

**CO<sub>2</sub> capture in industries and distributed energy  
systems:  
Possibilities and limitations**

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# **CO<sub>2</sub> capture in industries and distributed energy systems: Possibilities and limitations**

CO<sub>2</sub>-afvang in industrie en gedecentraliseerde energieconversiesystemen:  
Mogelijkheden en beperkingen  
(met een samenvatting in het Nederlands)

産業部門および分散型エネルギーシステムからの二酸化炭素回収  
可能性と限界について  
(日本語要旨付)

## **Proefschrift**

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## Abbreviations and units

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ASU	Air Separation Unit
ATR	Autothermal Reforming
AZEP	Advanced Zero Emission Power
BAT	Best Available Technique
BF	Blast Furnace
BFG	Blast Furnace Gas
BOF	Basic Oxygen Furnace
CCP	CO <sub>2</sub> Capture Project
CCS	Carbon Capture and Storage
CEPCI	Chemical Engineering Plant Cost Index
CHP	Combined Heat and Power plant
COE	Cost of Electricity
CO <sub>2</sub> -eq.	CO <sub>2</sub> equivalent
CSPC	Cell Stack Production Cost
d	day
DG	Distributed Generation
DH	District Heating
DOE	U.S. Department of Energy
DRI	Direct Reduced Iron
EPCCI	European Power Capital Cost Index
EU	European Union
FCC	Fluid Catalytic Cracker
GHG	Greenhouse Gas
GJ	Gigajoule
GT	Gas Turbine
Gt	Gigatonne
HEX	Heat Exchanger
HHV	Higher Heating Value
HPR	Heat-to-Power Ratio (for CHP plants)
HRSG	Heat Recovery Steam Generator
IEA	International Energy Agency
IEA GHG	International Energy Agency Greenhouse Gas R&D Programme
IGCC	Integrated Gasification Combined Cycle
IPCC	Intergovernmental Panel on Climate Change
KS-1	Chemical solvent developed by Mitsubishi Heavy Industries
kW	kilowatt
kWh	kilowatt-hour
LCA	Life Cycle Assessment

LHV	Lower Heating Value
LPG	Liquid Propane Gas
LT	Long Term future
MCFC	Molten Carbonate Fuel Cell
MCM	Mixed Conducting Membrane
MEA	Monoethanolamine
MGA	Membrane Gas Absorption
Mt	Megatonne
MT	Mid Term future
MW	Megawatt
MWh	Megawatt-hour
NECD	National Emissions Ceilings Directive
NG	Natural Gas
NGCC	Natural Gas Combined Cycle
NMVO	Non-Methane Volatile Organic Compounds
O&M	Operation and Maintenance
OCM	Oxygen Conducting Membrane
OECD	Organisation for Economic Co-operation and Development
OMA	Order of Magnitude Analysis
PC	Pulverized Coal
PM	Particulate Matter
ppm(v)	Parts Per Million (by volume)
PSA	Pressure Swing Adsorption
SEWGS	Sorption Enhanced Water Gas Shift
SF	Scaling Factor
SOEC	Solid Oxide Electrolyzer Cell
SOFC	Solid Oxide Fuel Cell
ST	Short Term
ST/MT	Short-Mid Term future
t	metric ton (tonne)
TCR	Total Capital Requirement
TGRBF	Top Gas Recycling Blast Furnace
TPC	Total Plant Cost
UNFCCC	United Nations Framework Convention on Climate Change
UNIDO	United Nations Industrial Development Organization
PSA/VSA	Pressure/Vacuum Pressure Swing Adsorption
WADE	World Alliance for Decentralised Energy
WGS	Water-Gas Shift
WGSMR	Water-Gas Shift Membrane Reactor
yr	year



## Introduction

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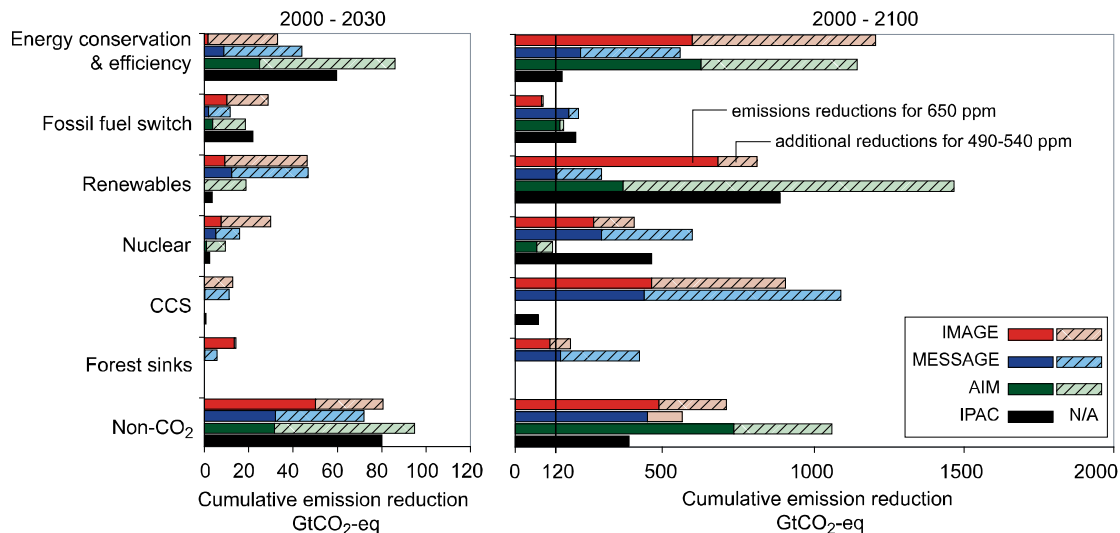
### 1.1 Background

#### 1.1.1 Climate change, greenhouse gas (GHG) emissions and carbon dioxide capture and storage (CCS)

Global warming is unequivocal from the observations of changes in atmosphere, cryosphere and oceans (2007b). Many natural systems are being affected by regional climate changes, in particular temperature increases (IPCC, 2007a). The Fourth Assessment Report (AR4) on Climate Change from the Intergovernmental Panel on Climate Change (IPCC, 2007a) reported that most of the global average warming observed since the mid-20<sup>th</sup> century is very likely due to anthropogenic greenhouse gases (GHG) increases. Among anthropogenic greenhouse gases, carbon dioxide (CO<sub>2</sub>) is the most important. The primary source of the increased atmospheric concentration of CO<sub>2</sub> since the pre-industrial period is fossil fuel use (IPCC, 2007b). For increases in global average temperature exceeding 1.5-2.5 °C and in concomitant atmospheric CO<sub>2</sub> concentrations, major changes in ecosystem structure and function, species' ecological interactions and shifts in species' geographical ranges with predominantly negative consequences for biodiversity and ecosystem goods and services such as water and food supply are projected to occur (IPCC, 2007a).

To reduce climate change risks, impacts and damages, to date over hundred countries have adopted a global warming target limiting the average temperature increase to 2 °C or below compared to pre-industrial levels as a guiding principle for mitigation efforts (Meinshausen et al., 2009). To reach this target, it is indicated that atmospheric GHG concentration needs to be stabilized at a level of 350-400 ppm(v) CO<sub>2</sub> or 445-490 ppm(v) CO<sub>2</sub>-eq. (IPCC, 2007c) and thus, global CO<sub>2</sub> emissions need to be reduced by 50 to 80% and maybe more, by 2050 compared to the 2000 level (IEA, 2008a).

In order to achieve such a stabilization target, a portfolio of mitigation measures will be needed (IPCC, 2005). Conventional measures to reduce CO<sub>2</sub> emissions such as fuel switching (e.g., from coal to natural gas), energy efficiency improvement, energy savings, renewables, and nuclear energy, may not be sufficient. CO<sub>2</sub> capture and storage (CCS) is considered an important option to reduce GHG emissions given the continued large-scale use of fossil fuels over the coming decades (e.g., IEA (2008a) and Stern (2006)). CCS comprises the separation of CO<sub>2</sub> from industrial and energy-related sources, transport of CO<sub>2</sub> to a storage location and long-term isolation from the atmosphere (IPCC, 2005). CCS may play a substantial role in the smooth and cost-effective transition to a sustainable, low-carbon energy future (IPCC, 2005; Rubin et al., 2007; Turkenburg, 1997). As seen in Figure 1-1, if a stabilization level of 490-540 ppm(v) CO<sub>2</sub>-eq. is to be achieved, some studies estimate the contribution of CCS to CO<sub>2</sub> emissions reduction as large as that of renewables, energy conversion and efficiency, and nuclear energy.

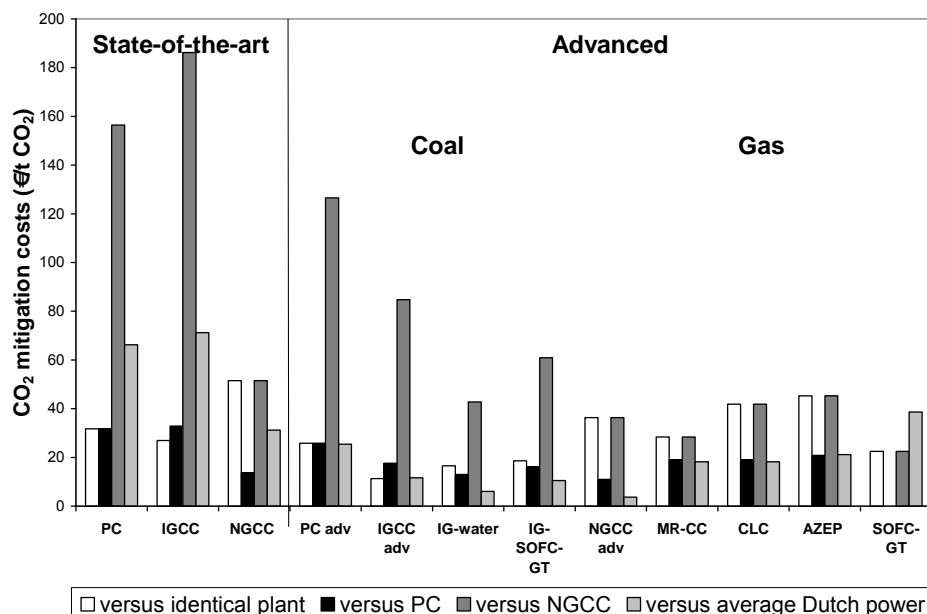


**Figure 1-1:** Cumulative emissions reductions for alternative mitigation measures for 2000 to 2030 (left-hand figure) and for 2000-2100 (right-hand figure). The figures show illustrative scenarios from four global models (AIM, IMAGE, IPAC and MESSAGE) for stabilization levels of 490-540 ppmv CO<sub>2</sub>-eq and 650 ppmv CO<sub>2</sub>-eq, respectively. Source: IPCC (2007c).

### 1.1.2 General description of CCS

CCS is comprised of three major components: capture, transport, and geological storage. A detailed description of each component can be found in, e.g., IEA (2008a), IEA (2005) and elsewhere. CO<sub>2</sub> capture systems are often categorized into four types: post-combustion capture, pre-combustion capture, oxyfuel combustion capture, and capture from industrial processes, and they are described in detail in, e.g. IEA (2005) and elsewhere. In post-combustion capture, CO<sub>2</sub> is captured from flue gas produced by combustion processes. In pre-combustion capture, fuel is converted into synthesis gas mainly comprised of CO and H<sub>2</sub> (so-called ‘syngas’) by reacting the fuel with oxygen or air and/or steam (IPCC, 2005). CO is reacted with steam (shift reaction) in a catalytic reactor to generate more H<sub>2</sub> and CO<sub>2</sub>. The CO<sub>2</sub> is separated from the H<sub>2</sub>-rich gas which can be used in various applications. In the oxyfuel concept, nearly pure oxygen is used for the combustion instead of air. The flue gas contains mainly CO<sub>2</sub> and H<sub>2</sub>O, which can easily be separated. With regard to CO<sub>2</sub> capture from industrial processes, a number of processes, e.g. purification of natural gas, ammonia production, alcohols and synthetic liquid fuels, already remove CO<sub>2</sub> as a necessary step of industrial production or a step to facilitate more cost-effective production. In general, the first three systems refer to power plants, but they can also be applied to industrial processes.

The captured CO<sub>2</sub> is compressed for transport via pipeline, ship, train or truck. The mode of transport depends on the transport volume, distance and location of the storage site. Pipeline and ship are suited for large-scale transport. The final step in the CCS chain is the



**Figure 1-2:** CO<sub>2</sub> mitigation costs of electricity production with CCS via vis-à-vis different reference systems, including indirect emissions of fuel extraction and transport (Damen et al., 2007). PC: pulverized coal-fired plant, IGCC: integrated gasification combined cycle, NGCC: natural gas combined cycle, SOFC-GT: solid oxide fuel cell with gas turbine MR: membrane reformer, CLC: chemical looping combustion, AZEP: advanced zero emission power plant. Note that the cost figures are expressed in €<sub>2003</sub> and do not take into account the energy and material price increase observed in recent years.

storage of CO<sub>2</sub> in geological formations at depths lower than 800 m. Nearly depleted hydrocarbon reservoirs, coal seams and saline formations, i.e., deep underground porous reservoir rocks saturated with brackish water or brine, can be used for storing CO<sub>2</sub> (IPCC, 2005). CO<sub>2</sub> is injected into the underground storage space in the pore of sedimentary rocks in a supercritical state (IPCC, 2005). Injection of CO<sub>2</sub> can increase the recovery of hydrocarbons when it is injected in oil and gas reservoirs (enhanced oil recovery (EOR) and enhanced gas recovery (EGR)) or coal beds that are too deep to mine (enhanced coal bed methane (ECBM) recovery). CO<sub>2</sub> can remain trapped underground by a number of mechanisms, e.g. trapping below caprock<sup>1</sup>; retention as an immobile phase trapped in the pore spaces of the storage formation; dissolution in the *in situ* formation fluids; adsorption onto organic matter in coal and shale; and/or formation of carbonate minerals by reacting with the minerals in the storage formation and caprock (IPCC, 2005).

Regarding the economic performance of CCS, CO<sub>2</sub> avoidance costs are currently estimated to be around 40-50 €<sub>2008</sub>/tCO<sub>2</sub><sup>2</sup> for coal-fired power plants and 40-75 €<sub>2008</sub>/tCO<sub>2</sub>

<sup>1</sup> Caprock is an impermeable, confining layer of very-low-permeability shale or salt (IPCC, 2005)

<sup>2</sup> Original values expressed in 2008 US dollars are converted to 2008 Euros using an exchange rate of 1\$=0.683 € (OANDA, 2011).

for gas-fired power plants (IEA, 2008a). The costs are expected to reduce considerably over time (Figure 1-2) due to developments in power plant efficiency, advanced CO<sub>2</sub> capture technologies and better system integration. The cost of CO<sub>2</sub> capture is the largest component of overall CCS costs for most large emission sources (IPCC, 2005) and thus, significant amount of research has been performed to reduce them. All these cost estimates, however, assume that CCS is deployed at full-scale power plants. To date CO<sub>2</sub> capture from power plants has only been demonstrated at small scale and it is indicated that the demonstration of CO<sub>2</sub> capture from full-scale power plants in the next ten years will be crucial to the wide deployment of CCS between 2020 and 2050 (IEA, 2007a). As of April 2011, there were 78 active and planned R&D projects on CO<sub>2</sub> capture around the world (GCCSI, 2011) and many of these projects are focusing on the application of CCS to large-scale power plants.

### 1.1.3 CCS in the industry

Although the power sector has been the main focus of RDD&D for CCS, the potential for CCS in the industry and petroleum refineries is expected to be very significant. Industry and petroleum refineries are among the largest contributors to anthropogenic CO<sub>2</sub> emissions. In 2006 direct and indirect CO<sub>2</sub> emissions from these sectors accounted for nearly 40% of total global emissions (IEA, 2008d, 2009; IEA GHG, 2006). CCS has a large potential not only because of their total CO<sub>2</sub> emissions but also because there are many industrial processes that generate gas streams rich in CO<sub>2</sub>, or in some cases pure CO<sub>2</sub>, which could increase the cost-effectiveness of the option by reducing the amount of effort needed to separate CO<sub>2</sub> from flue gases. Table 1-1 shows the properties of gas streams that can possibly be inputted to a CO<sub>2</sub> capture process (IPCC, 2005).

A study made by the International Energy Agency (IEA) estimated that in a scenario (BLUE Map scenario) aiming to half global GHG emissions in 2050 compared to 2007 level (comparable to a stabilization target of 2.0-2.4 °C), half of all CCS deployed (up to more than 10Gt per year) would be in industrial processes and the fuel transformation sector, see Figure 1-3(a) (IEA, 2008a). Moreover, the same IEA study also projects that considerable CCS deployment for power generation would be in industrial Natural Gas Combined Cycle (NGCC) Cogeneration of Heat and Power (CHP) plants (Figure 1-3(b)) (IEA, 2008a).

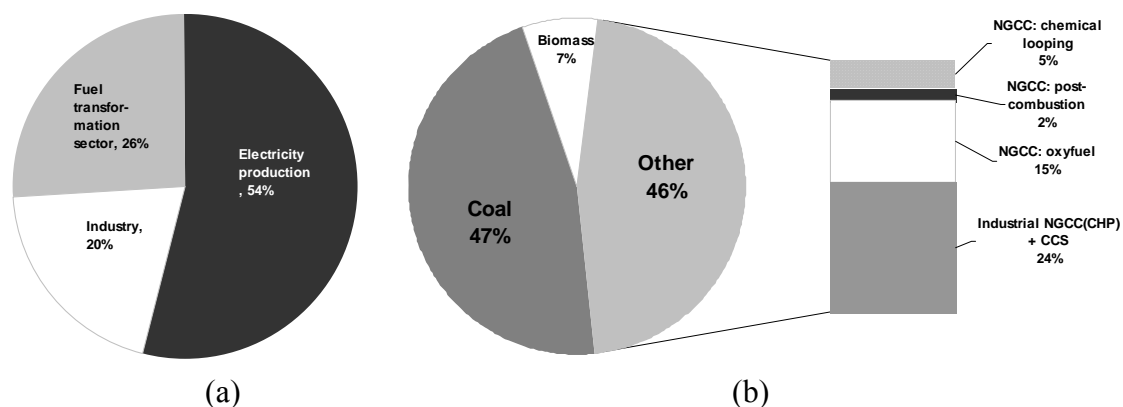
In recent years, some CO<sub>2</sub>-intensive industries have started up R&D programmes for technologies to achieve significant reduction in CO<sub>2</sub> emissions, including CCS. For example, the CO<sub>2</sub> Capture Project (CCP), formed by eight of the world's largest energy companies, has carried out extensive case studies for CO<sub>2</sub> capture from existing industrial plants around the world for both conventional and advanced CO<sub>2</sub> capture technologies (CCP, 2005). Other programs include the European Ultra Low CO<sub>2</sub> Steelmaking (ULCOS) program (ULCOS, 2011) and the CCS research project of the European Cement Research Academy (ECRA, 2009b). In addition, the United Nations Industrial Development Organization (UNIDO) has launched the "Carbon Capture and Storage-

Industrial Sector Roadmap” project aiming to develop a technology roadmap for CCS in various industrial sectors, particularly in developing countries (UNIDO, 2010a).

**Table 1-1:** Properties of gas streams that can be inputted to a CO<sub>2</sub> capture process. Source: IPCC (2005) with additional data from various sources (Damen et al., 2009; Duc et al., 2007; IEA GHG, 1999b; IPCC, 2009a).

<i>Source</i>	<i>CO<sub>2</sub> concentration [%vol (dry)]</i>	<i>Pressure of gas stream [bar]</i>	<i>CO<sub>2</sub> partial pressure [bar]</i>
<b>CO<sub>2</sub> from fuel combustion</b>			
Power station flue gas			
Natural gas fired boilers	7-10	1	0.07-0.1
Gas turbines	3-4	1	0.03-0.04
Oil fired boilers	11-13	1	0.11-0.13
Coal fired boilers	12-14	1	0.12-0.14
IGCC: after combustion	12-14	1	0.12-0.14
Oil refinery and petrochemical plant fired heaters	8-13	1	0.08-0.13
<b>CO<sub>2</sub> from chemical transformations + fuel combustion</b>			
Blast furnace gas			
Before combustion	16-26	2-3	0.3-0.8
After combustion	24-27	1	0.24-0.27
Cement kiln off-gas	14-33	1	0.14-0.33
<b>CO<sub>2</sub> from chemical transformations before combustion</b>			
IGCC: synthesis gas after gasification	8-20	20-70	1.6-14
<b>Chemical reaction</b>			
Ammonia production	~100	1	1
Ethylene oxide	~100	1	1
Hydrogen production	15-20	1	1
Methanol production	~100	1	1
<b>Other processes</b>			
Natural gas processing	~100	1	1

Because of the increasing interest in the deployment of CCS in the industry, a relatively large number of techno-economic feasibility studies on CO<sub>2</sub> capture from industrial processes have been published in recent years (e.g. CCP, 2005; Rootzén et al., 2009). However, it is currently difficult to determine which technologies are truly promising and what the possible costs for CO<sub>2</sub> capture are for different timeframes. There are a number of reasons that makes the comparison and assessment of the reported CO<sub>2</sub> capture performance difficult. Firstly, the technologies required in the industry are more diverse than in the power sector (UNIDO, 2010b). While some CO<sub>2</sub> capture technologies are of an add-on nature, others are integrated with the industrial process potentially affecting the material and energy flows of the entire industrial plant considerably. To enable a fairer comparison of CO<sub>2</sub> capture performance, not only the CO<sub>2</sub> capture process but also the entire industrial plant needs to be taken into account within consistent system boundaries. Secondly, each literature source contains specific assumptions that affect the economic performance calculations. For economic performance assessment, assumptions on fuel



**Figure 1-3:** Breakdown of CCS deployment by sector (a) and breakdown of electricity generation from power plants fitted with CCS by technology and fuel (b), in a scenario to reduce global greenhouse gas (GHG) emissions to half in 2050 compared to 2007 level (IEA, 2008a).

price, capital cost estimation, interest rate, and economic lifetime, have a large impact on the results. Comparative assessment of CO<sub>2</sub> capture technologies proposed in the literature on the basis of consistent system boundaries and standardized parameters have been performed for energy conversion systems by, e.g., Damen et al. (2006). Such an analysis, however, has not yet been performed for the wide range of studies on CO<sub>2</sub> capture from industrial processes available in the literature.

Besides being a large CO<sub>2</sub> emitter, industry has been one of the major sources of environmental pollution. Large efforts have been put in the last decades to minimize the environmental impacts. Nevertheless, problems such as air pollution due to the emissions of, e.g., NO<sub>x</sub>, SO<sub>2</sub> and particulate matter (PM), still remain. The evaluation of new technologies such as CCS should, therefore, take into account the potential environmental effects other than CO<sub>2</sub> emissions reduction. It is very likely that CCS would somehow affect these air pollutant emissions because they are usually contained in gas streams from which CO<sub>2</sub> is captured. There are a number of studies published on post-combustion CO<sub>2</sub> capture from power plants that have looked at the changes in the emissions of air pollutants such as NO<sub>x</sub>, SO<sub>2</sub>, PM and NH<sub>3</sub> (an overview of literature can be found in Koornneef et al. (2010)). To our knowledge, such assessments have not been performed systematically for CO<sub>2</sub> capture in industrial processes.

#### 1.1.4 CCS in distributed energy systems

As discussed, Figure 1-3 illustrates a significant potential of CCS for industrial CHP plants. CHP is one of the major applications of distributed generation (DG), which is defined as “an electric power source connected directly to the distribution network or on the customer site of the meter” (Ackermann et al., 2001). The scale of the DG stations is

irrelevant for this definition, but they are generally smaller than centralized power plants because the plant scales are determined by the heat demand of the user<sup>3</sup>. DG is expected to become increasingly important in the future energy supply infrastructure, particularly in future electric utilities in the economies where deregulation takes place (Ogden, 2002). Moreover, DG may make a significant contribution to the mitigation of CO<sub>2</sub> emissions. CHP is a popular DG option because it is more energy efficient than the separate generation of electricity and heat. Its deployment, therefore, leads to reductions in primary energy consumption and CO<sub>2</sub> emissions. The IEA estimates that in the BLUE Map scenario, 30% of global electricity generation from fossil fuels would come from gas-fired CHP plants, with more than a quarter of the gas-fired CHP power generation decarbonized with CCS (IEA, 2010a).

The increase of DG capacity, however, means an increase in CO<sub>2</sub> emissions from smaller scale CO<sub>2</sub> emission sources. Distributed energy systems today already account for a substantial proportion of total global CO<sub>2</sub> emissions. The IEA Greenhouse Gas R&D Programme (Hendriks et al., 2009; IEA GHG, 2007a) estimated the CO<sub>2</sub> emissions from “medium scale”<sup>4</sup> installations to be 10-15% of total global CO<sub>2</sub> emissions, with the largest shares of emissions from boilers and furnaces, particularly those that are fired by coal. It is also estimated that in 2030, CO<sub>2</sub> emissions from distributed energy systems may account for about 20% for current global energy-related CO<sub>2</sub> emissions (see Chapter 6). If stringent CO<sub>2</sub> emissions reduction is needed, also distributed energy systems have to consider options that enable large CO<sub>2</sub> emissions reduction, such as CCS.

CCS in distributed energy systems are generally considered to be more expensive than in centralized power plants due the economies of scale (Damen et al., 2007; IPCC, 2005). However, DG plants operate on different conditions than centralized power plants. For example, a CHP plant operates according to the heat demand of an individual consumer while a centralized power plant to the total electricity demand of a city of a region. Moreover, the heat integration of CHP plants is configured differently than the centralized power plants. Therefore, typical CO<sub>2</sub> capture energy penalties and costs for centralized power plants many not apply to DG systems. It is important to identify and quantify these differences. Meanwhile, a number of studies indicate that advanced technologies may enable low-cost CO<sub>2</sub> capture in the long term (e.g., Damen et al., 2006; Hendriks et al., 2009). For example, Damen et al. (2006) shows that CO<sub>2</sub> capture costs for small-scale energy conversion systems (2-20 MW<sub>e</sub>) may become competitive to those for large-scale systems (600 MW<sub>e</sub>) in the long term. Therefore, it is necessary to identify distributed energy systems that are worth considering CCS already for the short-term and those only in the long-term. To date, however, no study assessed and compared various CO<sub>2</sub> capture

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<sup>3</sup> There is also a limit to the maximum capacity for DG stations (100-150 MW<sub>e</sub>) due to technical constraints regarding the voltage level within the distribution system (Ackermann et al., 2001).

<sup>4</sup> “Medium scale” is defined as installations with a fuel input of between 1 and 100 MW<sub>th</sub>. A 100 MW<sub>th</sub> coal fired plant operating for 7,500 h/yr would produce about 250 ktCO<sub>2</sub>/yr of CO<sub>2</sub> assuming an emission factor of 94 kg CO<sub>2</sub> per GJ, and a natural gas fired plant would produce about 150 ktCO<sub>2</sub>/yr assuming an emission factor of 56 kg CO<sub>2</sub> per GJ.

technologies for distributed energy systems explicitly taking into account differences in timeframe, fuel type, energy system type, plant scale and operational conditions.

## 1.2 Objectives and outline of the thesis

In this context, the objective of this thesis is to assess technical and economic performance for CO<sub>2</sub> capture from industrial plants and distributed energy systems. The research questions are as follows:

RQ.1: What are technical and economic possibilities for CO<sub>2</sub> capture in key industrial sectors?

RQ.2: What are possible impacts of CO<sub>2</sub> capture technologies on the emission of air pollutants (NO<sub>x</sub>, SO<sub>2</sub>, PM and NH<sub>3</sub>) from key industrial processes?

RQ.3: What are technical and economic possibilities for CO<sub>2</sub> capture in distributed energy systems?

Table 1-2 shows an overview of the research questions addressed in each chapter of this thesis. The structure of this thesis is as follows. Chapters 2 and 3 address RQs 1 and 2, respectively. RQ.3 is principally addressed in Chapter 6, with Chapters 4 and 5 addressing specific issues regarding CO<sub>2</sub> capture from distributed energy systems.

**Table 1-2:** Overview of research questions addressed in each chapter.

	<i>Chapter 2</i>	<i>Chapter 3</i>	<i>Chapter 4</i>	<i>Chapter 5</i>	<i>Chapter 6</i>
RQ.1	X				
RQ.2		X			
RQ.3			x	x	X

Chapter 2 performs a techno-economic assessment and comparison of CO<sub>2</sub> capture technologies for key industrial sectors (iron and steel, cement, refining and petrochemicals). The assessment is based on an extensive literature review and interviews with experts, covering both industrial reports and peer-reviewed scientific articles. Key parameters, e.g., capacity factor, energy prices, interest rate, economic plant lifetime, CO<sub>2</sub> compression pressure, and grid electricity CO<sub>2</sub> intensity, were standardized to enable a fair comparison of technologies. The analysis focuses on the changes in energy, CO<sub>2</sub> emissions and material flows, due to the deployment of CO<sub>2</sub> capture technologies. CO<sub>2</sub> capture technologies are categorized into short-mid term (ST/MT) and long-term (LT) technologies.

Chapter 3 quantitatively assesses the changes in the emissions of NECD substances (NO<sub>x</sub>, SO<sub>2</sub>, PM and NH<sub>3</sub>) from industrial processes due to CO<sub>2</sub> capture technologies in the



European Union in the short-mid term (timeframe: 2020-2025). This chapter investigates the following industrial sectors: cement, iron and steel (integrated steel plants), and petroleum refinery furnaces and boilers (combined stacks). The analysis includes onsite emissions as well as the emissions associated with the changes in grid electricity consumption. The changes in emissions are assessed at plant level as well as at EU-27 sector level.

Chapter 4 investigates the techno-economic performance of post-combustion CO<sub>2</sub> capture from 50-200 MW<sub>e</sub> (maximum electrical capacity) scale industrial NGCC-CHP plants in comparison with large-scale (400 MW<sub>e</sub>) condensing NGCC plants, in the short term (2010) and the mid-term future (2020-2025). The potential reduction of CO<sub>2</sub> capture energy penalty and its effect on CO<sub>2</sub> capture cost is analyzed when the unused capacity of the partial-load operating industrial CHP plants is utilized to supply energy for CO<sub>2</sub> capture. Also the effect of scale on the CO<sub>2</sub> capture performance is investigated.

Chapter 5 examines whether potentially low-cost CO<sub>2</sub> capture from solid oxide fuel cell (SOFC) systems could enhance the penetration of SOFC in the energy market in a highly carbon-constrained society in the mid-term future (up to year 2025). The technical performance of a 5 MW<sub>e</sub> SOFC system for industrial combined heat and power generation (CHP) with CO<sub>2</sub> capture using existing commercial technology was calculated using a process modeling software. For CO<sub>2</sub> capture, oxyfuel combustion of anode off-gas using commercially available technologies was selected. The competitiveness of the SOFC-CHP system was compared with conventional gas turbine (GT-) CHP technology.

Chapter 6 provides an overview of CO<sub>2</sub> capture opportunities for DG systems and smaller scale energy systems and their economic performance both in the short-term and the long term. This chapter also investigates the short-term technical and economic performance of CO<sub>2</sub> capture from district heating (DH-) CHP plants. The potential for heat integration specific for the application to DH-CHP plants and the effect of plant scale was carefully examined.

In Chapter 7, the objectives and the key findings of this thesis are summarized, conclusions are drawn, and recommendations for further research are presented.

Finally, some important parameters are the fossil fuel prices, the economic lifetime of the investments, and the interest rate when performing economic analyses. In this thesis, natural gas price was assumed 7.5 €/GJ<sub>LHV</sub> for the short-term (up to 2020) and 8 €/GJ<sub>LHV</sub> for the mid-term (2020-2030) and long-term future (beyond 2030). Coal price is assumed 2.6 €/GJ<sub>LHV</sub> for all timeframes. The interest rate is assumed 10-12% for the industry and 10% for other sectors. The economic lifetime of plants are assumed 35 years for coal-fired energy systems and 20-25 years for other installations.



## Comparative assessment of CO<sub>2</sub> capture technologies for carbon-intensive industrial processes

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### Abstract

This article presents a consistent techno-economic assessment and comparison of CO<sub>2</sub> capture technologies for key industrial sectors (iron and steel, cement, petroleum refineries and petrochemicals). The assessment is based on an extensive literature review, covering studies from both industry and academia. Key parameters, e.g. capacity factor (91-97%), energy prices (natural gas: 8 €<sub>2007</sub>/GJ, coal: 2.5 €<sub>2007</sub>/GJ, grid electricity: 55 €<sub>2007</sub>/MWh), interest rate (10%), economic plant lifetime (20 years), CO<sub>2</sub> compression pressure (110 bar), and grid electricity CO<sub>2</sub> intensity (400 g/kWh), were standardized to enable a fair comparison of technologies. The analysis focuses on the changes in energy, CO<sub>2</sub> emissions and material flows, due to the deployment of CO<sub>2</sub> capture technologies. CO<sub>2</sub> capture technologies are categorized into short-mid term (ST/MT) and long term (LT) technologies. The findings of this study identified a large number of technologies under development, but it is too soon to identify which technologies would become dominant in the future. Moreover, a good integration of industrial plants and power plants is essential for cost-effective CO<sub>2</sub> capture because CO<sub>2</sub> capture may increase the industrial onsite electricity production significantly.

For the iron and steel sector, 40-65 €/tCO<sub>2</sub> avoided may be achieved in the ST/MT, depending on the iron making process and the CO<sub>2</sub> capture technique. Advanced LT CO<sub>2</sub> capture technologies for the blast furnace-based process may not offer significant advantages over conventional ones (30-55 €/tCO<sub>2</sub> avoided). Rather than the performance of CO<sub>2</sub> capture technique itself, low-cost CO<sub>2</sub> emissions reduction comes from good integration of CO<sub>2</sub> capture to the ironmaking process. Advanced smelting reduction with integrated CO<sub>2</sub> capture may enable lower steel production cost and lower CO<sub>2</sub> emissions than the blast furnace-based process, i.e., negative CO<sub>2</sub> mitigation cost. For the cement sector, post-combustion capture appears to be the only commercial technology in the ST/MT and the costs are above 65 €/tCO<sub>2</sub> avoided. In the LT, a number of technologies may enable 25-55 €/tCO<sub>2</sub> avoided. The findings also indicate that, in some cases, partial CO<sub>2</sub> capture may have comparative advantages. For the refining and petrochemical sectors, oxyfuel capture was found to be more economical than others at 50-60 €/tCO<sub>2</sub> avoided in ST/MT and about 30 €/tCO<sub>2</sub> avoided in the LT. However, oxyfuel retrofit of furnaces and heaters may be more complicated than that of boilers.

Crude estimates of technical potentials for global CO<sub>2</sub> emissions reduction for 2030 were made for the industrial processes investigated with the ST/MT technologies. They amount up to about 4 Gt/yr: 1 Gt/yr for the iron and steel sector, about 2 Gt/yr for the cement sector, and 1 Gt/yr for petroleum refineries. The actual deployment level would be much lower due to various constraints, about 0.8 Gt/yr, in a stringent emissions reduction scenario.

## 2.1 Introduction

Industry and petroleum refineries are among the largest contributors to anthropogenic CO<sub>2</sub> emissions. In 2006 these two sectors together emitted more than 11GtCO<sub>2</sub><sup>5</sup> directly and indirectly, accounting for nearly 40% of total global CO<sub>2</sub> emissions (IEA, 2009, 2010b; IEA GHG, 2006). Alongside energy efficiency improvement, renewables and nuclear energy, CO<sub>2</sub> capture and storage (CCS) is considered a promising option to achieve significant reduction in CO<sub>2</sub> emissions. CCS has a large potential in industry and petroleum refineries not only because of its large CO<sub>2</sub> emissions but also because there are many industrial processes that generate gas streams rich in CO<sub>2</sub>, or in some cases pure CO<sub>2</sub>, which could reduce the costs of CCS. A recent study by the International Energy Agency (IEA) estimates that in a scenario to reduce global greenhouse gas (GHG) emissions to half in 2050 compared to today's level, about half of all CCS deployed (up to more than 10Gt per year) would be in industrial processes (cement, iron and steel and chemicals) and fuel transformation sector (petroleum refineries and liquefied natural gas production) (IEA, 2008a). The United Nations Industrial Development Organization (UNIDO) recently carried out a project to develop a roadmap for CCS in various industrial sectors (UNIDO, 2010a).

### 2.1.1 Overview of key literatures on CCS in the industry

One of the first comprehensive studies on the techno-economic performance of CO<sub>2</sub> capture from carbon-intensive industrial processes was performed in 1995 by Farla et al. (1995) for the iron and steel industry, and the petrochemical sector. The study was focused on applying chemical absorption to capture CO<sub>2</sub>. It concluded that, in comparison with CO<sub>2</sub> capture from the flue gases of a thermal power plant, cost figures are comparable for the iron and steel sector and higher for the petrochemical sector. The IEA Greenhouse Gas R&D Programme (IEA GHG) also delivered a set of reports on the techno-economic performance of CO<sub>2</sub> capture from cement plants and oil refineries in the late 1990s (IEA GHG, 1999c, 2000a). These reports investigate the performance of CO<sub>2</sub> capture technologies other than the chemical absorption method. For the cement sector, the results suggest that the kiln operation in a CO<sub>2</sub>/O<sub>2</sub> atmosphere may be a promising technique to

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<sup>5</sup> Industry accounted for 10.6 Gt of direct and indirect emissions, and petroleum refineries accounted for 0.8 Gt of direct emissions.

recover CO<sub>2</sub> and that chemical absorption method seems less appropriate because of the high heat requirement whereas the heat is not easily available from the cement production process. For refinery heaters, the economic performances of amine-based capture from the flue gas and oxyfuel combustion capture are very similar.

Since the late '90s a number of studies have been published on a broad range of industrial sectors with various CO<sub>2</sub> capture technologies that are potentially more cost-effective than the chemical absorption method. The CO<sub>2</sub> Capture Project (CCP), executed by eight of the world's largest energy companies, carried out extensive case studies for CO<sub>2</sub> capture from an existing refinery in the UK (CCP, 2005) using both conventional and advanced CO<sub>2</sub> capture technologies. For furnaces and heaters, the results suggest that oxyfuel combustion capture can achieve 48% reduction in CO<sub>2</sub> avoidance cost compared to post-combustion capture using amine solvent (CCP, 2005). Also other industrial associations and consortia have been carrying out extensive research and development (R&D) activities for cost-effective CO<sub>2</sub> capture. Examples are the European Ultra Low CO<sub>2</sub> Steelmaking (ULCOS) program (ULCOS, 2011) and the CCS research project of the European Cement Research Academy (ECRA, 2009a, b). In particular, the ULCOS program investigated the economic performance of a large number of low-CO<sub>2</sub> iron and steel making technologies, with and without CCS, under various scenarios. Their results suggest that chemical absorption capture is less economical than other commercially available CO<sub>2</sub> capture technologies (Birat and Lorrain, 2009). Most of these studies have been reviewed and compared in the UNIDO Industrial CCS roadmap project, from which five sectoral assessment reports (biomass, high-purity CO<sub>2</sub> sources, iron and steel, refineries, and cement) (Birat and Maizière-lès-Metz, 2010; Carbo, 2011; Det Norske Veritas, 2010; Mott MacDonald, 2010; Zakkour and Cook, 2010), a technical synthesis report (de Coninck and Mikunda, 2010), and a policy brief prepared for the 16<sup>th</sup> Conference of the Parties (COP16) of the United Nations Framework Convention on Climate Change (UNFCCC) held in Cancun, Mexico, in 2010 (UNIDO, 2010c).

Although considered less economical, feasibility of chemical absorption CO<sub>2</sub> capture has been continually investigated because it is the only available technology for most emission sources if CCS is to be deployed in the near future. An assessment of the techno-economic performance of post-combustion capture retrofit for iron and steel, cement, and oil refining plants in Australia has been performed recently by Ho et al. (2010). The study concludes that among three sectors, the iron and steel sector may be able to deploy CCS at an early stage due to the moderate cost of CO<sub>2</sub> capture and economies of scale. The study also concludes that for the other two sectors, technological improvements or financial incentives are required before CCS is applied.

### **2.1.2 Rationale and objectives**

Literature reviews on the CO<sub>2</sub> capture technologies for the industry and petroleum refineries include those from the IEA (2004, 2008a) and the Special Report on CO<sub>2</sub> capture and Storage (SRCCS) from the Intergovernmental Panel on Climate Change

(IPCC, 2005). These reports reviewed literature for the following sectors: natural gas sweetening, steel, cement, pulp and paper, and petroleum refineries. Table 2-1 presents an overview of some key literatures that investigated and/or reviewed the economic performance of CO<sub>2</sub> capture for a wide range of industrial processes. All these studies are informative, but they are incomplete in one way or another when CO<sub>2</sub> capture technologies are concerned. Firstly, some reviews (IEA, 2008a; IPCC, 2005) are more of an introductory nature. Although these reviews provide a good overview of the potentially feasible CO<sub>2</sub> capture options and as well as R&D activities in the industry on CO<sub>2</sub> capture, they do not go deep into the technical and economic details of CO<sub>2</sub> capture technologies partly because the aim of these reviews is to cover all aspects of CCS in a comprehensive manner. Secondly, partly due to the aim of these reports, they do not look into assumptions behind the calculated results in each publication. For economic performance assessment, however, assumptions on system boundaries, fuel price, capital costs, interest rate, and economic lifetime, can have a large impact on the outcomes. Without standardizing key parameters, a fair comparison of technical and economic performance of CO<sub>2</sub> capture technologies published in the literature is not possible. For power and hydrogen production, Damen et al. (2006; 2007) performed a comparative assessment of CCS technologies with a standardization of key parameters. Melien (2005) also compared various technologies researched in the CO<sub>2</sub> Capture Project (CCP) based on standardized economic parameters. Such an analysis, however, has not yet been performed yet for the wide range of studies on CO<sub>2</sub> capture from industrial processes available in the literature. Therefore, the objective of this paper is to perform a consistent assessment and comparison of the technical and economic performance of CO<sub>2</sub> capture technologies for industrial processes.

While large fraction of the literature covered in this study have been reviewed the series of reports from the UNIDO project (Birat and Maizière-lès-Metz, 2010; Carbo, 2011; de Coninck and Mikunda, 2010; Det Norske Veritas, 2010; Mott MacDonald, 2010; Zakkour and Cook, 2010), the main added value of this study is the assessment and comparison of the economic performance of CO<sub>2</sub> capture based on uniform system boundaries and standardized underlying parameters, i.e., capacity factor, energy prices, interest rate, economic plant lifetime, CO<sub>2</sub> compression pressure, and grid electricity CO<sub>2</sub> intensity. Moreover, this study will take an in-depth look into various aspects that may affect the techno-economic prospects for CO<sub>2</sub> capture from industry and petroleum refineries including possibilities for retrofit.

**Table 2-1:** Overview of key literatures that investigated or reviewed the economic performance of CO<sub>2</sub> capture for a wide range of industrial processes.

Reference	Cost year	Interest rate	Plant economic lifetime	CO <sub>2</sub> avoidance costs					
				Iron and steel	Cement	Chemical and petrochemical	Petroleum refineries	Pulp and paper	Other industrial processes
<b>Original studies</b>									
Farla et al. (1995): chemical absorption	US\$ <sub>1990</sub>	7%	25	35 \$/t	Not assessed.	Combined stacks: 46 \$/t Ammonia plants: 8 \$/t	Not investigated.	Not assessed.	Not assessed.
Melien ((2005), in CCP (2005))	US\$ <sub>2003</sub>	11% annuity factor		Not assessed.			41-85\$/t (various technologies)	Not assessed.	Oil and gas field and oil sand/synthetic crude facility
Ho et al. (2010)(all retrofits)	US\$ <sub>2008</sub>	7%	25	BF: 60 \$/t (WGS) - 68 \$/t (MEA) COREX : 32 \$/t (WGS) - 52 \$/t (MEA)	68 \$/t (MEA)	Not assessed.	Combined stacks. 87 \$/t (MEA)	Not assessed.	Not assessed.
McKinsey & Company (2009) <sup>1)</sup> (for 2030)	€ <sub>2005</sub>	4%	20-30	New plants: ~25 €/t Retrofit: ~30 €/t	New plant: ~25 €/t Retrofit: ~35 €/t	Combustion: ~30 €/t Ammonia: ~25 €/t	Upstream: ~50 €/t Downstream: 35-50 €/t	Not assessed.	Not assessed.
<b>Reviews and overview studies</b>									
IPCC (2005)	Varied (original figures from reviewed studies)			Current: cites only Farla et al. (1995)	No citation.	No citation.	Current: 74-116 \$/t (cites two studies (CCP, 2005; IEA GHG, 2000a)) Advanced: 41-76 \$/t (cites (Melien, 2005) only)	Current: 23-53 \$/t (cites only one study)	No citation.
IEA (2008a). All figures from “Roadmap”, unless otherwise stated.	US\$ <sub>2008</sub>	Country/region specific	Not stated.	BF: 40-60 \$/t (2030) Smelting reduction: 30-50 \$/t (2030) 20-40 \$/t (2050)	Post-combustion: 100 \$/t (2030), 75 \$/t (2050) Oxyfuel: 50 \$/t (2030), 40 \$/t (2050)	Large CHP plants: “similar to that of other power plants”. Ammonia: 20 \$/t	Heaters: 238 \$/t investment cost (from (CCP, 2005))	Black liquor boilers: 40 \$/t (2030), 35 \$/t (2050) Gasifiers : 30 \$/t (2030) , 25 \$/t (2050)	Natural gas sweetening, heavy oil and tar sands, H <sub>2</sub> production, gasification and hydrocarbon synfuel production
Rootzen et al. (2009)	Not stated.	Not stated.	Not stated.	~20 €/t captured (top gas recycling)	Oxyfuel combustion ~34 €/t captured	Not assessed.	Short-mid term: Furnaces and boilers ~ 30 €/t captured (oxyfuel combustion) Catalytic crackers ~45 €/t captured (MEA)	Not assessed.	Not assessed.

<sup>1)</sup> No specific CO<sub>2</sub> capture technology or CO<sub>2</sub> capture plant scale is indicated. The original cost figures include CO<sub>2</sub> transport and storage costs. The figures presented in the table do not include these two cost components.

The focus of this study is on the following sectors: iron and steel, cement, petroleum refineries and petrochemicals. The iron and steel, cement, and petrochemical sectors account for nearly three-fourths of global total industrial CO<sub>2</sub> emissions (IEA, 2009). Petroleum refineries account for an additional 0.8 Gt of CO<sub>2</sub> per year. Industrial processes that generate pure CO<sub>2</sub> streams, e.g., natural gas sweetening, ammonia production, and coal and oil gasification, are not investigated in this study as CO<sub>2</sub> separation is already practiced in an economical manner (currently the separated CO<sub>2</sub> is vented). The pulp and paper sector is also excluded from this study due to its relatively small energy consumption and CO<sub>2</sub> emissions compared to other industrial sectors. The sector accounts for 6% of total global industrial energy consumption with half of it coming from biomass (IEA, 2009), and is consequently accountable for only 2% of total global industrial CO<sub>2</sub> emissions (IEA, 2009). The techno-economic assessment of CO<sub>2</sub> capture technologies for chemical and petrochemical sector and for petroleum refineries are performed together because the main CO<sub>2</sub> emission sources in these sectors are similar: furnaces and boilers, and onsite (combined heat and) power plants. Note that CCS from biofuel production (synfuels and H<sub>2</sub>) is expected to play a significant role in the future; its potential is projected to be more than 25% of total industrial CCS potential in 2020 and more than 50% in 2050 (Carbo, 2011). The performance of biofuel production with CO<sub>2</sub> capture was not investigated in this study due to the scarcity of data (Carbo, 2011).

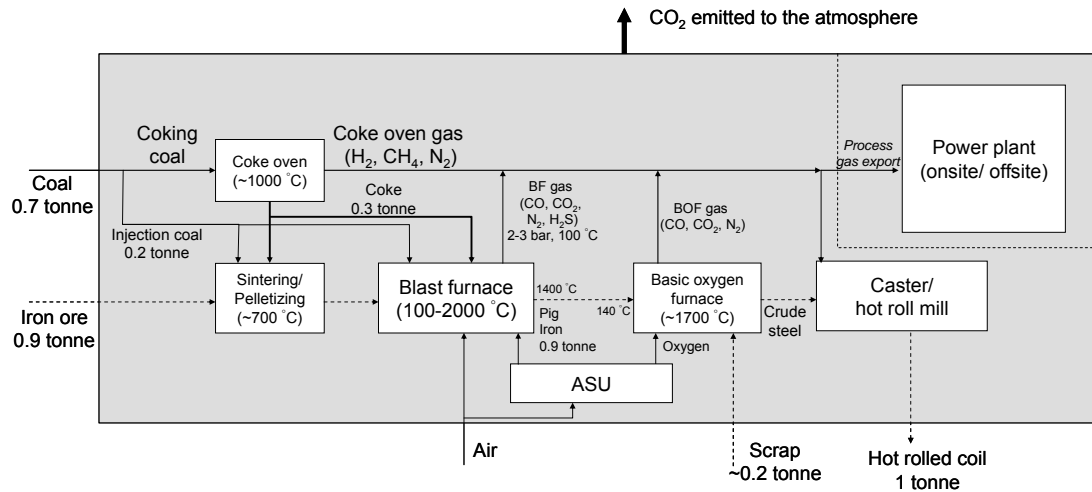
Finally, this study considers CO<sub>2</sub> capture plant installations in the industrialized world. With regard to timeframe, the implementation of CO<sub>2</sub> capture in the short-mid term future (ST/MT: 10-15 years) and in the long term future (LT: 20 years or more) is considered. ST/MT technologies are defined as those that are either in pilot plant, demonstration or commercialization phase today (Peeters et al., 2007). The technologies are categorized as short-mid term technologies also when all required components are commercially available today, even if the process as a whole has not been tested or demonstrated. All other technologies, either in modeling or laboratory phase today, are considered to be LT options. Note that the performance data used in this study assumes that the technologies are commercially mature. Cost estimates for first-of-a-kind plants are, therefore, excluded.

## **2.2 Description of the industrial sectors studied**

### **2.2.1 Iron and steel sector**

The iron and steel sector is one of the largest energy-consuming manufacturing sectors in the world. Final energy use in 2007 was 26 EJ worldwide, accounting for nearly 20% of total industrial energy consumption (IEA, 2007a). The iron and steel sector alone emitted 2.3GtCO<sub>2</sub> worldwide in 2007, accounting for 30% of total direct industrial CO<sub>2</sub> emissions and nearly 10% of global total CO<sub>2</sub> emissions (IEA, 2007a). The worldwide crude steel production was 1219 Mt in 2009 (World Steel Association, 2010) and may increase up to about 1600 Mt in 2030 (Oda et al., 2007).





**Figure 2-1:** Simplified flow diagram of an integrated iron and steel production process. Typical material flow rates are taken from Birat et al. (2009).

Around 60% of world total steel is made from pig iron produced in blast furnaces (BF) and most of the rest is made from steel scrap (IEA, 2007a). The integrated steelmaking process using BF will continue to play a dominant role in the industry in the longer term (Birat et al., 2003). Figure 2-1 shows a simplified flow diagram of an integrated iron and steel making process via pig iron. A general description of carbon flows in an integrated iron and steel production process can be found in, e.g., Farla et al. (1995), IEA (2007a), and elsewhere. An integrated iron and steel making process consists of mainly five sections: coking, iron ore agglomeration, blast furnace, basic oxygen furnace (BOF), and final product manufacturing, e.g. steel casting, rolling and finishing. Coke is produced in a coke oven by pyrolyzing coal or lignite. Volatile organic compounds (coke oven gas), tar, and sulfur compounds are removed in this process. Coke oven gas is used in the steel making process. Coke breeze (a residue from the screening of heat-treated coke (IUPAC, 1997)) is used in an iron ore agglomeration process, which aggregates fine ore so that it can be used in blast furnaces. Sintering and pelletizing are two common agglomeration processes. In a BF, pig iron and blast furnace gas (BFG) is produced by reducing iron ore sinters and pellets with coal and coke. BFG is partly used for other processes within the iron and steel making plant, and the rest can be exported for power generation. The pig iron produced in the blast furnace is converted into steel in BOF, where the large part of carbon contained in the pig iron is removed by blowing pure oxygen.

Other iron and steel production processes currently operating around the world include the smelting reduction process and the direct reduced iron (DRI) production process. The smelting reduction process is the latest development in pig iron production, which omits coke production by combining the gasification of non-coking coal with the reduction of iron ore in a liquid bath (Price et al., 2002). The smelting reduction reactor resembles the lower part of a blast furnace. The reduction process generates a large amount of residual gas which, in the most effective designs, is used for pre-reduction of the solid ore (IEA, 2009). The smelting reduction facilitates CO<sub>2</sub> capture because the flue gas has a higher

CO<sub>2</sub> concentration than conventional blast furnace gas as the furnace is blown with pure oxygen (some nitrogen needs to be injected in order to maintain momentum and heat transfer within the furnace). As of 2008, the COREX process is the only smelting reduction process commercially operating worldwide (IEA, 2008c). Advanced smelting reduction could enable the direct use of iron ores. One approach that recently became commercial in recent years is the FINEX process developed by POSCO and Siemens VAI (Siemens VAI, 2007). The FINEX process is currently being operated at a scale of 1.5Mt/yr hot metal (POSCO, 2008a). The FINEX process is very similar to the COREX process; the major difference is that the FINEX process enables the direct use of sinter feed fine ore (Siemens VAI, 2007). The scale of typical smelting reduction steel making plants is smaller than that of typical BF-based plants.

DRI is produced by reduction of iron ore in a solid form in small-scale plants (below 1Mt/yr) (Price et al., 2002). In contrast to pig iron, DRI contains all gangue (mixture of valueless minerals) elements of the iron ores because there is neither melting nor slag phase, requiring a separation process in the electric arc furnace (EAF) (IEA, 2009). The iron ore can be reduced by coal, coke, or natural gas. As of 2004, around 5% of world steel is produced from DRI, most of it is natural gas based (IEA, 2007a). Commercialized DRI technologies include the MIDREX process, which uses natural gas for iron ore reduction. CO<sub>2</sub> capture from the reduction gas is already widely applied in DRI production (Gielen, 2003) and is, therefore, is not assessed in this study.

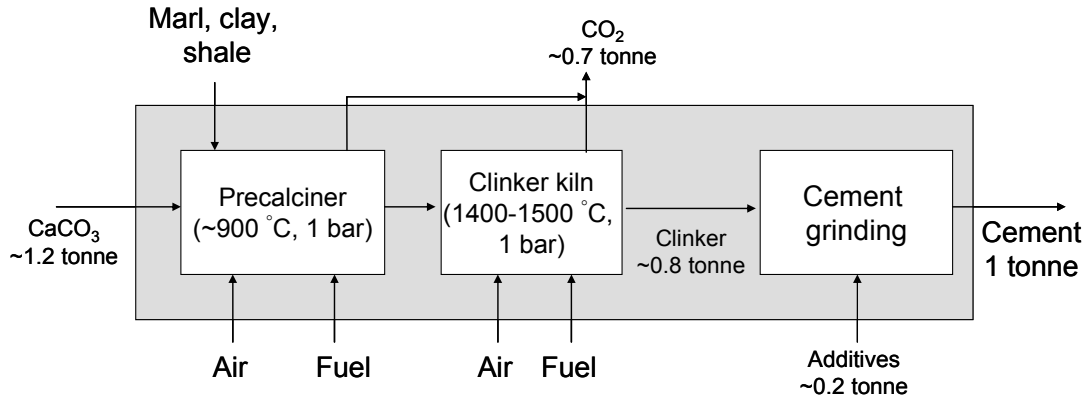
### 2.2.2 Cement sector

Cement production is the second most CO<sub>2</sub> intensive industrial process, accounting for 2 GtCO<sub>2</sub>/year worldwide in 2007 (IEA, 2010a). High CO<sub>2</sub> emission intensity of cement production is not only due to the large energy requirement but also to the emissions from raw materials. The cement production can be categorized into four different process routes: wet process, semi-wet process, semi-dry process and dry process. These process routes are distinguished by the moisture content of the feed going into the kiln (IEA GHG, 2008). The best available technique (BAT)<sup>6</sup> for cement production today is based on the dry process (IEA GHG, 2008). Figure 2-2 presents a schematic diagram of dry process cement production. The raw material mix is mainly comprised of calcium carbonate (CaCO<sub>3</sub>; main component of limestone) with some silica, alumina and iron oxide. The raw material powder is heated up to a sintering temperature of over 1400 °C in a kiln to produce clinker. In this process (calcination), calcium carbonate decomposes into CO<sub>2</sub> and calcium oxide (Eq.(2.1)).



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<sup>6</sup> The term “best available techniques” is defined by the EU Directive (European Commission, 2008) as “the most effective and advanced stage in the development of activities and their methods of operation which indicate the practical suitability of particular techniques for providing in principle the basis for emission limit values designed to prevent and, where that is not practicable, generally to reduce emissions and the impact on the environment as a whole”.



**Figure 2-2:** Typical schematic of cement production (dry process). Typical material flow rates are taken from IPCC (2010b).

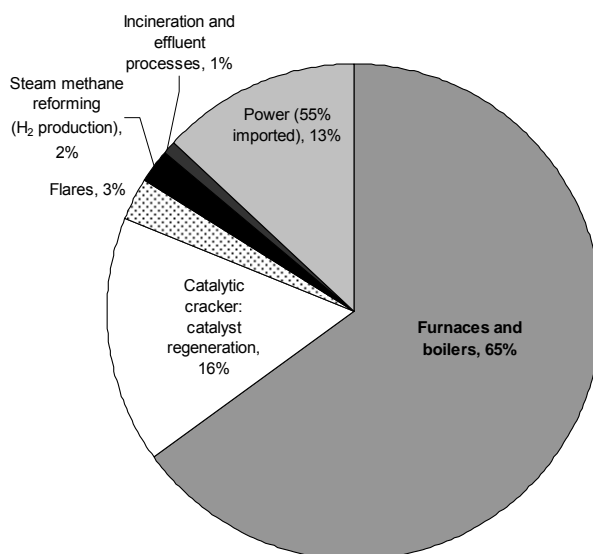
The produced clinker is cooled and grinded with some additives to make cement. The specific heat demand for cement production using the dry process is around 2.9-4.6 GJ/t clinker (FLSMidth, 2006). Specific CO<sub>2</sub> emissions is reported to be in the order of 0.9-1.0 t/t cement (energy and feedstock combined) (IPPC, 2010b). Depending on the clinker/cement ratio, around 60% of the CO<sub>2</sub> originates from the calcination process and the rest is related to fuel combustion (IPPC, 2010b), mostly coal. Specific electricity demand is around 0.32-0.54 GJ/tonne cement (IPPC, 2010b).

Waste-derived fuels are also often used for cement production. In the ENCI cement plant in the Netherlands, for example, more than 95% of the total fuel used is either biomass or alternative fuels such as processed sludge and waste (ENCI, 2008). The CO<sub>2</sub> concentration in the cement plant flue gas is generally around 15-30% (IEA GHG, 2008). Typical kiln size today is around 3000 t/d clinker (IPPC, 2010b), which corresponds to about 1 Mt/yr clinker production.

Waste-derived fuels are also often used for cement production. In the ENCI cement plant in the Netherlands, for example, more than 95% of the total fuel used is either biomass or alternative fuels such as processed sludge and waste (ENCI, 2008). The CO<sub>2</sub> concentration in the cement plant flue gas is generally around 15-30% (IEA GHG, 2008). Typical kiln size today is around 3000 t/d clinker (IPPC, 2009b), which corresponds to about 1 Mt/yr clinker production.

### 2.2.3 Petroleum refineries

CO<sub>2</sub> emissions from petroleum refineries are close to 1Gt/yr worldwide, or about 4% of global total emissions (van Straelen et al., 2010). At the refineries, CO<sub>2</sub> is emitted from various sources (besides hydrogen production) and catalytic crackers. As is the case with chemical industry, onsite electricity and heat production are responsible for the bulk of the CO<sub>2</sub> emissions from petroleum refineries (Wilkinson et al., 2003). A breakdown of CO<sub>2</sub> emissions from petroleum refineries worldwide (Figure 2-3) shows that a large fraction of CO<sub>2</sub> is attributable to onsite heat and power production, and fluid catalytic crackers



**Figure 2-3:** Breakdown of CO<sub>2</sub> emissions from refineries worldwide by source (van Straelen et al., 2010).

(FCCs) (IEA GHG, 1999b). The CO<sub>2</sub> concentration of flue gases from petroleum refineries range between 4% for CHP plants and 12% for plants burning heavy residue (van Straelen et al., 2010).

#### 2.2.4 Chemical and petrochemical sector

The chemical and petrochemical sector is the largest industrial energy user that accounted for 37 EJ final energy use and emitted nearly 1.3 Gt CO<sub>2</sub> in 2007 (IEA, 2007a). The main CO<sub>2</sub> emission sources in the petrochemical sectors are steam boilers and an increasing number of CHP plants (IEA, 2008a). Among various point sources of CO<sub>2</sub> emissions, ethylene production is often considered to be the largest one (IPCC, 2005). Ethylene is produced often near a refinery via steam cracking of light hydrocarbons, e.g. naphtha, LPG and ethane. In Western Europe, naphtha accounts for three quarters of the total steam cracker input, while plants operating on natural gas liquids<sup>7</sup> dominate in the USA (Neelis et al., 2005). Steam cracking is considered to be one of the most energy consuming process and also the most CO<sub>2</sub> emitting process in the chemical and petrochemical industry, accounting for around 180 Mt/yr CO<sub>2</sub> worldwide today (Ren et al., 2006). CO<sub>2</sub> emissions from ethylene production are predominantly attributable to fuel combustion to supply heat for endothermic cracking reactions. Steam cracking process produces not only high value basic petrochemicals but also fuel-grade by-products (e.g., hydrogen and methane) and low value by-products. Fuel grade by-products are combusted to heat the cracker, while low value by-products are partly used to heat the cracker and partly recycled back to the refinery (Ren et al., 2006). The CO<sub>2</sub> partial pressure in ethylene

<sup>7</sup> Natural gas liquids are “hydrocarbons in natural gas that are separated from the gas as liquids through the process of absorption, condensation, adsorption, or other methods in gas processing or cycling plants.” “Generally such liquids consist of propane and heavier hydrocarbons and are commonly referred to as lease condensate, natural gasoline, and liquefied petroleum gases” (US DOE, 2010c).

furnace flue gas is low: the flue gas is emitted at atmospheric pressure with a CO<sub>2</sub> concentration around 12 vol% (Damen et al., 2009; IEA, 2008a). The average ethylene plant capacity per location is about 500 kt/yr in the EU countries (IPPC, 2003b). Regarding specific CO<sub>2</sub> emissions, the majority of plants in Europe operate below 2.1 tCO<sub>2</sub>/t ethylene and a considerable number of plants below 1.5 tCO<sub>2</sub>/t ethylene (IPPC, 2003b).

## 2.3 Methodology

### 2.3.1 System boundaries and performance indicators

Figure 2-4 shows the system boundaries of an industrial process as defined for this study. Besides direct emissions from the industrial process and CO<sub>2</sub> capture and compression, the CO<sub>2</sub> emissions accountable for the import/export of process gas, electricity and steam due to CO<sub>2</sub> capture and compression are also taken into account. This approach incorporates the effect of process modification on material and energy flows of the industrial process due to CO<sub>2</sub> capture. CO<sub>2</sub> transport and storage are excluded from the system boundaries.

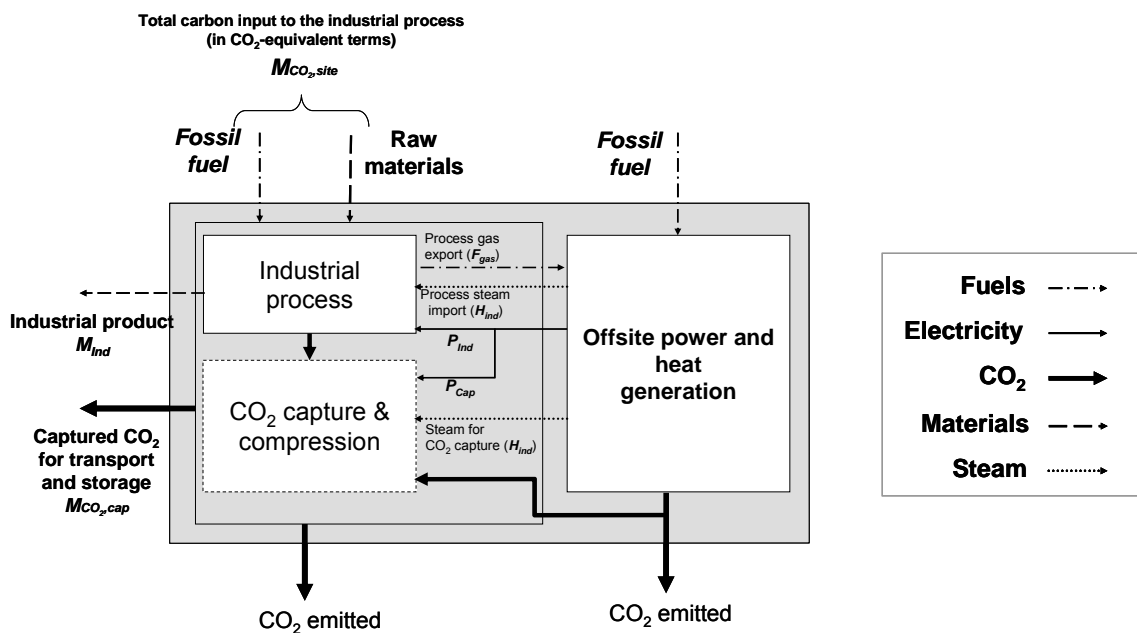


Figure 2-4: System boundaries of an industrial process defined for this study.

In this study, the performance of an industrial plant with CO<sub>2</sub> capture is compared with the identical industrial plant without CO<sub>2</sub> capture.

#### 2.3.1.1 Technical indicators

This study uses specific CO<sub>2</sub> emissions avoided ( $\Delta Em_{CO_2,Sp,Ind}$ ) (tCO<sub>2</sub>/t industrial product) as the main technical indicator of CO<sub>2</sub> capture performance. An important approach here is that in the calculations, the costs and CO<sub>2</sub> emissions both the exported process fuel gas and the imported/exported process steam are estimated as the lost/gained electricity from

the gas turbine combined cycle plant and the steam turbine power plant, respectively.  $\Delta Em_{CO_2,Sp,Ind}$  is defined as (Eq.(2.2)):

$$\Delta Em_{CO_2,Sp,Ind} = \frac{M_{CO_2,cap} - [\Delta M_{CO_2,site} + \{(\Delta P_{Ind} + \Delta H_{Ind} * f_{St,Ind}) + (P_{Cap} + H_{Cap} * f_{St,Cap}) - \Delta F_{gas} * f_{PP}\} * Em_{Sp,Elec}]}{M_{Ind}} \quad (2.2)$$

where:

$M_{CO_2,cap}$ : CO<sub>2</sub> capture rate (tonne/s)

$\Delta M_{CO_2,site}$ : change of total carbon input to the industrial process due to CO<sub>2</sub> capture (tCO<sub>2</sub>-equivalent/s),

$M_{Ind}$ : production rate of the industrial product (tonne/s),

$\Delta P_{Ind}$ : change in the electricity import for the industrial process due to CO<sub>2</sub> capture (MW),

$P_{Cap}$ : electricity import for CO<sub>2</sub> capture and compression (MW),

$\Delta H_{Ind}$ : change in the steam import for the industrial process due to CO<sub>2</sub> capture (MW);

$H_{Cap}$ : steam import for CO<sub>2</sub> capture and compression (MW),

$f_{St,Ind}$ : power equivalent factor for industrial process steam (dimensionless),

$f_{St,Cap}$ : power equivalent factor for CO<sub>2</sub> capture steam (dimensionless),

$\Delta F_{gas}$ : change in the net process gas export from the industrial process to power plants due to CO<sub>2</sub> capture (MW),

$f_{PP}$ : gas-fired power plant efficiency, and

$Em_{Sp,Elec}$ : CO<sub>2</sub> emission factor of grid electricity (tCO<sub>2</sub>/MJ<sub>e</sub>).

For CO<sub>2</sub> capture from heaters and furnaces in the refinery and petrochemical sectors, the industrial product is assumed to be the CO<sub>2</sub>. With regard to the power equivalent factors for steam ( $f_{St,Ind}$  and  $f_{St,Cap}$ ), high pressure steam is assumed for the industrial process steam ( $f_{St,Ind}$ ). For the CO<sub>2</sub> capture steam ( $f_{St,Cap}$ ), there are cases where low pressure steam is used (e.g., post-combustion capture). We considered two cases for the valuation of steam of such quality: (1) low  $f_{St,Cap}$  value based on the steam turbine efficiency (based on exergy content), (2) high  $f_{St,Cap}$  value assuming that the steam from boilers are equally valuable regardless of their pressures and temperatures (based on energy content).

### 2.3.1.2 Economic indicators

We use CO<sub>2</sub> avoidance cost ( $C_{CO_2}$ ) as the main economic indicator for CO<sub>2</sub> capture performance (Eq.(2.3)):

$$C_{CO_2} = \frac{\alpha * \Delta I + \Delta C_{energy} + \Delta C_{O\&M} + \Delta C_{Mat}}{\Delta Em_{CO_2,Sp,Ind} * M_{Ind,annual}} \quad (2.3)$$

where  $\alpha$  is the annuity factor (yr<sup>-1</sup>),  $\Delta I$  is the additional capital requirement (€),  $\Delta C_{energy}$  is the additional annual cost of energy due to CO<sub>2</sub> capture (€/yr),  $\Delta C_{O\&M}$  is the incremental annual operation and maintenance (O&M) costs (€/yr),  $\Delta C_{Mat}$  is the additional annual cost of raw materials due to CO<sub>2</sub> capture (€/yr), and  $M_{Ind,annual}$  is the annual production of the industrial product (t/yr). In some industrial sectors a variety of manufacturing routes can

be found for a single product, e.g. steel. In such cases, costs of manufacture ( $C_{Ind}$ : €/t industrial product) are also calculated (Eq.(2.4)):

$$C_{Ind} = \frac{\alpha * \Delta I + C_{energy} + \Delta C_{O\&M} + \Delta C_{Mat}}{M_{Ind,annual}} \quad (2.4)$$

### 2.3.2 Data collection and standardization of key parameters

An extensive literature review was performed to assess the technical and economic performance of the aforementioned four industrial sectors with and without CO<sub>2</sub> capture. To enable a fair comparison of technologies, some underlying parameters were standardized. We followed the steps suggested by Damen et al. (2006):

#### 2.3.2.1 Indexation

All cost figures were converted to €<sub>2007</sub>. Inflation and material price increases are accounted for by applying the Chemical Engineering Plant Cost Index (CEPCI, 2009). Costs that are reported in U.S. dollars were first standardized to US\$<sub>2007</sub> using CEPCI, then a year-average €/€ currency conversion rate for the year 2007 (1€=1.37\$) (OANDA, 2011) was applied. When the cost data in the literature are expressed in other currencies, they were first converted to US\$ at the currency exchange rate of year the cost data are reported, then updated to US\$<sub>2007</sub> by applying the CEPCI, followed by the conversion to €<sub>2007</sub>.

#### 2.3.2.2 Normalization of CO<sub>2</sub> compression pressure and CO<sub>2</sub> purity

CO<sub>2</sub> compression pressure was standardized at 110 bar. Moreover, we also assumed additional CO<sub>2</sub> purification processes in case the literature reports CO<sub>2</sub> concentrations lower than 95 vol%, a typical concentration at which existing CO<sub>2</sub> pipelines operate (de Visser et al., 2008). When electricity consumption for CO<sub>2</sub> compression is not reported in the original literature, specific power consumption is estimated using the following equation adapted from Damen et al. (2007):

$$E_{s,comp} = \frac{ZRT_1}{M\eta_{is}\eta_m} \frac{N\gamma}{\gamma-1} \left\{ \left( \frac{p_c}{p_l} \right)^{\gamma-1/N\gamma} - 1 \right\} \quad (2.5)$$

where  $E_{s,comp}$  is the specific electricity requirement (kJ/kgCO<sub>2</sub>),  $Z$  is the CO<sub>2</sub> compressibility factor at 1.013 bar, 15 °C (0.9942),  $R$  is the universal gas constant (8.3145 J/(mole K)),  $T_1$  is the suction temperature (313.15 K),  $\gamma$  is the specific heat ratio ( $c_p/c_v$ ) (1.294),  $M$  is the molar mass (44.01 g/mole),  $\eta_{is}$  is the isentropic efficiency (80%),  $\eta_m$  is the mechanical efficiency (99%),  $p_l$  is the suction pressure (101 kPa),  $p_c$  is the critical pressure (7,380 kPa),  $p_l$  is the outlet pressure (11,000 kPa),  $N$  is the number of compressor stages (= 4),  $\rho$  is the density of CO<sub>2</sub> during pumping (630 kg/m<sup>3</sup>), and  $\eta_P$  is the pump efficiency (75%). When the compression pressure reported in the literature differs from the value used in this study (110 bar), we also adjusted the specific electricity consumption using Eq.(2.5). Note that the assumption on the CO<sub>2</sub> compressibility factor is conservative because the value becomes smaller at higher pressures.

An additional CO<sub>2</sub> purification process was assumed in case CO<sub>2</sub> purity reported in the literature is lower than 95 vol%<sup>8</sup>. In such cases, CO<sub>2</sub> capture rate was adjusted by a multiplication factor  $\eta_{Rec}$  as some CO<sub>2</sub> would be vented together with the removed impurities in the purification process.  $\eta_{Rec}$  is assumed to be 90% for  $\chi_{Ref} < 75$  vol%, 92% for  $75 \text{ vol}\% \leq \chi_{Ref} < 80$  vol%, and 94% for  $80 \text{ vol}\% \leq \chi_{Ref}$  as suggested in Ritter et al. (2009).

### 2.3.2.3 Normalization of cost figures

With regard to capital investment, we considered total capital requirement (TCR), which includes the following cost components:

- Process plant cost (costs for the equipment pieces and their installation) plus engineering fees and contingencies;
- Owner costs (royalties, preproduction costs, inventory capital, land costs and site preparation) and interests during construction.

### 2.3.2.4 Normalization of plant scales

Capital costs are strongly influenced by the plant capacity. For a fair comparison of various CO<sub>2</sub> capture technologies, it is necessary to standardize the plant scale. The base scales for various industrial processes were determined based on the literature review. Capital costs were standardized by applying a generic scaling relation as shown in Eq.(2.6):

$$\frac{Cost_A}{Cost_B} = \left( \frac{Scale_A}{Scale_B} \right)^{SF} \quad (2.6)$$

where  $SF$  is the scaling factor.

The technical and economic parameters and variables common for all industrial sectors investigated in this paper are shown in Table 2-2. Other parameters that are specific for individual sectors are presented in later sections of the paper.

After the performance data from the literature are standardized, the literature sources were selected using a number of criteria to obtain figures on the CO<sub>2</sub> capture performance. This study uses the criteria proposed by Damen et al. (2006). Preference is given to recent, highly detailed and transparent studies that ideally include data on production and capture efficiencies, capital cost and O&M costs. Note that advanced CO<sub>2</sub> capture technologies are generally studied in less detail. Technical performance figures are often forecasts, and cost projections are highly uncertain. Cost figures for advanced technologies are generally forecasted values based on the assumption that the technologies are mature, and do not explicitly account for the effect of technological learning (Damen et al., 2007).

---

<sup>8</sup> 95 vol% is a typical concentration at which existing CO<sub>2</sub> pipelines operate (de Visser et al., 2008).



**Table 2-2:** Parameters standardized for technical and economic performance calculations in this study.

<i>Parameters and variables</i>	<i>Unit</i>	<i>Nominal value</i>	<i>Range used for sensitivity analysis</i>
Annual plant operation time			
Cement sector	h/yr	8000 <sup>1)</sup>	
Other sectors	h/yr	8500	
Economic plant lifetime	yr	20	Total ± 30%
Real interest rate <sup>2)</sup>		10%	
Total capital requirement <sup>3)</sup>	% - total plant cost	110	for annualized
Total plant cost <sup>3)</sup>	% - process plant cost	130	capital cost
Energy prices			
Non-coking coal <sup>4)</sup>	€/GJ	2.6	2-3.2
Natural gas <sup>4)</sup>	€/GJ	8	5-11
Electricity <sup>5)</sup>	€/MWh	55	40-70
Energy content of fuels (LHV) <sup>6)</sup>			
Coal	MJ/kg	26.7	
CO <sub>2</sub> emission factors			
Natural gas <sup>6)</sup>	g/MJ LHV	56	
Coal <sup>6)</sup>	g/MJ LHV	95	
Grid electricity ( $Em_{Sp,Elec}$ ) <sup>7)</sup>	g/kWh	400	320-480
Power equivalent factor for steam <sup>8)</sup>			
High temperature steam ( $f_{St,Ind}$ )		0.45	0.4-0.5
Low pressure steam ( $f_{St,Cap}$ )		0.23	0.21-0.25
Gas turbine combined cycle power plant efficiency ( $f_{PP}$ )		0.5	0.4-0.6

<sup>1)</sup> The literature values range between 7350 h/yr and 8000 h/yr (Hegerland et al., 2006; IEA GHG, 2008).

<sup>2)</sup> The interest rates for the industrial plants found in the literature were 10% for cement production (IEA GHG, 2008), 10% for methanol and hydrogen production (Hamelinck and Faaij, 2002), 10-15% for paper mills (Möllersten et al., 2006; Möllersten et al., 2003), and 12% for iron and steel production (Gielen, 2003).

<sup>3)</sup> Process plant cost (PPC) comprises equipment cost and installation costs. Total plant cost (TPC) comprises PPC and engineering fees and contingencies. Total capital requirement (TCR) comprises TPC, owner costs and interests during construction. The values are within the ranges observed for power plant construction (van Horssen et al., 2009).

<sup>4)</sup> Nominal value is from IEA (2010b). The high and low values assumed here agree with those forecasted by the IEA for the EU, the US and Japan for years between 2020 and 2030 (IEA, 2010b).

<sup>5)</sup> Electricity price for large industries differs significantly by country, from 0.028 €/kWh in Russia (in 2006) to 0.177 €/kWh in Italy (in 2007). The price used in this study is similar to that in the USA (0.048 €/kWh in 2007), South Korea (0.052 €/kWh in 2007) and Poland (0.061 €/kWh in 2007) (US DOE, 2010b).

<sup>6)</sup> IPCC (2006).

<sup>7)</sup> The changes in electrical consumption due to CO<sub>2</sub> capture are likely to affect base load power generation by base-load fossil fuel-fired power plants. We assumed that in the industrialized world where CCS is also deployed for industrial processes, around 40% of base-load fossil fuel-fired power plants are equipped with CO<sub>2</sub> capture. Assuming an average 75% CO<sub>2</sub> avoidance rate by CO<sub>2</sub> capture, we estimated that the CO<sub>2</sub> emissions from base-load fossil fuel-fired power plants would be reduced by 30% due to CO<sub>2</sub> capture. The nominal value assumes that natural gas- and coal-fired power plants share the electricity

generation 50% each. The low end value corresponds to a ratio of 80:20 between natural gas power plants and coal power plants, and the high end value corresponds to a ratio of 20:80.

- <sup>8)</sup> The low  $f_{St,Cap}$  value is taken from Bolland and Undrum (2003) for NGCC plants, assuming a steam temperature of 140 °C. The high  $f_{St,Cap}$  value and the  $f_{St,Ind}$  value is based on a typical electrical conversion efficiency for steam turbines.

### 2.3.3 Sensitivity analysis

Material and energy flows in industrial processes are often more complex than in power plants. In particular, imports and exports of various valuable products such as steam, fuel-quality process gas and electricity are common for industrial processes and they are affected by CO<sub>2</sub> capture. It is therefore probable that the assumptions on the economic values and the CO<sub>2</sub> emission factors of these energy products affect CO<sub>2</sub> avoidance cost significantly. Thus, the sensitivity of the results to the parameter values needs to be assessed. The parameters considered for the sensitivity analysis are: all energy prices, annualized capital cost, grid electricity CO<sub>2</sub> emission factor, power equivalent factors for steam ( $f_{St,Ind}$  and  $f_{St,Cap}$ ), and gas turbine combined cycle efficiency ( $f_{PP}$ ). The value ranges of the parameters are presented in Table 2-2.

To assess the combined effect of the parameters on the results, we also performed a sensitivity analysis by applying a general equation for uncertainty propagation as described in, e.g., (Brealey and Myers, 2000; Taylor, 1997). For a performance indicator  $C=F(X_1, X_2, \dots, X_n)$ , which is a function of variables  $X_1, X_2, \dots, X_n$ , the standard deviation of  $C$  ( $\sigma_C$ ) is calculated as follows:

$$\sigma_C = \sqrt{\sum_{i=1}^n \left( \frac{\partial C}{\partial X_i} \right)^2 \sigma_{X_i}^2 + 2 \sum_{i=1}^{n-1} \sum_{j=i+1}^n \frac{\partial C}{\partial X_i} \frac{\partial C}{\partial X_j} \sigma_{X_i} \sigma_{X_j} r_{ij}} \quad (2.9)$$

where  $\sigma_{X_i}$  is the standard deviation of variable  $X_i$ ,  $r_{ij}$  is the correlation coefficient between two variables  $X_i$  and  $X_j$ . In this study it is assumed that there is some degree of correlation among energy product prices. Particularly, a high degree of correlation between natural gas and electricity prices has been observed for the OECD countries (Yang and Blyth, 2007). The correlation coefficients between natural gas price and electricity, between natural gas price and coal price, and between coal price and electricity price, are assumed to be 0.9, 0.5, and 0.5, respectively. These coefficients only reflect our qualitative understanding of the relations between the parameters. The correlation coefficients for other combinations of variables are assumed to be zero.

### 2.3.4 Future CO<sub>2</sub> price

In this study, the obtained economic performance results are compared with a possible future CO<sub>2</sub> price of 30-75 €<sub>2007</sub>/tonne. The range is based on the IEA's World Energy Outlook (WEO) 2010 (IEA, 2010b). In the reference scenario in which carbon pricing is limited to the power and industry sectors in EU countries, CO<sub>2</sub> price in the EU Emissions Trading System is projected to reach 43\$<sub>2008</sub>/tonne (30€<sub>2007</sub>/tonne) in 2020 and 54\$<sub>2008</sub>/tonne (38€<sub>2007</sub>/tonne) in 2030. Alternatively, in a scenario where global

greenhouse gas (GHG) emissions are reduced to half in 2050 compared to today's level (450 scenario), the study estimated that the CO<sub>2</sub> price reaches 50\$<sub>2007</sub>/tonne (37€<sub>2007</sub>/tonne) in 2020 and 110\$<sub>2007</sub>/tonne (75€<sub>2007</sub>/tonne) in 2030 in all the OECD countries and non-OECD EU countries.

## **2.4 Assessment of CO<sub>2</sub> capture technologies by industrial sector**

### **2.4.1 Iron and steel sector**

For CO<sub>2</sub> capture from the iron and steel sector, this study exclusively focuses on ironmaking process, i.e., blast furnace and smelting reduction processes, because large fraction of the total carbon input to the iron and steel plant flows through the ironmaking process. Firstly, different ironmaking processes and the effects of integrating CO<sub>2</sub> capture on the energy and material flows of the ironmaking processes are described. Secondly, the technical and economic performance of various CO<sub>2</sub> separation technologies proposed in the literature are presented. Figure 2-5 presents four ironmaking processes investigated in this paper.

#### **2.4.1.1 Ironmaking with CO<sub>2</sub> capture: changes in energy and material flows**

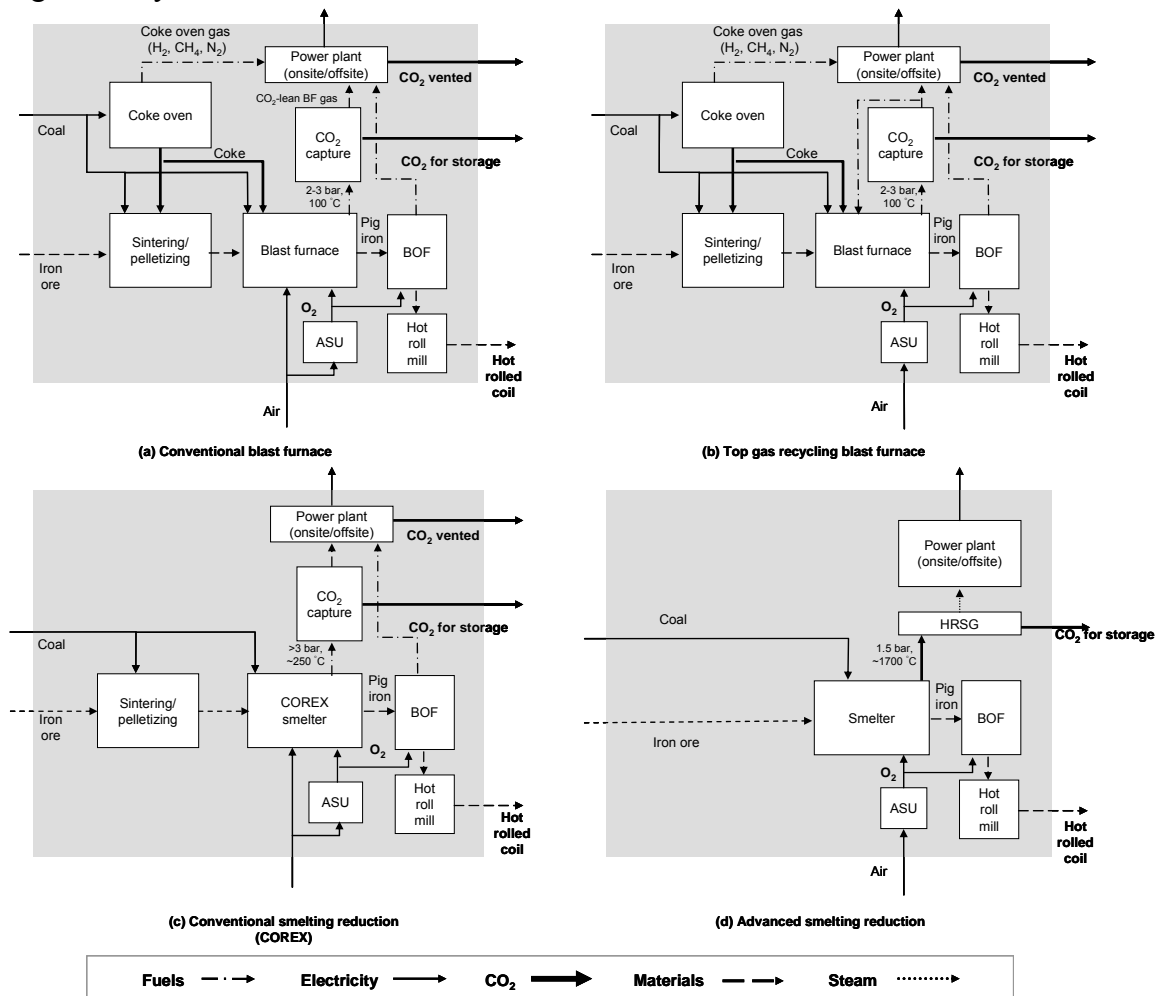
##### *Blast furnace: add-on CO<sub>2</sub> capture (no modification to the blast furnace)*

In the blast furnace (BF) -based process (Figure 2-5 (a)), around 70% of the carbon introduced into the process flows through the BF (Farla et al., 1995). BF gas has a pressure of about 2-3 bar (Farla et al., 1995) and contains CO<sub>2</sub> (17-25%), CO (20-28%), H<sub>2</sub> (1-5%), N<sub>2</sub> (50-55%) (IPPC, 2001). After dust removal, BF gas flows through expansion turbines to recover some power before being distributed as a fuel (Farla et al., 1995). There are two approaches for capturing CO<sub>2</sub> from BF as add-on technologies (no modification of the BF): (1) capture directly from BF gas, (2) capture after the conversion of CO to CO<sub>2</sub>. In the first case, less than 50% of the total carbon contained in the BF gas is captured because the remaining fraction is in the form of CO, which is not captured. For the second approach, CO can be either combusted or shifted with steam to be converted to CO<sub>2</sub>. The main advantage is on higher CO<sub>2</sub> capture rates.

##### *Blast furnace: process-integrated CO<sub>2</sub> capture (modification of the blast furnace)*

When the modification of the air-blown BF is concerned, the Top Gas Recycling Blast Furnace (TGRBF) (Figure 2-5 (b)) technology may enable more energy efficient blast furnace operation while incorporating CO<sub>2</sub> capture in the ST/MT. TGRBF has been studied for decades to reduce coke consumption (Nishio and Miyashita, 1973), and today it is being developed for commercialization by the European Ultra Low CO<sub>2</sub> Steelmaking (ULCOS) program (ULCOS, 2011). Since the technology is based on conventional blast furnace, TGRBF can be commercialized in the ST/MT (Birat and Lorrain, 2009; Torp, 2005). TGRBF is oxygen-blown so its top gas contains little nitrogen and is rich in CO (40-50 vol%) and CO<sub>2</sub> (about 35 vol%), thus enabling the top gas to be recycled as a

reducing agent after CO<sub>2</sub> is removed. Consequently, the coke consumption can be reduced significantly.



**Figure 2-5:** Schematics of different ironmaking processes with CO<sub>2</sub> capture.

For TGRBF, two CO<sub>2</sub> removal technologies have been tested to date on a pilot scale: monoethanolamine (MEA) (Tseitlin et al., 1994) and vacuum pressure swing adsorption (VPSA) (Danloy et al., 2009). The test results show a reduction of the total carbon input by up to 30% compared to conventional air-blown BF and the onsite CO<sub>2</sub> emissions can be reduced by up to 76% compared with conventional BF (Danloy et al., 2009). The overall CO<sub>2</sub> emissions reduction will, however, be somewhat smaller because the reduced BF gas export needs to be compensated for, and a large amount of electricity is required to produce high-purity oxygen. The recycling of H<sub>2</sub> instead of CO after shift reaction is also theoretically possible as H<sub>2</sub>, like CO, is also a reductant. Such systems have been proposed in (Gielen, 2003; Hooey et al., 2010) and may prove to be technologically viable in the LT (Treadgold, 2009). TGRBF can be retrofitted to conventional blast furnaces, although it may require major modifications to the furnace. In addition, the steel production loss during retrofitting the blast furnace period may be significant. The consequent economic losses, however, have not been investigated in the publicly available literature.

### *Smelting reduction*

CO<sub>2</sub> capture from smelting reduction process gas (Figure 2-5 (c)) is considered to be more cost-effective than that from air-blown BF gas because of the higher CO<sub>2</sub> concentration, around 25-35 vol%. CO<sub>2</sub> removal from the smelter process gas is already in operation at commercial scale. At the Saldanha steel plant in South Africa, CO<sub>2</sub> is removed from the COREX process gas before it is used as reduction gas to produce direct reduced iron (DRI) (Wieder et al., 2005). For the FINEX process, the process gas is recycled after the removal of CO<sub>2</sub> as part of cost-effective plant operation (POSCO, 2008a). Part of CO-rich gas is recycled for the pre-reduction of iron ore. Currently, the CO<sub>2</sub> removed from both Saldanha and FINEX plants are vented. As is the case with BF, CO<sub>2</sub> can be captured either directly from the smelter process gas or after the conversion of CO to CO<sub>2</sub>.

In the LT, advanced smelting reduction technologies will be operated without any N<sub>2</sub> input (Figure 2-5(d)). One of these technologies being developed under the ULCOS programme is the HIsarna process (Birat and Lorrain, 2009; Birat et al., 2009). The main characteristics of the HIsarna process are that both the iron ores and the non-coking coal can directly be input to the smelter. In this setup, the input carbon is fully oxidized within the smelter so that CO<sub>2</sub> removal is unnecessary. Some heat is recovered from the off-gas to generate steam. The HIsarna process will be further developed in a pilot plant at Tata steelworks in IJmuiden (the Netherlands) in the next couple of years (ULCOS, 2011). The HIsarna process is expected to capture 95% of the carbon input to the iron making process (Burke, 2009).

### *Sector-specific assumptions*

The technical and economic performance of various steelmaking processes with CO<sub>2</sub> capture is calculated for a plant scale of 4 Mt/yr hot rolled coil production. Table 2-3 presents the input rates of raw materials, intermediate products, energy products and labor to individual processes in the iron steel plants for the calculation of specific material, energy and labor use per tonne of hot rolled coil. The production efficiency of BF is not expected to improve significantly in the future because it is already very efficient (Schmöle and Lünge, 2004). For iron making, a tradeoff between the amount of total coal and coke input and the amount of exported gas can be seen between different processes. Table 2-4 presents the capital cost figures of individual processes in the iron and steel production, and Table 2-5 presents prices of raw materials for iron and steel production.

**Table 2-3:** Raw materials, intermediate products, energy products and labor inputs to individual processes in the iron and steel plants. The values are from Daniëls (2002), unless otherwise stated.

Process Unit output	Coke plant <sup>1)</sup> [1t coke]	Sinter plant [1t sinter]	Pellet plant [1t pellet]	ASU <sup>2)</sup> [1t O <sub>2</sub> ]	Pig iron production				Steel production (BOF) <sup>3)</sup> [1t hot rolled coil]
					BF <sup>3)</sup> [1t pig iron]	TGRBF <sup>5)</sup> (excl. CO <sub>2</sub> capture)	COREX	Advanced smelting reduction <sup>4)</sup>	
<i>Raw material inputs</i>									
Air	-	-	-	5.00	-	-	-	-	-
Coking coal (t)	1.25	-	-	-	-	-	-	-	-
Steam coal (t)	-	-	-	-	0.20	0.20	0.95	0.68	-
Fine ore (t)	-	0.90	0.95	-	-	-	-	1.5	-
Lump ore (t)	-	-	-	-	0.17	0.17	-	-	-
Limestone (t)	-	0.12	0.05	-	-	-	0.05	0.07	-
Scrap (t)	-	-	-	-	-	-	-	-	0.16
Byproducts (GJ)	-1.6	-	-	-	-	-	-	-	-
Gas (GJ)	-4.8	0.17	0.40	-	-3.6	-0.02	-11.2	-	-0.63
Steam (GJ)	-	-	-	-	-	-	-	-	-
Electricity (GJ)	0.16	0.04	0.21	0.94 <sup>2)</sup>	0.12	0.12	-	-0.90	0.09
<i>Intermediate material inputs</i>									
Coke (t)	-	0.05	-	-	0.32	0.21	-	-	-
Sinter (t)	-	-	-	-	1.1	1.1	-	-	-
Pellet (t)	-	-	-	-	0.49	0.49	1.50	-	-
O <sub>2</sub> (t)	-	-	-	-	0.08	0.35	0.76	0.98	0.072
Pig iron (t)	-	-	-	-	-	-	-	-	0.86
CO <sub>2</sub> capture (t) <sup>6)</sup>									
Excl. CO-derived	-	-	-	-	0.44	0.83	1.0	1.8	-
Incl. CO-derived	-	-	-	-	0.89	Not considered	2.5	-	-
Labor (man hour)	0.55	0.14	0.14	0.1	0.25	0.25	0.7	0.25	0.3

- <sup>1)</sup> All values are from Diemer et al. (2004), except for labor requirements. Byproducts include coal tar, benzene, toluene and xylene.
- <sup>2)</sup> Specific energy consumption for ASU is obtained from Chiesa et al. (2005).
- <sup>3)</sup> The values are from (2009), except for labor requirements and gas consumption. The gas consumption value for BF is from Schmöle and Lungen (2004).
- <sup>4)</sup> The process includes steam turbines for power generation. Values are from Treadgold (2009), except for labor, fine ore and limestone requirements. The labor requirement value is taken from Daniëls (2002) for cyclone converter furnace. Fine ore and limestone requirements are from Link (2008).
- <sup>5)</sup> All values are from Birat (2006), except for labor requirements and gas consumption. The gas consumption value is from Schmöle and Lungen (2004).
- <sup>6)</sup> CO<sub>2</sub> capture rates for air-blown BF are based on the carbon flows indicated in Schmöle and Lungen (2004), and an assumption of CO<sub>2</sub>/CO molar ratio of 1:1 as suggested in, e.g., (Farla et al., 1995; Lie et al., 2007). It is assumed that CO<sub>2</sub> is not captured from the BF gas that is used for BF stove. CO<sub>2</sub> capture rate for TGRBF is based on the carbon flows indicated in Schmöle and Lungen (2004), and an assumption of CO<sub>2</sub>/CO molar ratio of 36:47 as suggested in various ULCOS studies, e.g., (Duc et al., 2007; Lie et al., 2007; Torp, 2005). The CO<sub>2</sub> capture rate for COREX gas (excluding CO-derived CO<sub>2</sub>) is based on a number of assumptions. 950kg coal input per tonne of pig iron equals to about 2800 kg CO<sub>2</sub>. CO/CO<sub>2</sub> molar ratio in the COREX gas is between 1.1 and 1.5 (Lampert and Ziebig, 2007). Considering the CO<sub>2</sub> capture efficiency of the solvents or sorbents and the carbon content of 174 kg CO<sub>2</sub>-eq./t pig iron (Schmöle and Lungen, 2004), 1tCO<sub>2</sub> captured per tonne of pig iron is a reasonable capture rate. The CO<sub>2</sub> capture rate for advanced smelting reduction assumes that all carbon input, except for the carbon contained in the pig iron, to the smelter is captured.

**Table 2-4:** Capital costs of individual processes in the iron and steel manufacture. Source: Gordon et al. (2009), unless otherwise stated.

<i>Process</i>	<i>Base scale</i>	<i>Capital cost [M€]</i>	<i>Scaling factor</i>
Coking <sup>1)</sup>	1.9 Mt coke/yr	490	0.83
Pelletizing <sup>2)</sup>	2.8 Mt pellet/yr	250	1
Sintering <sup>3)</sup>	6.6 Mt sinter/yr	530	1
O <sub>2</sub> separation <sup>4)</sup>	1839 tO <sub>2</sub> /d	37	0.7
<i>Ironmaking</i>			
Blast furnace <sup>5), 6)</sup>	2.8 Mt pig iron/yr	610	1
COREX <sup>5)</sup>	1 Mt pig iron/yr	230	1
Advanced smelting reduction <sup>7)</sup>	1 Mt pig iron/yr	230	1
<i>Steelmaking</i>			
BOF <sup>5)</sup>	2.8 Mt crude steel/yr	380	1

<sup>1)</sup> The presented scaling factor is derived from the values given in this table and the capital cost of 450 M€ for a plant scale of 1.7Mt coke/yr.

<sup>2)</sup> Scaling factor is derived from the data presented in the table and the other data reported in the same study (870 M€ for 9.4 Mt pellet/yr). We therefore concluded that scaling effect is negligible.

<sup>3)</sup> The same scaling factor for the pelletizing plant is used.

<sup>4)</sup> Values are from Kreutz et al. (2005).

<sup>5)</sup> Two units are assumed for the plant of 5.6 Mt pig iron/yr, so it is assumed that there are no economic benefits for the scale-up for the plant scales considered in this study.

<sup>6)</sup> TGRBF is assumed to be identical to air-blown BF.

<sup>7)</sup> There are no capital cost estimates available for the advanced smelting reduction plant in the publicly available literature. Therefore the specific capital cost is assumed to be the product of the specific capital cost for the conventional BF used in this study and the ratio of specific capital cost between converter cyclone furnace and conventional BF given in Daniëls (2002).

**Table 2-5:** Price of raw materials for iron and steel production

<i>Material</i>	<i>Unit</i>	<i>Value</i>
Fine ore <sup>1)</sup>	€/tonne	65
Lump ore <sup>2)</sup>	€/tonne	86
Metal scrap <sup>3)</sup>	€/tonne	300
Coking coal <sup>4)</sup>	€/GJ	3.75
Lime <sup>5)</sup>	€/tonne	3
Labor <sup>6)</sup>	€/(man h)	20
Byproducts <sup>7)</sup>	€/GJ	5.3

<sup>1)</sup> Average price for 2008 for Europe from Ponte da Madeira, Brazil (Steelonthenet.com, 2010).

<sup>2)</sup> Lump ore prices are assumed to be 33% more expensive than the fine ore price (Daniëls, 2002).

<sup>3)</sup> Metal scrap price in 2008 ranged between 140-450 €/tonne (Steelonthenet.com, 2010).

<sup>4)</sup> Between years 2000 and 2008, coke prices were higher than steam coal prices (per unit weight) in most of the major steel manufacturing countries: 10-50% in Japan, 30-100% in South Korea, and 80-190% in the USA (US DOE, 2010a). In this study, we assume that the coke price (per tonne) is 50% more expensive than the steam coal price.

<sup>5)</sup> Price data for cement production (IEA GHG, 2008).

<sup>6)</sup> Labor cost differs significantly from country to country. Our assumption agrees well with the labor costs for a number of industrialized countries such as Italy, Japan and the USA (Steelonthenet.com, 2010).

<sup>7)</sup> In Daniëls (2002), the economic value of the byproducts, e.g., coal tar, benzene, toluene and xylene, are assumed to be two-thirds of that of gases.



#### **2.4.1.2 Overview of CO<sub>2</sub> capture technologies**

Table 2-7 shows key parameters for various CO<sub>2</sub> capture technologies for the iron and steel sector. The values include CO<sub>2</sub> compression to 110 bar. A capital cost scaling factor of 0.85 is used, unless otherwise stated. All references provide both the energy consumption data and the cost data, except for the reference (Tobiesen et al., 2007), which only provides specific steam consumption data. Note that a significant number of low-CO<sub>2</sub> iron and steel making routes have been assessed and compared in the ULCOS programme (Birat and Lorrain, 2009; Birat et al., 2009), but the presented results are either aggregated or in relative terms. Therefore, we could make only limited use of them. All CO<sub>2</sub> capture options presented here are, in principle, applicable for retrofit. Chemical absorption is appropriate for CO<sub>2</sub> capture from gases with low CO<sub>2</sub> partial pressure, which is the case for BF and COREX gases. Besides the techno-economic studies presented in Table 2-6, the energy performance of the following chemical absorption technologies have been assessed and compared for BF gas by Tobiesen et al. (2007): monoethanolamine (MEA), MEA/piperazine mix and sterically hindered primary amines. Advanced tertiary amine solvents and sterically hindered primary amines react more like tertiary amines because of the molecular structure, and they are expected to bring down the CO<sub>2</sub> capture heat requirement below 2.5 GJ/t CO<sub>2</sub> captured (Fujioka, 2010; Tobiesen et al., 2007).

Some CO<sub>2</sub> capture technologies such as physical absorption, pressure swing adsorption (PSA) and vacuum pressure swing adsorption (VPSA), operate at higher CO<sub>2</sub> partial pressures. This CO<sub>2</sub> capture technology can therefore be used to separate CO<sub>2</sub> from the process gas after water-gas shift (WGS) reaction, which is operated at about 20 bar (Gielen, 2003). As described earlier, CO<sub>2</sub> removal from smelting reduction processes has been practiced for years. At Saldanha steel plant, CO<sub>2</sub> is removed from the COREX process gas by VPSA (Wieder et al., 2005). For the FINEX process, CO<sub>2</sub> in the process gas is removed using PSA technique (POSCO, 2008a, b)<sup>9</sup>. These CO<sub>2</sub> capture technologies require process gas compression. Their performance seems to depend on how the net electricity consumption for gas compression can be minimized. Ho et al. (2010) argued that the CO<sub>2</sub> capture cost is similar to that of a chemical absorption process using MEA, due to the significantly increase in power consumption for additional feed gas compressor (Ho et al., 2010). When the process gas after CO<sub>2</sub> removal is used in power plants, however, Lampert et al. (2010) indicated that the energy penalty for add-on CO<sub>2</sub> capture from BF and COREX gases can be partially offset because the exported gas is compressed anyway prior to combustion in the power plant. The study also shows a small increase in the power plant electrical output due to the improvement in fuel quality (Lampert et al., 2010).

With regard to the capture of CO-derived CO<sub>2</sub> in the iron-making process gas, there are many technically feasible options that have not been studied for iron-making applications.

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<sup>9</sup> Although IEA (2009) stated that the FINEX process uses chemical absorption technique for CO<sub>2</sub> removal, there is no other source that confirms this.

Besides the study by Wiley et al. (2010) presented in Table 2-6, the relative economic performance of post-combustion capture from the BF combustion flue gas against other CO<sub>2</sub> capture technologies for BF is reported in Birat and Lorrain (2009). Oxyfuel combustion of BF gas using a specially made gas combustor is planned to be tested in the Netherlands (CATO, 2010). Both the technical details of the demonstration plant and the feasibility study results for this technique, however, are not publicly available.

Regarding CO<sub>2</sub> capture from iron-making process gases after WGS, the process gas after CO<sub>2</sub> removal is H<sub>2</sub>-rich. This could enable higher electrical efficiency when it is used for power generation, but also requires important modifications in the gas turbines (Lampert et al., 2010). Vlek (2007) also suggested that shifting all BFG affects the composition of the processed gas in such a manner that it will not fit the existing installation anymore. For both BFG and COREX process gas, the use of shift reaction reduces specific CO<sub>2</sub> capture energy consumption. For BFG, however, the reduction in energy consumption is limited. As a result, the economic benefit of shifting CO is offset by the additional capital costs for shift reactor, feed gas compressor and H<sub>2</sub> turbine (Ho et al., 2010). For COREX process gas, in contrast, the economic benefit of shifting CO is significant because of larger amount of CO per unit of flue gas and higher CO concentration compared to BFG (Ho et al., 2010).

A potentially promising WGS technology in the mid-term future is the Sorption Enhanced Water Gas Shift (SEWGS) process. The SEWGS process drives the WGS reaction ( $\text{CO} + \text{H}_2\text{O} \leftrightarrow \text{CO}_2 + \text{H}_2$ ) to the right by removing CO<sub>2</sub> from the reaction gas using a CO<sub>2</sub> adsorbent (e.g. hydrotalcites) in the reactor. A multi-column test rig has been in operation since 2007 at the Energy research Centre of the Netherlands (ECN, 2009). Currently the SEWGS has been researched for the application of CO<sub>2</sub> capture from natural gas (CAESAR, 2010; CCP, 2005) and for IGCC (van Dijk et al., In Press), but it can theoretically also be applied to refinery residues and off-gases of blast furnaces in the steel industry (CAESAR, 2010). One possible LT technology may be the water gas shift membrane reactor (WGSMR), which enables the conversion of CO to CO<sub>2</sub> and the separation of H<sub>2</sub> from CO<sub>2</sub> taking place in a single reactor.

### *Assumptions*

The technical and economic performances of various steelmaking processes with CO<sub>2</sub> capture are calculated for a plant scale of 4 Mt/yr hot rolled coil production. Table 2-7 presents the key parameters of CO<sub>2</sub> capture technologies for different steelmaking processes.

**Table 2-6:** Key parameters of various CO<sub>2</sub> capture options for blast furnace and other steelmaking processes reported in the literature.

<i>CO<sub>2</sub> capture technique</i>		<i>Specific energy consumption [GJ/tCO<sub>2</sub> captured]</i>		<i>Specific capital cost [€/tCO<sub>2</sub> captured/yr]</i>	<i>References</i>
		<i>Steam</i>	<i>Electricity</i>		
<i>Air-blown BF</i>					
Chemical absorption	MEA	3.2-4.4	0.51-0.55	70-90	Fujioka (2010), Ho et al. (2010), Tobiesen et al. (2007)
	KS-1	3.0	0.5-0.51	60-70	RITE (2005)
	Other amines	2.5-4.7	0.35-0.51	70-80	Farla et al. (1995), RITE (2005), Tobiesen et al. (2007)
	Advanced solvents	2.2-2.5	0.50	70	Fujioka (2010), Tobiesen et al. (2007)
Physical absorption (Selexol)		-	0.77	180	Vlek (2007)
Shift + physical absorption (Selexol) <sup>1)</sup>		0.50 <sup>2)</sup> -0.62	0.63-0.91 <sup>3)</sup>	20-190	Gielen et al. (2003), Ho et al. (2010), Vlek (2007)
Selective carbon membrane		-	0.69-0.89	80	Lie et al. (2007)
Hydrate crystallization		-	4.7	220	Duc et al. (2007)
<i>Top gas recycling BF</i>					
Chemical absorption (MEA)		3.3	0.62	60	Torp (2005)
Vacuum pressure swing adsorption <sup>4)</sup>		-	0.94	50	Torp (2005)
Physical absorption (Selexol)		0.21	0.93	60	Torp (2005)
Selective carbon membrane		-	0.79-0.88	60-90	Lie et al. (2007), Torp (2005)
Hydrate crystallization		-	1.5	70	Duc et al. (2007)
<i>COREX</i>					
Chemical absorption (MEA)		4.4	0.45	40	Ho et al. (2010)
Physical absorption (Selexol)		-	0.97	40	Lampert and Ziebig (2007), Lampert et al. (2010)
Shift + physical absorption (Selexol) <sup>1)</sup>		0.63	0.60 <sup>5)</sup>	20-110	Gielen et al. (2003), Ho et al. (2010)

<sup>1)</sup> CO<sub>2</sub> capture using WGS improves the quality of exported fuel gas (higher hydrogen content). Gielen (2003) assumes a benefit of 0.28 GJ/t CO<sub>2</sub> captured. However, our study took a conservative approach of excluding this benefit. The power consumption and cost figures differ significantly between Gielen (2003) and Vlek (2007). The difference in specific power consumption is partly because the study by Vlek (2007) takes into account not only gas compression and steam consumption but also power consumption for solvent circulation and loss of heating value due to shift reaction. The cost scaling factor was calculated to be 0.68, based on the figures from Danloy et al. (2003).

<sup>2)</sup> Steam required for water-gas shift is recovered from waste heat streams.

<sup>3)</sup> Ho et al. (2010) reports 1.36 GJ/tCO<sub>2</sub> captured, including the steam consumption converted to electrical terms.

<sup>4)</sup> CO<sub>2</sub> capture efficiency of VPSA is from (2008).

<sup>5)</sup> Ho et al. (2010) reports 0.83 GJ/tCO<sub>2</sub> captured, including the steam consumption converted to electrical terms.

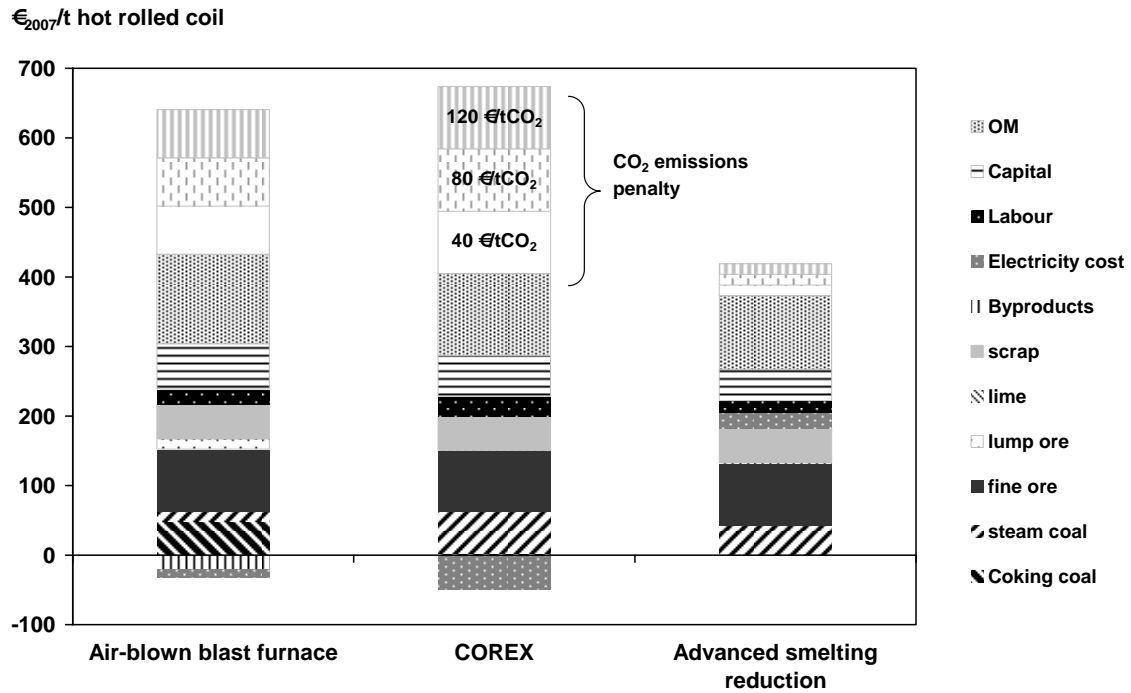
**Table 2-7:** Key parameters of CO<sub>2</sub> capture technologies for different steelmaking processes

Timeframe and iron making process	CO <sub>2</sub> capture technique		Specific energy consumption [GJ/tCO <sub>2</sub> captured]		Specific capital cost [€/tCO <sub>2</sub> captured /yr]	O&M cost
			Steam	Electricity		
<i>Short-mid term</i>						
Air-blown blast furnace	Chemical absorption	MEA	4.40	0.54	90	5 %
		KS-1	3.0	0.51	70	12 %
		Advanced solvent	2.5	0.50	70	5 %
		Physical absorption (Selexol)	-	0.77	180	5 %
		Shift + Physical absorption (Selexol)	0.62	0.91	190	5 %
Top gas recycling	Chemical absorption (MEA)		3.30	0.62	50	8 %
	Vacuum pressure swing adsorption		-	0.94	40	6 %
COREX	Chemical absorption (MEA)		4.40	0.45	70	5 %
	Physical absorption (Selexol)		-	0.97	40	5%
	Shift + physical absorption (Selexol)		0.63	0.60	110	5%
<i>Long term</i>						
Air-blown blast furnace	Selective carbon membrane		-	0.76	80	5%
Top gas recycling	Selective carbon membrane		-	0.88	90	4%
Top gas recycling	Hydrate crystallization		-	1.5	70	3%

### 2.4.1.3 Results

#### *Steel production costs and CO<sub>2</sub> emissions*

Figure 2-6 shows the breakdown of costs for the production of 1 tonne of hot rolled coil from BF-based process without CO<sub>2</sub> capture, the COREX process without CO<sub>2</sub> capture, and advanced smelting reduction process with integrated CO<sub>2</sub> capture. The prices of raw materials are given in Table 2-5. Note that these production cost figures are merely indicative. These three processes use different types and amounts of raw materials input and products. Therefore, steel production costs and their ranking by process depends largely on raw material costs and energy prices, which vary significantly from location to location. Based on our assumptions, both the COREX process and the advanced smelting reduction technology offers lower steel production cost compared to BF-based process. However, the COREX process is accountable for significantly more CO<sub>2</sub> than the BF-based process. Our results show that the COREX process appears more economical than the integrated steelmaking process until a CO<sub>2</sub> price of about 120 €/tonne is reached. The advanced smelting reduction process may not become considerably cheaper than the COREX process because of the large reduction in process gas production (some steam is generated instead), but by far more advantageous when the CO<sub>2</sub> emissions are taken into



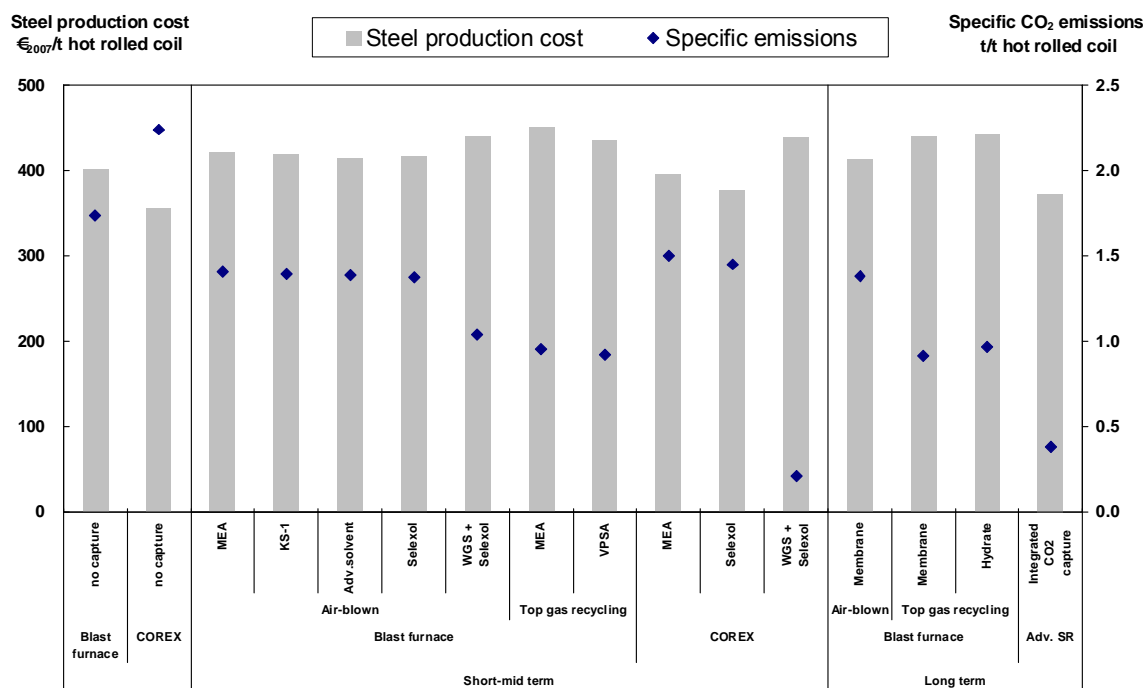
**Figure 2-6:** Production cost of 1t hot rolled coil from air-blown blast furnace (without CO<sub>2</sub> capture), the COREX process (without CO<sub>2</sub> capture), and the advanced smelting reduction process (integrated CO<sub>2</sub> capture). The CO<sub>2</sub> emissions penalty stacks represent the expected increase in steel production costs.

account. The results strongly suggest that the effect of CO<sub>2</sub> capture on the steel production process may be significant and CO<sub>2</sub> capture may affect the steel manufacturers' technology of choice. The results also indicate that with forecasted high energy and material prices, capital costs will only be a limited fraction of total production cost.

Figure 2-7 presents the production costs for one tonne of hot rolled coil from various steelmaking processes and their respective specific CO<sub>2</sub> emissions. The comparison shows that for the BF-based process, specific CO<sub>2</sub> emissions are nearly halved when the CO in the BF gas is shifted or the BF is converted to TGRBF. The figure also shows that the COREX process with CO<sub>2</sub> capture enables lower hot rolled coil production cost and lower specific CO<sub>2</sub> emissions compared to the reference BF-based process. However, the reduction in specific CO<sub>2</sub> emissions compared to the reference BF-based process will only be about 15%. Advanced smelting reduction process shows very promising results: reducing hot rolled coil production cost by 15% and specific CO<sub>2</sub> emissions by 90% compared to the reference BF-based process.

### CO<sub>2</sub> capture performance

Figure 2-8 shows CO<sub>2</sub> avoidance costs and CO<sub>2</sub> avoidance rates for various CO<sub>2</sub> capture routes for iron and steel production process in the ST/MT (10-15 years) and the LT (20



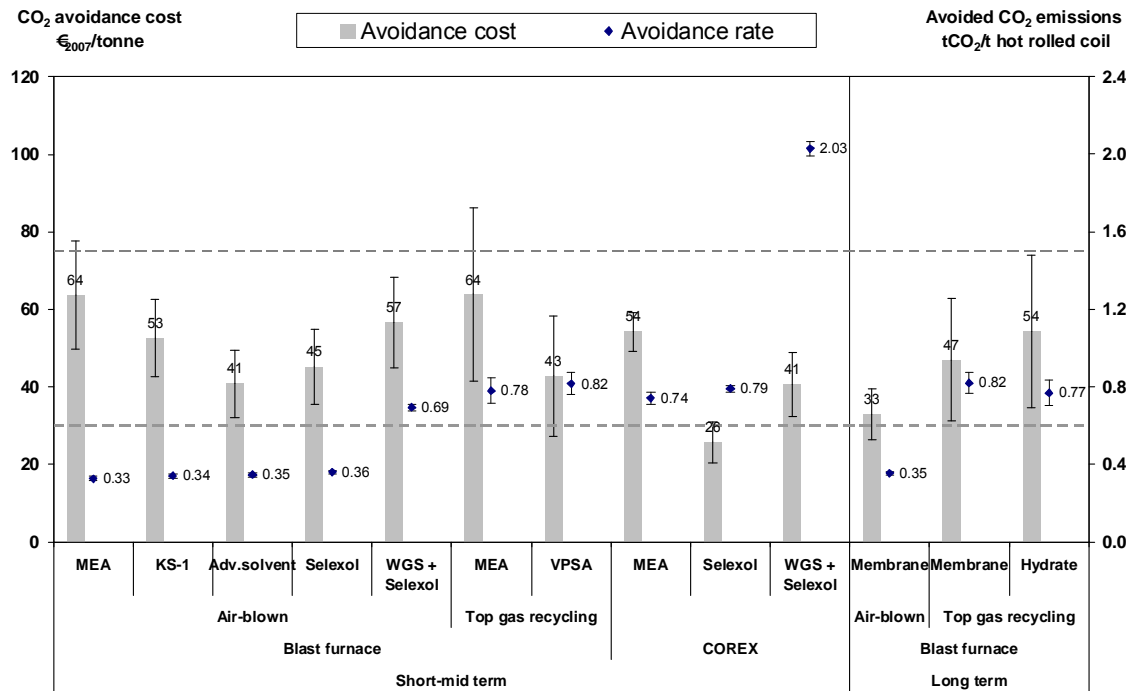
**Figure 2-7:** Production costs for one tonne of hot rolled coil from various steelmaking processes and the specific CO<sub>2</sub> emissions. CO<sub>2</sub> price is not considered in the calculations.

years or more). The error bars present the uncertainties derived from the ranges of the parameter values presented in Table 2-2.

In the ST/MT, CO<sub>2</sub> avoidance costs lie around 40-65 €/tCO<sub>2</sub> for BF-based processes. For both air-blown BF and TGRBF, CO<sub>2</sub> capture using MEA was found to be more expensive than most other capture technologies. For CO<sub>2</sub> capture from air-blown BF after water-gas shift (WGS), note that the benefits of generating H<sub>2</sub>-rich fuel are not considered in this study, which could reduce CO<sub>2</sub> avoidance costs considerably. Although not presented in the figure, CO<sub>2</sub> capture using physical absorption (Selexol) shows similar CO<sub>2</sub> avoidance rate and CO<sub>2</sub> avoidance cost. In comparison, CO<sub>2</sub> capture from TGRBF using MEA performs poorly due to its high thermal energy requirement for solvent regeneration.

With regard to the COREX process, CO<sub>2</sub> capture from the COREX process may be economically more attractive option for the iron and steel industry in the ST/MT, mainly due to the higher CO<sub>2</sub> concentration in the COREX gas. With WGS, the COREX process may achieve very high CO<sub>2</sub> avoidance rate of over 2 t/t hot rolled coil. This is because the COREX process has significantly larger coal input and the COREX gas has high CO concentration.

In the LT, CO<sub>2</sub> avoidance costs were found to decrease to about 30 €/tonne for air-blown BFs using selective carbon membrane. Membrane separation technique was found to be



**Figure 2-8:** CO<sub>2</sub> avoidance costs and avoidance rates for iron and steel production in the short-mid term future (10-15 years) and the long term future (20 years or more). The error bars present uncertainty ranges. The dotted lines indicate the probable future CO<sub>2</sub> price range (30-75 €/tonne).

considerably more economical when it is applied to air-blown BF gas than to TGRBF gas (Figure 2-8). For TGRBF, the economic advantage of selective carbon membrane technology over the conventional VPSA technology is minimal. Hydrate crystallization of CO<sub>2</sub> in the air-blown blast furnace gas suffers economically from high electricity requirement. The results indicate that advanced CO<sub>2</sub> separation technologies do not deliver significant improvement on the economic performance for TGRBF. In comparison with the probable future CO<sub>2</sub> price range (30-75 €/tonne), the results show that all CO<sub>2</sub> capture options for both the ST/MT and the LT can be economically feasible.

CO<sub>2</sub> avoidance costs for TGRBF cases show particularly large uncertainties because the modification from air-blown BF to TGRBF changes the inputs and outputs of energy products (coal, coke, steam, process gas and power) substantially. Moreover, the assumption on the grid electricity CO<sub>2</sub> emission factor influences not only CO<sub>2</sub> avoidance costs but also the CO<sub>2</sub> emissions reduction performance considerably.

## 2.4.2 Cement sector

### 2.4.2.1 Overview of CO<sub>2</sub> capture technologies

Key parameters of various CO<sub>2</sub> capture options for cement production plants obtained from the literature are presented in Table 2-8. The CO<sub>2</sub> capture technologies for the cement sector are compared on a plant scale of 1 Mt/yr clinker production at 8000 h/yr

operation time. We chose to standardize the plant scale on the basis of clinker production instead of cement production because most of the CO<sub>2</sub> emissions are attributable to clinker production. The capital cost figures are adjusted by applying a generic cost scaling factor (*SF*) of 0.67.

### *Post-combustion capture*

Literature seems to agree that post combustion CO<sub>2</sub> capture is the only option that can be implemented with a low technical risk and that enables retrofitting in the short-term (Hegerland et al., 2006; IEA GHG, 2008). However, a significant amount of heat will be required to regenerate CO<sub>2</sub> capture solvent. The major difference between centralized power plants and industrial plants such as cement plants is that the former have large quantities of low grade heat that can be used for solvent regeneration, whereas the latter generally do not (IEA GHG, 2008). Solvent regeneration heat can be either generated onsite or imported. When the regeneration heat is generated onsite, CO<sub>2</sub> generated from heat production can also be captured. In such cases, the fuel choice for the onsite heat generator has a large influence on the technical and economic performance of CO<sub>2</sub> capture (Hegerland et al., 2006). The pros and cons of natural gas and coal for a CHP plant are discussed in Hegerland et al. (2006). The economic performance of the post-combustion capture system with gas-fired CHP plant has been studied in Hegerland et al. (2006), while a system with a coal-fired CHP plant has been investigated in IEA GHG (2008). Besides the difference in the fuel used for the CHP plant, these two studies assume different de-NO<sub>x</sub> and de-SO<sub>2</sub> technologies. It is therefore difficult to perform a “like-for-like” comparison (IEA GHG, 2008).

The use of recovered waste heat for CO<sub>2</sub> capture may significantly improve the performance of chemical absorption CO<sub>2</sub> capture. Literature indicates some potential for waste heat recovery. Egberts et al. (2003) suggests that 30% of the post-combustion CO<sub>2</sub> capture heat requirement can be met by waste heat recovery. However, the amount of steam that can be recovered is likely to differ significantly by plant. One case study estimates a waste heat recovery potential of 1.3 GJ/t clinker from flue gas and clinker cooling (Hassan et al., 2007). Assuming a capture rate of 0.7 tCO<sub>2</sub>/t clinker and a specific heat requirement of 3.5GJ/tCO<sub>2</sub> captured, this equals to about half of the CO<sub>2</sub> capture heat requirement. Another case study estimates that 0.32 GJ/t clinker can be recovered from the kiln flue gas in the form of 8 bar steam, which equals to 13% of the CO<sub>2</sub> capture heat requirement, at a simple payback time of less than 1.5 years (Engin and Ari, 2005). It is not clear from the literature if CO<sub>2</sub> avoidance cost for post-combustion capture would differ between retrofit and new plants. McKinsey study (McKinsey & Company, 2009) assumes higher CO<sub>2</sub> avoidance costs for retrofit cases (about 10 €/tonne), while ECRA (ECRA, 2009a) suggests no difference between retrofit and new plants.



**Table 2-8:** Key parameters of various CO<sub>2</sub> capture options for cement production plants indicated in the literature. The cost figures are adjusted to a plant size of 1 Mt/yr clinker production.

CO <sub>2</sub> capture technique	System details		Retrofit possible?	Specific energy consumption [GJ/t CO <sub>2</sub> captured]			CO <sub>2</sub> capture rate [tCO <sub>2</sub> /t clinker]	Incremental capital cost [€/tCO <sub>2</sub> captured/yr]	References
				Primary energy	Steam	Electricity			
Post-combustion capture <sup>1)</sup>	MEA	Steam import	Yes	-	3.7-4.4	0.37-0.73	0.68-0.86	60-160	Hassan et al. (2007), Ho et al. (2010), IEA GHG (2008), RITE (2005)
	MEA	Onsite NG-CHP + CO <sub>2</sub> capture	Yes	3.6	-	0.13	0.77	160	Hegerland et al. (2006)
	MEA	Onsite coal CHP + CO <sub>2</sub> capture	Yes	5.5	-	-0.35	1.2	310	IEA GHG (2008)
	KS-1	Onsite coal CHP + CO <sub>2</sub> capture	Yes	4.0	-	-0.38	1.0	260	IEA GHG (2008)
	Adv. solvents	Steam import	Yes	0	2.7	0.54	0.81	80	RITE (2005)
Oxyfuel kiln	Precalciner only <sup>2)</sup>		Yes (major modification of precalciner)	0.06 (coal)	-	0.73	0.51 <sup>3)</sup>	150	IEA GHG (2008)
	Entire kiln		Preferred for new plants	0.86 (coal)	-	0.99	0.80 <sup>4)</sup>	110	ECRA (2009b)
Calcium looping (separate combustor and precalciner)	Precalciner only		Yes (major modification of precalciner)	1.6 (coal)	-	0.54	0.52	50 <sup>5)</sup>	Rodríguez et al. (2008)

<sup>1)</sup> 85% CO<sub>2</sub> capture efficiency is used for all studies except for (RITE, 2005), which assumes 90% capture efficiency.

<sup>2)</sup> Impurities such as SO<sub>2</sub> are assumed to be co-sequestered with CO<sub>2</sub>.

<sup>3)</sup> CO<sub>2</sub> capture efficiency was calculated to be 62%.

<sup>4)</sup> CO<sub>2</sub> capture efficiency was assumed to be 100% in the referenced study.

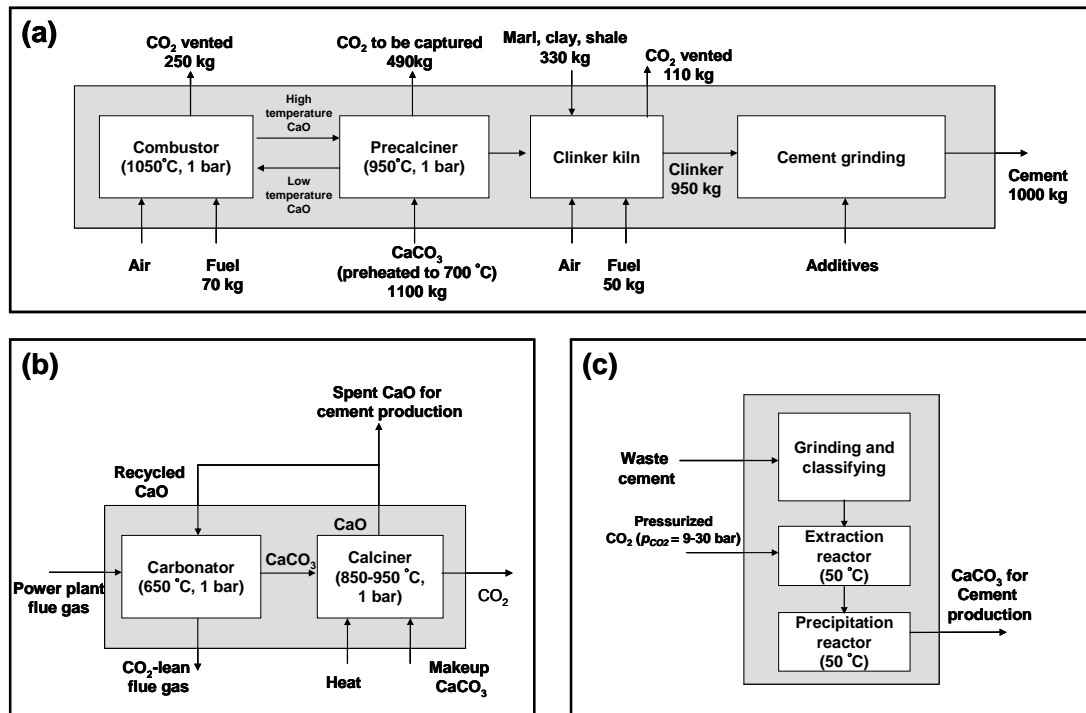
<sup>5)</sup> It is not clear which cost components are included.

### *Oxyfuel combustion with CO<sub>2</sub> capture*

Fuels can be combusted with pure oxygen mixed with recycled CO<sub>2</sub> instead of air to heat kiln furnaces. Oxyfuel combustion of only precalciner is suggested to have the lowest technical risk, avoiding the difficulties of undertaking oxyfuel combustion in the kiln and of minimizing air in-leakage in the kiln and raw mill while achieving high CO<sub>2</sub> capture rate (IEA GHG, 2008). Oxyfuel CO<sub>2</sub> capture from the entire cement plant is likely to be available for new plants in the longer term. Oxyfuel combustion can be retrofitted to existing kilns, but it requires most of the core units in the cement plant to be rebuilt (IEA GHG, 2008). Nevertheless, the economic advantage of oxyfuel CO<sub>2</sub> capture over post-combustion capture may be significant enough to carry out a major rebuild of the kiln (i.e., from single to twin preheater towers) and undertake the oxyfuel combustion retrofit rather than implementing post-combustion capture (IEA GHG, 2008). Oxyfuel kiln technology is currently in the process modeling and laboratory phase, and thus the precalciner operation in a CO<sub>2</sub>-rich atmosphere is not proven to date (IEA GHG, 2008). However, the effect of the increased CO<sub>2</sub> partial pressure on the calcination of limestone has been investigated and it suggests that a stable temperature profile between the precalciner and the kiln becomes very important (Zeman, 2009). The temperature at which the calcination reaction takes place rises from 670 °C at  $p_{\text{CO}_2} = 0$  to 900 °C at 1 bar and the reaction type also changes from a gradual onset to threshold reaction. Calcium oxide promptly returns to calcium carbonate through the carbonation reaction if the temperature drops below the threshold (Zeman, 2009). As with post-combustion capture, it is not clear from the literature if CO<sub>2</sub> avoidance cost for oxyfuel CO<sub>2</sub> capture would differ between retrofit and new plants.

### *Other advanced CO<sub>2</sub> capture technologies*

Another advanced CO<sub>2</sub> capture technique that can be applied to cement production is to separate combustion chamber of the precalciner (Figure 2-9(a)), which uses calcium oxide (CaO) as a heat carrier (Rodríguez et al., 2008). The proposed process separates the fuel combustion and calcination process into two chambers within the precalciner unit. Some CaO generated in the calcination chamber flows through the fuel combustion chamber and gets heated up to around 1000 °C. Then the heated CaO mixes with CaCO<sub>3</sub> in the calcination chamber, heats up the CaCO<sub>3</sub>, and drives the calcination reaction. The CO<sub>2</sub> capture rate is lower than that for post-combustion capture because CO<sub>2</sub> is captured only from the calcination chamber. The literature does not indicate whether the technology can be retrofitted. We reckon that retrofit is possible, but it will require rebuilding of the precalciner. Note that there is only one publicly available publication (Rodríguez et al., 2008) that assessed the techno-economic performance of CaO-looping CO<sub>2</sub> capture from cement production. Further research is therefore needed to gain insights into the techno-economic potential of this option.



**Figure 2-9:** (a) CO<sub>2</sub> capture by separating the combustion chamber from precalciner (Rodríguez et al., 2008), (b) cement production from spent CaO sorbent used for CO<sub>2</sub> capture from power plants (Abanades et al., 2005), and (c) carbonation of waste cement using pressurized CO<sub>2</sub> (Katsuyama et al., 2005).

### CO<sub>2</sub> emissions reduction by indirect CO<sub>2</sub> capture

Another possibility is the integration of cement production with power plants equipped with calcium looping CO<sub>2</sub> capture (e.g., Abanades et al. (2005), Figure 2-9(b)). With calcium looping, CO<sub>2</sub> contained in the combustion gas is absorbed by CaO to form CaCO<sub>3</sub> in the carbonation chamber at high temperatures. The generated CaCO<sub>3</sub> is then sent to the calcination chamber where CaO for subsequent carbonation cycles is regenerated and a gas stream rich in CO<sub>2</sub> is produced (Bosoaga et al., 2008). Spent sorbent from calcium-looping CO<sub>2</sub> capture, which is CaO that is sintered after numerous regeneration cycles and lost sorbent reactivity, can directly be used for cement production. Spent sorbent can be directly put into the cement mill for cement manufacture. The use of spent sorbent in the cement industry leads to reduction in extraction of CaCO<sub>3</sub> and consequent CO<sub>2</sub> emissions (Abanades et al., 2005). Blamey et al. (2010) estimated that if all coal-fired power plants in the UK and the USA are equipped with calcium looping CO<sub>2</sub> capture, the amount of spent CaO sorbent sums up to about three times their respective national cement production levels<sup>10</sup>.

<sup>10</sup> In the referenced study, the number of CaO loops is assumed to be about 10. A CaCO<sub>3</sub> makeup feed rate of 22.5kg/s is assumed for a 500MWe power station. This means that spent CaO generation rate is  $22.5 \times 56$  (molar mass CaO)/100 (molar mass CaCO<sub>3</sub>) = 12.6 kg/s. Assuming an electrical conversion efficiency of 40%(LHV) and a CO<sub>2</sub> emission factor of 95 kgCO<sub>2</sub>/GJ LHV for coal, CO<sub>2</sub> generation rate is about 120 kgCO<sub>2</sub>/s. Therefore, the number of CaO loops is  $120/12.6 = 9.5$ .

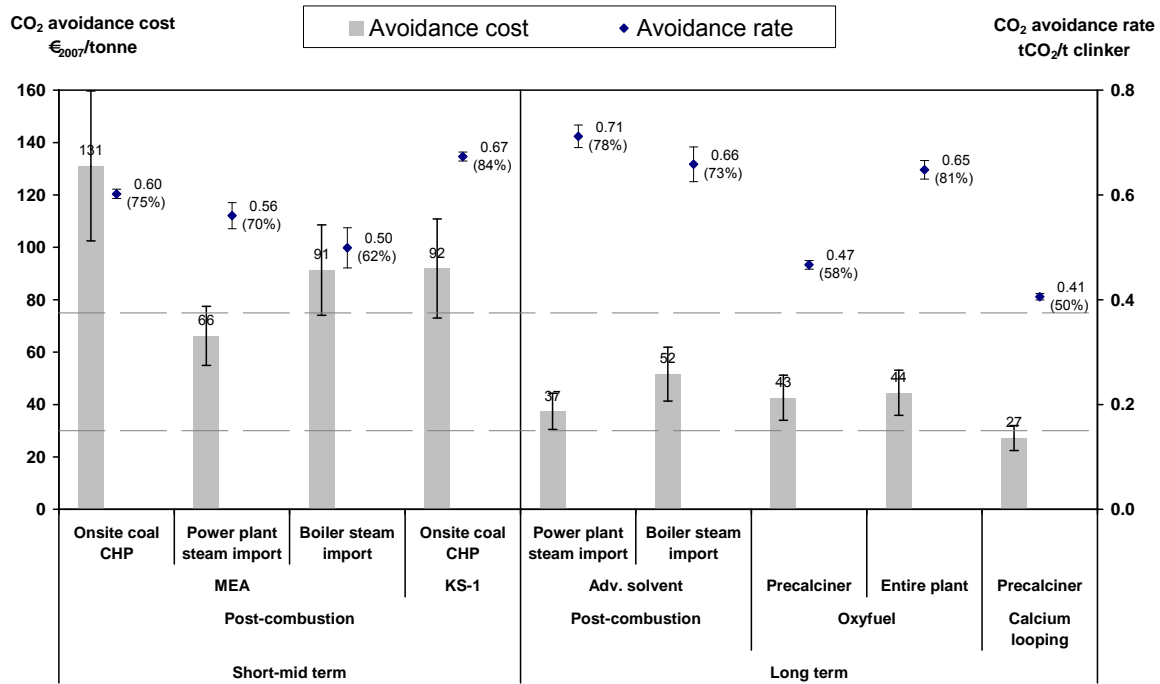
Another study (Katsuyama et al., 2005) proposed a process to produce CaCO<sub>3</sub> through precipitation process from waste cement by using pressurized CO<sub>2</sub> (Figure 2-9(c)), leading to a reduced natural limestone consumption consequent CO<sub>2</sub> emissions. The proposed process also reduces sulphur emissions contained in the flue gas. The produced CaCO<sub>3</sub> is of high purity that it can be used as heavy calcium carbonate or desulphurization agent. The economic analysis indicates that the production cost of CaCO<sub>3</sub> from the proposed process is between 136-323 \$<sub>2007</sub>/tonne, which is significantly higher than the typical price for limestone (e.g., the IEA GHG study (2008) assumes 3 €/tonne) but comparable to that for heavy calcium carbonate or desulphurization agent produced from natural limestone. This CO<sub>2</sub> emissions reduction option can be implemented without CO<sub>2</sub> transport and storage infrastructure. Although being perhaps attractive options, they are not investigated further in this paper as they are beyond our research scope.

#### 2.4.2.2 Calculation assumptions

Table 2-9 shows key performance parameters of the CO<sub>2</sub> capture technologies for the cement sector by timeframe. Cement production costs are not calculated because literature reviewed in this study suggest that the raw materials inputs will not be affected by the installation of CO<sub>2</sub> capture (IEA GHG, 2008; Rodríguez et al., 2008).

**Table 2-9:** Key parameters of various CO<sub>2</sub> capture options for cement production plants used for the assessment. The cost figures are adjusted to a plant size of 1 Mt clinker production per year.

CO <sub>2</sub> capture technique		Specific energy consumption [GJ/t CO <sub>2</sub> captured]			CO <sub>2</sub> capture rate [tCO <sub>2</sub> /t clinker]	Incremental capital cost [€/tCO <sub>2</sub> captured//yr]	O&M cost
		Primary energy	Steam	Electricity			
<i>Short-mid term</i>							
Post-combustion	MEA (steam import)	-	3.7	0.73	0.68	160	12%
	MEA (onsite coal CHP + CO <sub>2</sub> capture)	5.5	-	-0.35	1.17	310	7%
	KS-1 (onsite coal CHP + CO <sub>2</sub> capture)	4.0	-	-0.38	1.0	260	9%
<i>Long term</i>							
Post combustion	Adv. solvent (steam import)	-	2.7	0.54	0.81	80	8%
Oxyfuel kiln	Precalciner only	0.06 (coal)	-	0.73	0.51	150	6%
	Entire plant	0.86 (coal)	-	0.99	0.80	110	5%
Calcium looping		1.6 (coal)	-	0.54	0.52	50	5%



**Figure 2-10:** CO<sub>2</sub> avoidance costs and avoidance rates for cement production in the short-mid term future (10-15 years) and the long term future (20 years or more). The error bars present uncertainty ranges. The dotted lines indicate the probable future CO<sub>2</sub> price range (30-75 €/tonne). The CO<sub>2</sub> avoidance rate values are expressed in tCO<sub>2</sub>/t clinker. Note that the relative CO<sub>2</sub> emission avoidance rates compared to reference emissions are reported in brackets, and they do not follow any of the axes in the figure. The two CO<sub>2</sub> avoidance rate indicators are not always proportional because the reference emissions differ between technologies.

### 2.4.2.3 Results

Figure 2-10 shows CO<sub>2</sub> avoidance costs and avoidance rates for various CO<sub>2</sub> capture routes for cement production process in the ST/MT (10-15 years) and the LT (20 years or more).

#### Short-mid term (ST/MT)

The comparison of results for the cases with onsite coal-CHP and combined post-combustion CO<sub>2</sub> capture indicates the most advanced CO<sub>2</sub> capture solvents available today (KS-1) can only lead to about 30% reduction of overall costs of CO<sub>2</sub> capture compared to MEA. Nevertheless, a major CO<sub>2</sub> capture cost reduction may be achieved by purchasing low-cost steam. With low cost steam import, CO<sub>2</sub> avoidance cost may be lowered to about 65 €/tonne for using MEA, indicating that CO<sub>2</sub> avoidance cost around 50 €/tonne may be achieved by using state-of-the-art solvents such as KS-1. In comparison with the probable future CO<sub>2</sub> price range (30-75 €/tonne), the results show that post-combustion capture using MEA is unlikely to become economically feasible unless low cost steam is available.

### *Long term future*

The results shown in Figure 2-10 suggest that the CO<sub>2</sub> avoidance costs may decrease from above 60 €/tonne in the ST/MT down to 25-55 €/tonne in the long term future. The CO<sub>2</sub> avoidance rate is found to vary largely, depending on whether the CO<sub>2</sub> from processes other than the precalciner would be captured or not. Advanced solvents for post-combustion capture may compete with oxyfuel combustion capture technology both in terms of CO<sub>2</sub> avoidance rate and CO<sub>2</sub> avoidance cost.

### *Uncertainty of the results*

The calculated uncertainties of CO<sub>2</sub> avoidance costs are smaller than those observed for the iron and steel sector, which is partly attributable to the fact that CO<sub>2</sub> capture in the cement sector does not affect the inputs and outputs of valuable byproducts and raw and intermediate materials as is the case in the iron and steel sector. The figure also shows that the CO<sub>2</sub> avoidance cost for the post-combustion capture case integrated with onsite coal-fired CHP plant is significantly more expensive than other options regardless of the uncertainty in parameter values. Figure 2-10 shows that the grid electricity CO<sub>2</sub> emission factor and the steam-electricity conversion factor ( $f_{St,Cap}$ ) has a limited influence on the CO<sub>2</sub> avoidance performance.

## **2.4.3 Petroleum refineries and petrochemicals**

This section reviews the CO<sub>2</sub> capture technologies from furnaces and heaters for the petroleum refining and petrochemical sectors proposed in the literature.

### **2.4.3.1 Overview of CO<sub>2</sub> capture technologies**

Table 2-10 presents the technical and economic performance data of various CO<sub>2</sub> capture technologies proposed for the petroleum refineries, chemicals and petrochemical sectors. The capital costs for combined stacks and catalytic crackers are standardized to 2 MtCO<sub>2</sub>/yr and 1 MtCO<sub>2</sub>/yr, respectively. The capital cost figures are adjusted by applying a cost scaling factor ( $SF$ ) of 0.7.

### *Post-combustion capture*

Because the CO<sub>2</sub> is at a low partial pressure, post-combustion capture using chemical absorption is considered to be the only feasible option in the short term for both steam cracker furnace gas and refinery stack gases. The CCS project in Mongstad (Norway) plans to capture CO<sub>2</sub> in the flue gases from residue catalytic cracker and the CHP that supplies heat and power to the refinery using MEA (Statoil, 2009). At full scale the project aims to capture around 2 Mt/yr CO<sub>2</sub>. The uncertainty of the economic performance of this project is very high because of the extensive use of new technology and the fact that industry has no relevant experience with this type of project (Statoil, 2009), and in 2010 the investment decision has been postponed until 2014 (Norwegian Ministry of Petroleum and Energy, 2010). The most advanced, commercially available CO<sub>2</sub> capture processes, which are all solvent-based, will not reduce overall costs of CO<sub>2</sub> capture by more than around 25% compared to a conventional MEA solvent (van Straelen et al., 2010).

**Table 2-10:** Energy, CO<sub>2</sub> and economic performance data for petroleum refining and petrochemical sectors with CO<sub>2</sub> capture. The reference refinery stacks are assumed to emit 2 Mt/yr CO<sub>2</sub> and catalytic crackers are assumed to emit 1 Mt/yr CO<sub>2</sub>.

Emission source	CO <sub>2</sub> capture technique		CO <sub>2</sub> capture rate [t/t reference emissions]	Specific energy consumption			Incremental capital cost [€/tCO <sub>2</sub> captured/yr]	Reference
				[GJ/tCO <sub>2</sub> captured]				
				Primary energy (natural gas)	Steam	Power		
Combined stack	Post-combustion (MEA)	Onsite NG-CHP	0.85	5.7	-	-0.65	250	Hurst and Walker (2005)
		Steam import	0.9-0.99	-	3.3-4.4	0.48-0.62	100-210	Farla et al. (1995), Ho et al. (2010), IEA GHG (2000a)
	Oxyfuel (cryogenic)	Onsite power <sup>1)</sup>	0.78-1.07	1.9-2.5	-	-0.24- -0.06	200-240	Allam et al. (2005a)
		Power import	0.96	-	-	2.0	150	IEA GHG (2000a)
	Oxyfuel (OCM)	Onsite power <sup>1)</sup>	0.50-1.21	5.3-16.4	-	-8.5 - -1.5	170-390	Allam et al. (2005b)
Pre-combustion	WGSMR	0.77	4.8	-	-0.1	150	Middleton et al. (2005)	
Catalytic cracker	Post-combustion (MEA)	Steam import	0.94	-	4.6	0.6	150	de Mello et al. (2009)
	Oxyfuel	Cryogenic	0.94-1.04	-	-3.1- -3.3	2.4-2.5	190-200	de Mello et al. (2009)

1) The value ranges are large because three different system configurations can be considered. For case 1, the onsite power plant is a NGCC, and no steam is exported. For case 2, the onsite power plant is a NG-CHP, and the cogenerated steam replaces that from existing boilers. For case 3, autothermal reforming is applied for the CHP plant (Allam et al., 2005a; Allam et al., 2005b).

### *Oxyfuel capture*

It is expected that in the mid-term future, oxyfuel CO<sub>2</sub> capture may compete with or outperform post-combustion capture as a retrofit option with an additional advantage of significantly lower SO<sub>2</sub> and NO<sub>x</sub> emissions (IEA, 2008a). There are a number of pilot oxyfuel boiler plants up to 30 MW<sub>th</sub> scale (e.g. Schwarze Pumpe in Germany) being tested to date (IEA, 2008a). A techno-economic assessment of oxyfuel CO<sub>2</sub> capture from refinery boilers and heaters using cryogenic distillation ASU has been performed for an existing refinery in the UK (Allam et al., 2005a). About 2MtCO<sub>2</sub> per year is to be captured collectively from a number of furnaces and heaters of 10-110 MW firing duty. Because of large electricity requirement for oxygen generation, CO<sub>2</sub> capture and compression, the CO<sub>2</sub> capture system also included an onsite gas-fired power plant, either combined cycle or CHP. In case of CHP, the cogenerated steam replaces part of the boiler steam, which saves oxygen flow to the boilers. The feasibility study on the oxyfiring FCC also indicates significantly lower CO<sub>2</sub> capture costs compared to post-combustion capture (de Mello et al., 2009). The pilot plant test results also indicate that stable operation of an oxyfuel FCC without significant impact of catalyst regeneration and activity is technically possible (de Mello et al., 2009). With regard to retrofitting furnaces and heaters for oxyfuel combustion, it has been suggested that it differs from the case for boilers in a number of ways (Wilkinson et al., 2003). Compared to boilers, heaters have: (1) a greater air in-leakage due to wider range of furnace designs and different techniques being used for their construction, (2) hydrocarbons present in the furnace tubes, and (3) no automatic control of air/fuel ratio and draught. The conversion of heaters to oxyfuel operation may, therefore, encounter more technical issues than boilers do. A pilot test of oxyfuel CO<sub>2</sub> capture from fluid catalytic crackers has been performed in Brazil that confirmed no significant changes from normal operations in product profile, stability of operation and the effectiveness of coke burn (de Mello et al., 2009).

Advanced oxyfuel boilers and heaters integrated with oxygen conducting membrane (OCM) can be commercially available in the longer term. With OCM, oxygen ions are transferred under a gradient of oxygen partial pressure on the opposite side of the membrane at high temperatures typically around 800-900 °C. OCM can produce oxygen of very high purity, above 99% (Smith and Klosek, 2001). Depending on applications, membrane air separation is expected to reduce capital cost by 35-48% and specific power consumption by 35-68% compared to cryogenic separation (Armstrong and Fogash, 2007). The OCM technology is near commercialization. Air Products has been operating demonstration projects at 0.5 tO<sub>2</sub>/d scale since 2007 and is building a 5 tO<sub>2</sub>/d plant (Armstrong and Fogash, 2007).

When OCM is used, the CO<sub>2</sub> capture system configurations differ significantly from those using cryogenic distillation ASU. With cryogenic distillation, the ASU and the power plant are independent of each other. When OCM is used, the ASU and the power plant can be integrated into one unit: the compressed hot air that comes out of OCM can be expanded in a gas turbine for power generation with some supplementary gas firing.



**Table 2-11:** Key performance data for petroleum refineries and petrochemical industry with CO<sub>2</sub> capture.

CO <sub>2</sub> capture technique	Process	CO <sub>2</sub> capture rate [t/t reference emissions]	Specific energy consumption [GJ/tCO <sub>2</sub> captured]			Incremental capital cost [€/tCO <sub>2</sub> captured/yr]	O&M cost
			Primary energy (NG)	Steam	Power		
<i>Short-mid term</i>							
Post-combustion	Combined stack	0.85	5.7	-	-0.1	190	8%
	Catalytic cracker	0.94	-	4.6	0.6	150	12%
Oxyfuel (cryogenic)	Combined stack	0.87	2.5	-	-0.2	240	2%
	Catalytic cracker	0.94	-	-3.3	2.5	200	4%
<i>Long term</i>							
Oxyfuel (OCM)	Combined stack	0.50	12.5	-	-8.4	390	6%
Pre-combustion (WGSMR)	Combined stack	0.77	4.8	-	-0.1	150	5%

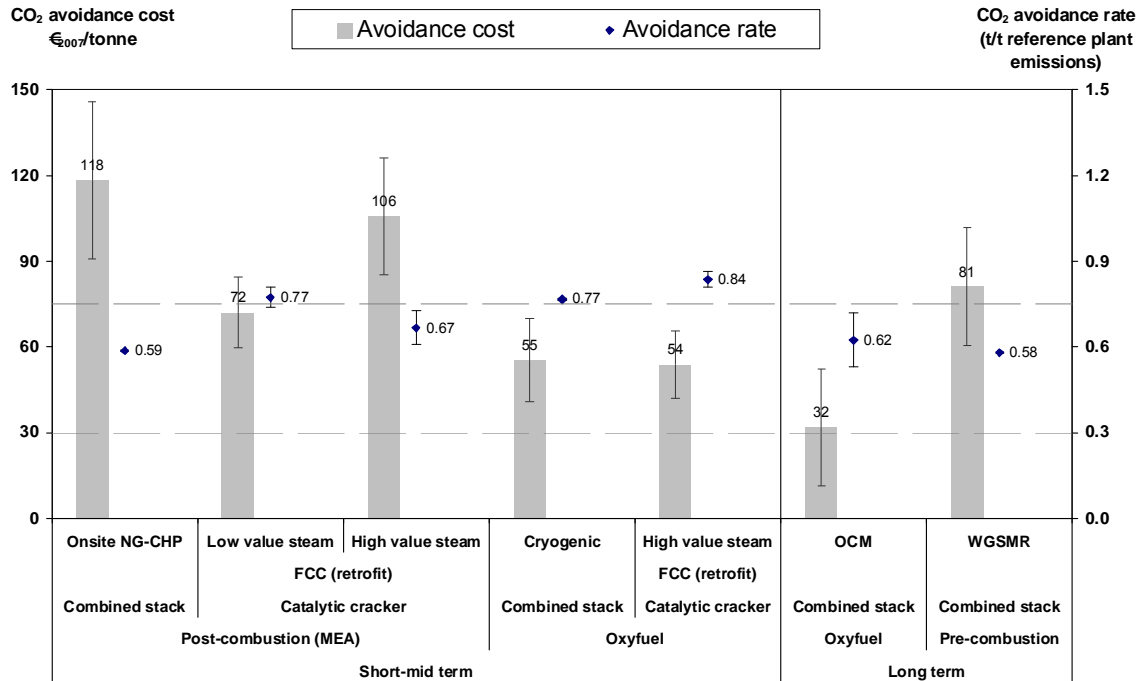
Because the power plant is scaled on the amount of air used for OCM, the power generation was found to be much larger than for the case using cryogenic distillation ASU (Allam et al., 2005b). The power plant can be operated in CHP mode and replace part of the boiler steam to save oxygen flow to the boilers. Allam et al. (2005b) has shown that it is more economical to use the steam from the oxyfuel gas turbine power plant for industrial processes and replace existing boilers than expanding the steam in the turbine and generating additional electricity.

#### *Pre-combustion capture*

Pre-combustion CO<sub>2</sub> capture is also likely to become technically feasible in the ST/MT. Unlike oxyfuel combustion, no modification of steam boilers and fired furnaces is necessary because they can combust nearly pure hydrogen (Thomas and Kerr, 2005), and hydrogen production from various hydrocarbons have been commercially practiced. A potentially promising pre-combustion capture technology for the mid-term is the SEWGS process (see section 2.4.1). However, there is no study that assessed the performance of the SEWGS process for refinery gases. With regard to the LT technologies, the water-gas shift membrane reactor (WGSMR) has been extensively researched in the CO<sub>2</sub> Capture Project (CCP, 2005). The WGSMR is located downstream of a conventional autothermal reformer, where fuel gas is converted to H<sub>2</sub>, CO and CO<sub>2</sub> by injecting steam and oxygen. In the WGSMR, the WGS reaction and H<sub>2</sub> separation takes place simultaneously (Middleton et al., 2005).

#### **2.4.3.2 Calculation assumptions**

Table 2-11 shows the key performance parameters of the CO<sub>2</sub> capture technologies for the refining and petrochemical sectors by timeframe.



**Figure 2-11:** CO<sub>2</sub> avoidance cost and avoidance rates for CO<sub>2</sub> capture from petroleum refining and petrochemical sectors in the short-mid term future (10-15 years) and the long term future (20 years or more). The error bars present uncertainty ranges. The dotted lines indicate the probable future CO<sub>2</sub> price range (30-75 €/tonne).

### 2.4.3.3 Results

Figure 2-11 shows CO<sub>2</sub> avoidance costs and CO<sub>2</sub> avoidance rates for various CO<sub>2</sub> capture routes for chemical and petrochemical sectors and petroleum refineries in the ST/MT (10-15 years) and the LT (20 years or more). The error bars present the uncertainties derived from the ranges of the parameter values presented in Table 2-2. CO<sub>2</sub> avoidance costs for post-combustion capture in the ST/MT are found to be above 100 €/tonne unless low-cost steam is available in the vicinity. The results are in line with those from van Straelen et al. (2010), who estimated a CO<sub>2</sub> avoidance cost around 90 €/tonne at 7% discount rate. In comparison with the probable future CO<sub>2</sub> price range (30-75 €/tonne), post-combustion capture using MEA will unlikely become an economically feasible technology unless low cost steam is available. Note that first-of-a-kind plants would be significantly more expensive: CO<sub>2</sub> avoidance cost calculated for the Mongstad demonstration plant (Statoil, 2009) was above 200 €/tonne. Oxyfuel capture, in contrast, may provide relatively low CO<sub>2</sub> avoidance costs around 50-60 €/tonne.

In the LT, advanced oxyfuel CO<sub>2</sub> capture using OCM may reduce CO<sub>2</sub> avoidance cost down to 30 €/tonne. CO<sub>2</sub> avoidance cost for the combined refinery heaters and boilers is sensitive to the price and CO<sub>2</sub> emission factor of grid electricity because of the excess electricity production is significant, amount up to nearly 300 MW. On the other hand, CO<sub>2</sub>

avoidance cost can be similar to the oxyfuel capture using conventional ASU under unfavorable conditions.

In comparison with the probable future CO<sub>2</sub> price range (30-75 €/tonne), pre-combustion capture does not seem economically promising. However, this study assumed that the generated H<sub>2</sub>-rich gas is simply combusted, as the process gas is used in the case without CO<sub>2</sub> capture. Note that there may be a significant economical benefit when the H<sub>2</sub>-rich gas generated as a result of CO<sub>2</sub> capture replaces the onsite H<sub>2</sub> production.

With regard to uncertainty in CO<sub>2</sub> avoidance rates, CO<sub>2</sub> capture options with large electricity import or export, i.e. catalytic crackers with post-combustion capture and combined stacks with advanced oxyfuel capture, show relatively large uncertainties.

#### **2.4.4 Summary of the results and global technical potentials for CO<sub>2</sub> emissions reduction**

Table 2-12 summarizes the technical and economic performance results for industrial processes investigated in this study. Compared to the cost figures presented in the literature (see Table 2-1), the results obtained in this study are relatively high. The exceptions are advanced industrial process technologies with integrated CO<sub>2</sub> capture for the LT.

We also roughly estimated the technical potential for worldwide CO<sub>2</sub> emissions reduction in 2030 for the industrial processes investigated with ST/MT technologies. The worldwide crude steel production from BF-based processes in 2030 is projected to be about 1.2 Gt/yr (Oda et al., 2007). Assuming a CO<sub>2</sub> avoidance rate of 0.8 t/t crude steel, the technical potential for CO<sub>2</sub> emissions reduction due to CO<sub>2</sub> capture can be up to about 1Gt/yr. The global cement production in 2030 is expected to be about 3500 Mt/yr (IEA, 2009). If all BF-based cement plants worldwide are equipped with CO<sub>2</sub> capture and assuming a clinker/cement ratio of 0.9, the CO<sub>2</sub> emissions reduction sums up to about 2 Gt/yr. For petroleum refineries, McKinsey & Company estimated that the CO<sub>2</sub> emissions from downstream refining in 2030 will be around 1.5 Gt/yr (McKinsey & Company, 2009). Assuming that furnaces/boilers and catalytic crackers account for 65% and 16% of total refinery CO<sub>2</sub> emissions, respectively (IEA GHG, 1999b), the CO<sub>2</sub> reduction potentials worldwide would be up to 1 Gt/yr for furnaces and boilers, and about 0.2 Gt/yr for catalytic crackers. The technical potential for CO<sub>2</sub> reduction from the industrial processes investigated altogether sums up to about 4 Gt/yr.

**Table 2-12:** Summary of technical and economic performance of CO<sub>2</sub> capture in the industrial processes investigated in this study.

Sector	Process	Reference plant scale (annual production)	CO <sub>2</sub> avoidance performance			
			Short-mid term		Long term	
			Avoidance cost [€/tonne]	Avoidance rate [t/unit product]	Avoidance cost [€/tonne]	Avoidance rate [t/unit product]
Iron and steel	Integrated steel-making	4 Mt hot rolled coil	40-65	0.3-0.8 tCO <sub>2</sub> /t hot rolled coil	30-55	0.35-0.8 tCO <sub>2</sub> /t hot rolled coil
	Smelting reduction		25-55	0.7-2 tCO <sub>2</sub> /t hot rolled coil	< 0 (advanced smelting reduction, versus BF)	~1.3 tCO <sub>2</sub> /t hot rolled coil (versus BF)
Cement		1 Mt clinker	65-135	0.5-0.7 tCO <sub>2</sub> /t clinker	25-55	0.4-0.7 tCO <sub>2</sub> /t clinker
Petroleum refineries and petrochemicals	Combined stacks	2 MtCO <sub>2</sub> reference plant emission	50-60 (oxyfuel) 70-120 (post combustion, depend largely on heat supply options)	~80% (oxyfuel) 60-80% (post combustion)	~30 (oxyfuel)	~80% reference plant emission
	Catalytic crackers	1 MtCO <sub>2</sub> reference plant emission				

Note that the actual deployment level in 2030 will likely be much smaller. The IEA's CCS roadmap (IEA, 2008a) estimates that actual deployment for the industry and petroleum refineries in the BLUE Map scenario would be around 0.8 Gt/yr altogether: 0.2 Gt/yr for the iron and steel sector, 0.15 Gt/yr for the cement sector, 0.1 Gt/yr from the petrochemical sector (from ammonia production) and 0.3 Gt/yr for the fuel transformation sector.

## 2.5 Discussion

### 2.5.1 Limitations of this study

Four key limitations have been identified. Firstly, this study only standardized key technical and economic parameters that underlie the studies reviewed in this paper. Other parameters may also have a substantial impact on the results. These parameters include: fuel and raw material specifications, product specifications, and plant location. The indicated specifications affect the the process gas from which CO<sub>2</sub> will be captured. With regard to plant location, a study on retrofitting post-combustion CO<sub>2</sub> capture to a refinery in Alaska has shown that a harsh climate and remote location lead to higher costs because: (1) only 2-3 months per year are available for construction, (2) need for modularizing and

pre-fabricating the equipment to minimize onsite construction activities, (3) need for a prolonged schedule due to limited sea-lift opportunities, (4) use of expensive cooling system, and (5) lack of fresh water (Hurst and Walker, 2005). This Alaska case may be an extreme one, but it stresses that the effect of plant location should not be underestimated.

Another issue is the selection of the reference systems for each CO<sub>2</sub> capture option. Damen et al. (2007) pointed out for the case of power plants that an integrated gasification combined cycle (IGCC) power plant equipped with CO<sub>2</sub> capture does not necessarily replace an IGCC power plant without CO<sub>2</sub> capture. The same can be said for some industrial processes. For iron and steel industry, for example, a COREX plant with CO<sub>2</sub> capture does not necessarily replace a COREX plant without CO<sub>2</sub> capture, but instead it may replace a blast furnace based integrated steelmaking plant without CO<sub>2</sub> capture. This may also be the case for cement plants, particularly for oxyfuel kiln and CaO looping technologies where the clinker production process environment changes considerably. The effects of CO<sub>2</sub> capture other than the add-on chemical absorption technique on the production process (raw material input rates, feedstock quality, production rate, product quality, etc) have been neglected in the literature reviewed in this study (IEA GHG, 2008; Rodríguez et al., 2008), but such effects should not be underestimated.

The third issue concerns the quality of data collected from the literature. During the literature survey, we experienced difficulties identifying the definition of “capital cost” in a large number of studies. For these studies we could not carry out the standardization procedure. Moreover, many studies do not present equipment cost data. Thus it was in some cases not possible to compare and evaluate whether the capital costs reported in the literature are within the reasonable range. This highlights the need for reporting more transparent data and calculations in the publicly available literature. Moreover, the number of techno-economic studies on CO<sub>2</sub> capture from industrial sources was limited. For the industry there was only one techno-economic study for most of the CO<sub>2</sub> capture routes.

The fourth issue is that this study did not look into the economic consequence of the production loss due to retrofitting CO<sub>2</sub> capture. Our study confirms that for many industrial processes, oxyfuel CO<sub>2</sub> capture will likely be more economical than post-combustion capture. The IEA GHG study on CO<sub>2</sub> capture from the cement sector (IEA GHG, 2008) suggests the retrofit of oxyfuel CO<sub>2</sub> capture over post-combustion capture even if a major modification of the process is required. *However, it has been indicated that for example, the retrofit construction for oxyfuel CO<sub>2</sub> capture from heaters and furnaces could be more complicated than that for boilers.* To the authors’ knowledge no study has examined in detail the significance of consequent production and economic losses.

Lastly, the uncertainty on the cost performance results obtained in this study may be larger than quantified by our analysis. In this study we assessed aggregate uncertainty of CO<sub>2</sub> capture performance regarding the following parameters: coal and gas prices, grid electricity CO<sub>2</sub> emission factor, and annualized capital cost. As stated in the introduction,

this study considered CO<sub>2</sub> capture plant installations in the industrialized world. The results obtained in this study may not stand for CO<sub>2</sub> capture installed in developing countries. The IEA GHG study on CO<sub>2</sub> capture from cement plants suggest that the cement production cost as well as CO<sub>2</sub> mitigation costs can be 40-50% lower in Asian developing countries than in the industrialized world due to the economy of scale of operating a larger plant, the lower equipment costs and cheaper labor (IEA GHG, 2008). The potentially cheaper CO<sub>2</sub> capture may also accelerate the relocation of industrial plants from industrialized countries to developing countries, which is already observed today. Moreover, even within the industrialized world, parameters such as energy and material prices, grid electricity CO<sub>2</sub> emission factor and interest rates differ significantly from country to country. In particular, price and CO<sub>2</sub> emission factor of grid electricity may be affected significantly by the deployment of CCS. Therefore, the CO<sub>2</sub> capture performance trends observed in our results do not necessarily hold for all industrialized world.

### 2.5.2 General observations

This study focused on the techno-economic performance of CO<sub>2</sub> capture from industrial processes as independent CO<sub>2</sub> emission sources, but industries may achieve significant CO<sub>2</sub> reductions by integrating with other processes. There seems to be a number of opportunities for cement manufacture that reduces limestone consumption by capturing CO<sub>2</sub> elsewhere, e.g. power plants (the use of spent sorbent from calcium-looping CO<sub>2</sub> capture (e.g., Abanades (2005)) and the use of waste cement and pressurized CO<sub>2</sub> captured (Katsuyama et al., 2005)). One important advantage of this approach is that CO<sub>2</sub> can be reduced from cement plants without waiting for the CO<sub>2</sub> transport and storage infrastructure to become available. Moreover, our results have shown that post-combustion CO<sub>2</sub> capture using state-of-the-art chemical solvent and low-grade steam from power plants may become relatively competitive. Our analysis indicates the importance of looking at CO<sub>2</sub> emissions reduction opportunities in a broader scope than simply trying to reduce onsite emissions.

Secondly, the introduction of CO<sub>2</sub> capture from industrial processes may lead to significant increase in electricity production from CHP plants. In the case of post-combustion CO<sub>2</sub> capture, a CHP plant will likely be built together with the CO<sub>2</sub> capture unit because this is the only way to generate steam efficiently for CO<sub>2</sub> capture solvent regeneration. Also with oxyfuel CO<sub>2</sub> capture, installing an onsite CHP plant can be an economical option because it is crucial to minimize the cost of power consumption for ASU. As the literature review shows, the installment of an onsite CHP plant for CO<sub>2</sub> capture can lead to significant electricity exports to the grid. The increase of CHP-derived electricity in the power market means energy savings, and possibly CO<sub>2</sub> emissions reduction. The economic effect will likely be more positive when the price of natural gas, the main fuel for CHP plants, are kept at lower levels.

## **2.6 Conclusions**

The objective of this paper was to provide a technical and economic assessment of CO<sub>2</sub> capture from iron and steel, cement, and refineries and petrochemical sectors based on a consistent basis. Parameters that are influential on the cost calculations were standardized, as well as direct emissions from the industrial process. Moreover, as we aimed to incorporate the effect of changes in material and energy flows of the industrial process due to process modification as a result of CO<sub>2</sub> capture, the CO<sub>2</sub> emissions accountable for the import/export of process gas, electricity and steam were also taken into account. The CO<sub>2</sub> capture technologies found in the literature were categorized into short-mid term (ST/MT) technologies (likely be commercialized in 10-15 years) and long-term (LT) technologies (likely be commercialized in 20 years or more). The following conclusions are drawn:

### **2.6.1 Iron and steel industry**

In the ST/MT, a CO<sub>2</sub> avoidance cost of 40-65 €/tCO<sub>2</sub> at a CO<sub>2</sub> avoidance rate of about 0.7-0.8 tCO<sub>2</sub>/t hot rolled coil, or 40-45% of the total carbon input to the plant, are possible with top gas recycling blast furnace (TGRBF) technique. This CO<sub>2</sub> capture technique can be retrofitted to existing blast furnaces, but the costs are likely to be somewhat higher. In addition, it is important to note that the uncertainty of the cost performance is high  $\pm 25$  €/tCO<sub>2</sub> avoided for CO<sub>2</sub> capture with VPSA), because the switch from air-blown BF to TGRBF changes the process energy and material flows considerably. Add-on CO<sub>2</sub> capture for air-blown blast furnace may also enable CO<sub>2</sub> capture at similar costs (40-65 €/tCO<sub>2</sub> avoided), but the CO<sub>2</sub> avoidance rate will be only about 0.3-0.4 tCO<sub>2</sub>/t hot rolled coil. CO<sub>2</sub> capture from the COREX process may be done at costs less than 30 €/tCO<sub>2</sub> avoided. However, in order to achieve significant reduction in CO<sub>2</sub> emissions, the COREX process gas needs to be shifted. Advanced add-on CO<sub>2</sub> capture technologies that may emerge in the LT, e.g., membrane separation and hydrate crystallization, may not offer significant economic advantages over conventional technologies (30-55 €/tCO<sub>2</sub> avoided). Selective carbon membranes will enable CO<sub>2</sub> capture from air-blown BF at around 30 €/tCO<sub>2</sub> avoided and the hydrate crystallization technology, which consumes significant amount of electricity, may become an attractive option when low-cost, low-carbon electricity is available.

When building new plants, smelting reduction technologies such as COREX process may become a strong competitor to conventional blast furnace based steel making process in a carbon-constrained society because of its low-cost CO<sub>2</sub> capture. The results also showed that the COREX process with CO<sub>2</sub> capture offers lower steel production cost and lower specific CO<sub>2</sub> emissions compared to the conventional blast furnace based process. Moreover in the LT, advanced smelting reduction technologies will incorporate CO<sub>2</sub> capture as part of production efficiency improvement while achieving significantly lower steel production costs compared to existing technologies. Although conventional iron and steel making using BF is expected to dominate the market in the LT, strong need for drastic CO<sub>2</sub> emissions reduction may drive the sector towards large scale implementation of advanced smelting reduction technologies.

It should also be noted that the DRI process may become important option for low-carbon steel production because it already removes CO<sub>2</sub>. A combination of DRI and smelting reduction with CO<sub>2</sub> removal may offer opportunities for low-cost CCS already in the ST/MT when the CO<sub>2</sub> transport and storage infrastructure is built.

### 2.6.2 Cement sector

For the ST/MT, post-combustion CO<sub>2</sub> capture may likely be the only commercially available technology for the cement sector. With regard to the use of MEA as CO<sub>2</sub> capture solvent, the most economic option seems to be the import of steam from neighboring power plants, which may enable about 65-70 €/tCO<sub>2</sub> avoided at a CO<sub>2</sub> avoidance rate of 0.6 tCO<sub>2</sub>/t clinker, which is about 60% of onsite CO<sub>2</sub> emissions. When the imported steam has a high economic value, the CO<sub>2</sub> avoidance cost may increase up to 90-135 €/tonne. Our results also indicate that it may be more economical not to capture CO<sub>2</sub> from the steam plant (boiler or CHP). The results also indicate that when the use of state-of-the-art CO<sub>2</sub> capture solvent (e.g. KS-1) is combined with the import of power plant steam, the CO<sub>2</sub> avoidance cost may reach below 50 €/tonne.

In the LT, a number of technologies may enable 25-55 €/tCO<sub>2</sub> avoided. Oxyfuel combustion of clinker kiln and precalciner will likely be commercialized, enabling CO<sub>2</sub> capture at around 40 €/tCO<sub>2</sub> avoided. CaO-looping CO<sub>2</sub> capture only from the precalciner may achieve even lower costs (25-30 €/tCO<sub>2</sub> avoided). Post-combustion capture using advanced solvents may also develop significantly with LT cost of 35-55 €/tCO<sub>2</sub> avoided. The cost performance uncertainty is smaller compared to that for the iron and steel sector, because CO<sub>2</sub> capture does not affect either raw material inputs or valuable product outputs.

### 2.6.3 Refineries and petrochemicals

For the ST/MT, our investigation suggests that oxyfuel CO<sub>2</sub> capture (50-60 €/tCO<sub>2</sub> avoided) is more economical than post-combustion capture (>70 €/tCO<sub>2</sub> avoided), though largely dependent on the economic evaluation of the consumed steam, for both combined heaters and furnaces of 2 MtCO<sub>2</sub>/yr scale and catalytic crackers of 1 MtCO<sub>2</sub>/yr scale. However, the retrofit construction for oxyfuel CO<sub>2</sub> capture from heaters and furnaces are expected to be more complicated than that for boilers. The state-of-the-art solvents and other advanced solvents currently under development may bring down the post-combustion capture costs considerably. In the longer term, oxyfuel CO<sub>2</sub> capture with an integrated ASU/CHP plant may become most economical. An onsite CHP plant supplies power for ASU and CO<sub>2</sub> compression, and steam which will replace some of the existing boilers. The CO<sub>2</sub> avoidance cost may become 30 €/tonne or lower, but due to the large excess electricity production, the overall economical performance will strongly be dependent on the conditions of the power market. The CO<sub>2</sub> avoidance rate, calculated around 70-80 % under nominal parameter values, will also depend largely on the credits that can be obtained from the exporting electricity to the grid.



#### **2.6.4 Final remarks**

The economic performance of CO<sub>2</sub> capture technologies will depend largely on individual plants due to the diversity in operational conditions and the type of products produced. Moreover, it was not possible to gain clear insights into the differences in costs between retrofit and greenfield investments for most of the cases due to lack of information in the previous studies. It should, however, be noted that some studies have indicated that the differences would be minor or non-existent, given the fact that CO<sub>2</sub> capture is not completely an add-on technology that could be retrofitted without changes in the process. In addition, CO<sub>2</sub> capture from industrial processes can significantly increase the electricity exports to the grid (chemical absorption capture and advanced oxyfuel combustion capture using OCM). Because of this, the overall CO<sub>2</sub> capture performance may become largely dependent on the conditions of the local power market. Thus a good integration of industrial plants and power plants is essential for cost-effective CO<sub>2</sub> capture. Furthermore, in the case of chemical absorption capture, process integration and the use of waste heat or low-grade industrial heat may become crucial for its economical operation. As indicated for the cement sector, the choice of CO<sub>2</sub> capture heat supply option may affect CO<sub>2</sub> avoidance by nearly 70 €/tonne. Last but not least, some advanced CO<sub>2</sub> capture technologies may lead to more energy efficient industrial production. That may be TGRBF or smelting reduction for iron and steel production, and advanced oxyfuel CO<sub>2</sub> capture for boilers and heaters in the refineries and petrochemical sectors.

This article is one of the first to assess CO<sub>2</sub> capture options for various industrial processes in detail and in a comprehensive way. The standardized key performance data presented in this paper could be a useful input to various energy-economic models that wants to incorporate CO<sub>2</sub> capture from industries. Moreover, the results obtained in this study may help policy makers to set climate policy measures for the industry or specific industrial processes based on their priorities, e.g., cost, CO<sub>2</sub> emissions reduction volume, or timeframe.

However, the results should be taken with caution since most CO<sub>2</sub> capture technologies are in an early stage of development. Therefore, changes in the economics or the technical behavior which were not captured in the present study may still occur. In fact, the current level of development and, more importantly, the lack of demonstration of CO<sub>2</sub> capture from industrial plants outside laboratory and simulation work imply that that it is too early to identify which CO<sub>2</sub> capture technologies would become dominant in the future.

Further research should focus on the topics such as:

- CO<sub>2</sub>, energy and economic performance of industrial CCS under area-specific energy market conditions, e.g., by means of case studies;
- Effect of CO<sub>2</sub> capture on the emissions of air pollutants, e.g. NO<sub>x</sub>, SO<sub>2</sub>, particulate matter, and NH<sub>3</sub>;
- In-depth analysis on the industrial production losses due to retrofitting CO<sub>2</sub> capture (in particular for technologies, e.g., oxyfuel combustion, that require modification in the core of the industrial processes);

- Assessment of potentials for the integration of industrial processes and power plants for cost-effective CO<sub>2</sub> emissions reduction;
- Bottom-up analysis for the cement industry to assess whether and how CO<sub>2</sub> capture will affect the production process;

Assessment and comparison of technical and economic performance of various biofuel production processes with CO<sub>2</sub> capture (raw material input rates, feedstock quality, production rate, product quality etc).

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## Effect of CO<sub>2</sub> capture on the emissions of air pollutants from industrial processes

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Submitted for publication.

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### Abstract

This study assesses whether the deployment of CO<sub>2</sub> capture technologies in the European industrial sector would result in significant changes in emissions of NO<sub>x</sub>, SO<sub>2</sub>, PM and NH<sub>3</sub> in the short-term. The industrial sectors investigated were: cement, petroleum refineries, and iron and steel. The analysis included onsite emissions and changes associated with grid electricity consumption due to CO<sub>2</sub> capture. Chemical absorption capture was considered for the cement sector and petroleum refineries, and top gas recycling blast furnace for the iron and steel sector.

The results show that when all three industrial sectors in the EU-27 are fully equipped with CO<sub>2</sub> capture, total industrial SO<sub>2</sub> emissions may decrease by 90% whereas NH<sub>3</sub> emissions may increase by nearly 900% (12% of the total EU-27 emissions). The large increase in NH<sub>3</sub> emissions is due to the degradation of MEA. Cement and petroleum refineries account for nearly all these changes. The results also show limited impact on NO<sub>x</sub> and PM emissions. Emission changes due to electricity import/export are found to be equally important as onsite emission changes. For the iron and steel sector, the changes in NECD emissions are limited because most emitting processes of the sector are unaffected by CO<sub>2</sub> capture.

### 3.1 Introduction

Industry and petroleum refineries are among the largest contributors to anthropogenic CO<sub>2</sub> emissions, accounting for nearly 40%<sup>11</sup> of these emissions globally (IEA, 2010a). Together with energy efficiency improvement and deploying nuclear energy and renewables, carbon capture and storage (CCS) is gaining attention as a promising option to achieve significant reduction of CO<sub>2</sub> emissions in the atmosphere. The potential of CCS in the industrial and petroleum refining sectors is considered to be significant. The International Energy Agency (IEA) estimates that, in a scenario to halve global greenhouse gas (GHG) emissions in 2050 compared to 2007 level, nearly half of all CCS deployed (up to more than 10 Gt/yr) would be in industrial processes (cement, iron and steel and chemicals) and the fuel transformation sector (petroleum refineries and liquefied natural gas production) (IEA, 2008a).

However, CO<sub>2</sub> is not the only substance emitted from the industrial sector. Since the industrial revolution, power plants and industries have been a major source of environmental pollution such as acidification, eutrophication and toxification of waters. Large efforts have been conducted by governments and industries to mitigate these effects. The deployment of (new) technologies at large scale, such as CCS, should therefore take into account the possibility of side-effects on other industrial plant emissions. Therefore, it is important to assess the possible impact of CO<sub>2</sub> capture technologies on the emissions of air pollutants such as NO<sub>x</sub>, SO<sub>2</sub>, ammonia, non-methane volatile organic compounds (NMVOC) and particulate matter (PM), which contribute to the formation of ground-level ozone and damage ecosystems by acidification and eutrophication. In the European Union (EU), current emission limits for the first four pollutants are set under the EU National Emissions Ceilings Directive (NECD) for 2010 (European Commission, 2001), and all four pollutants are covered by the Gothenburg Protocol of the United Nations Economic Commission for Europe to abate acidification, eutrophication and ground-level ozone (UNECE, 1999).

To date, in the scientific literature there are only a few studies that include possible effects of CO<sub>2</sub> capture on the emissions of air pollutants. These studies have mainly taken the form of life cycle assessment (LCA) studies, particularly those on post-combustion CO<sub>2</sub> capture from power plants. An overview of literature can be found in Koornneef et al. (2010). To our knowledge, such an assessment has not been performed systematically for CO<sub>2</sub> capture from key industrial emission processes.

This study aims to quantitatively assess the possible impacts of CO<sub>2</sub> capture on the emissions of air pollutants from key industrial sectors in Europe. We based the current study on our previous work (Chapter 2), where we present a comparative techno-economic assessment of various CO<sub>2</sub> capture technologies for carbon-intensive industrial processes.

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<sup>11</sup> Includes coke ovens and blast furnaces, and also CO<sub>2</sub> emissions from power generation and process emissions (IEA, 2007a).

In the current study we investigated the following industrial sectors: cement, iron and steel (integrated steel plants), and petroleum refining (furnaces and boilers). The focus is on four key air pollutants: NO<sub>x</sub>, SO<sub>2</sub>, particulate matter (PM) and NH<sub>3</sub>. NMVOC is not considered because there is a lack of quantitative information on the effect of CO<sub>2</sub> capture on NMVOC emissions in the literature (Koorneef et al., 2010). Taking into account the timeframe of the revised Gothenburg Protocol, we investigated CO<sub>2</sub> capture technologies that are likely to become commercially available around 2020. This study focuses on the effects from CO<sub>2</sub> capture and compression only: CO<sub>2</sub> transport and storage are outside the scope of the research. Finally, this study quantifies the *net* effect of CO<sub>2</sub> capture on the emissions of NECD substances. In cases where CO<sub>2</sub> capture leads to an increase in emissions of a NECD substance, additional abatement measures (e.g., wash tower) can be applied using current available technologies at additional cost. These abatement measures were, however, not included in this study.

### **3.2 General description of the sectors studied**

Table 3-1 shows the contribution of iron and steel, cement and petroleum refining sectors to total CO<sub>2</sub> and NECD substance emissions in the EU-27 in 2007 (EEA, 2010, 2011a; IIASA, 2011). The table also presents figures for the public electricity and heat production sector as a reference. Apart from direct CO<sub>2</sub> emissions presented in Table 3-1, industry is accountable for about 570 MtCO<sub>2</sub> emissions for the purchase of electricity from the power sector (EEA, 2011b). This makes industry responsible for about 1.5 GtCO<sub>2</sub> per year, or 36% of total EU-27 emissions. Regarding NECD substance emissions, the cement sector shows large NO<sub>x</sub> emissions and the petroleum refining sector shows large SO<sub>2</sub> emissions relative to CO<sub>2</sub> emissions. NH<sub>3</sub> emissions are negligibly small for all three sectors.

In 2007 the annual cement production in the EU was 283 Mt, accounting for about 11% of world production (Russell, 2008). The cement sector is one of the most CO<sub>2</sub> intensive industrial sectors, accounting for 130 MtCO<sub>2</sub> emissions in the EU-27 in 2007 (IEA, 2010a). High CO<sub>2</sub> emission intensity is not only caused by its large energy requirement but also by the emissions from raw materials, which is mainly comprised of calcium carbonate (main component of limestone). Depending on the clinker/cement ratio, around 60% of the CO<sub>2</sub> emissions originates in the calcination process where clinker is produced (IPPC, 2009b). The rest of the emissions is attributable to combustion of fossil fuels, mostly coal (IPPC, 2009b).

The world total capacity of petroleum refineries is around 4.5 Gt crude oil refined per year, 20% of which is in Europe (IPPC, 2010a). CO<sub>2</sub> emissions from petroleum refineries are nearly 1 Gt/yr worldwide, accounting for around 4% of global total emissions (van Straelen et al., 2010). At the refineries, CO<sub>2</sub> is emitted from various sources such as furnaces and heaters, onsite heat and power plants (including CHP plants), and catalytic crackers. Onsite generation of electricity and heat is responsible for the bulk of the CO<sub>2</sub> emissions in petroleum refineries (Wilkinson et al., 2003). In particular, furnaces and boilers account for 65% of CO<sub>2</sub> emissions from refineries worldwide (IEA GHG, 1999b).

**Table 3-1:** Contribution of iron and steel, cement and oil refining sectors to total CO<sub>2</sub> and NECD pollutant emissions in the EU-27 in 2007. For comparison, data for public electricity and heat production, the total industrial sector, and the total EU-27 emission values are also presented.

<i>Substance</i>	<i>CO<sub>2</sub></i> <sup>1)</sup> [Mt (%-total)]	<i>NO<sub>x</sub></i> [kt NO <sub>2</sub> - equivalent (%-total)]	<i>SO<sub>2</sub></i> [kt (%-total)]	<i>PM<sub>10</sub></i> [kt (%-total)]	<i>NH<sub>3</sub></i> [kt (%-total)]
<i>Sector</i>					
Public electricity and heat production	1.4×10 <sup>3</sup> (33)	2.1×10 <sup>3</sup> (19)	4.4×10 <sup>3</sup> (60)	1.3×10 <sup>2</sup> (6)	4 (0.1)
Industry <sup>2)</sup>	9.3×10 <sup>2</sup> (22)	1.8×10 <sup>3</sup> (17)	1.5×10 <sup>3</sup> (22)	5.8×10 <sup>2</sup> (23)	55 (2)
Iron and steel <sup>3)</sup>	2.1×10 <sup>2</sup> (5)	1.7×10 <sup>2</sup> (2)	2.5×10 <sup>2</sup> (3)	1.1×10 <sup>2</sup> (5)	<1
Cement <sup>4)</sup>	1.7×10 <sup>2</sup> (3)	3.2×10 <sup>2</sup> (3)	1.4×10 <sup>2</sup> (2)	15 (<1)	>1
Petroleum refining <sup>5)</sup>	1.3×10 <sup>2</sup> (3)	1.8×10 <sup>2</sup> (2)	5.8×10 <sup>2</sup> (8)	18 (1)	1
EU-27 total	4.2×10 <sup>3</sup> (100)	1.1×10 <sup>4</sup> (100)	7.3×10 <sup>3</sup> (100)	2.1×10 <sup>3</sup> (100)	3.9×10 <sup>3</sup> (100)

<sup>1)</sup> The figure excludes net CO<sub>2</sub> removals from land use, land-use change and forestry (LULUCF) and emissions from international aviation and international maritime transport (EEA, 2011a).

<sup>2)</sup> Includes the following NFR (New Format for Reporting) sectors: 1A2 (Combustion: Manufacturing industries and construction) and 2 (Industrial processes) (EEA, 2010).

<sup>3)</sup> Includes NFR sectors 1A2A (Combustion: Iron and steel) and 2C1 (Process: Iron and steel production) (EEA, 2010).

<sup>4)</sup> Because the EEA does not publish combustion-related emission figures specific for the cement sector, figures were taken from various sources. CO<sub>2</sub> emissions data (net emissions excluding emissions from electric power) are taken from WBCSD (2011). Net emissions are 10 Mt lower than the gross emissions because the emissions from alternative fuels are excluded from net emission figures.

Emissions data for NO<sub>x</sub>, SO<sub>2</sub> and PM are interpolated from the emissions projection for years 2005 and 2010 in the NEC 2010 Baseline scenario of the GAINS model (IIASA, 2010).

PM emissions data was obtained from EEA (2011a). Although the data does not include emissions from combustion, we did not consider this as an issue because PM emissions from combustion are limited (EEA, 2009).

The minimum value for NH<sub>3</sub> emissions was obtained from EEA (2011a) for process emissions. Data does not include emissions from combustion. EEA (2009) suggests that NH<sub>3</sub> emissions mainly originate in fuel combustion. GAINS model also does not consider NH<sub>3</sub> emissions from these processes.

<sup>5)</sup> Includes NFR sectors 1A1b (Combustion: Petroleum refining) and 1B2a iv (Refining/storage) (EEA, 2010).

The worldwide crude steel production was 1.35 Gt in 2007, of which 210 Mt was produced in the EU-27 (World Steel Association, 2010). The iron and steel sector alone emitted 2.3 GtCO<sub>2</sub> worldwide in 2007, accounting for 30% of total direct CO<sub>2</sub> emissions from the industry (IEA, 2010a). Significant amount of coal is consumed at an iron and steel plant not only to generate heat but also to reduce iron ore in the form of coke. Around 60% of world total steel production is derived from integrated steelmaking plants (IEA, 2007a), which consist of mainly five sections: coking, iron ore agglomeration, blast furnace (BF), basic oxygen furnace (BOF), and final product manufacturing. A general description of carbon flows in an integrated iron and steel production process can be found in, e.g., de Beer et al. (1998), Daniëls (2002) and elsewhere.

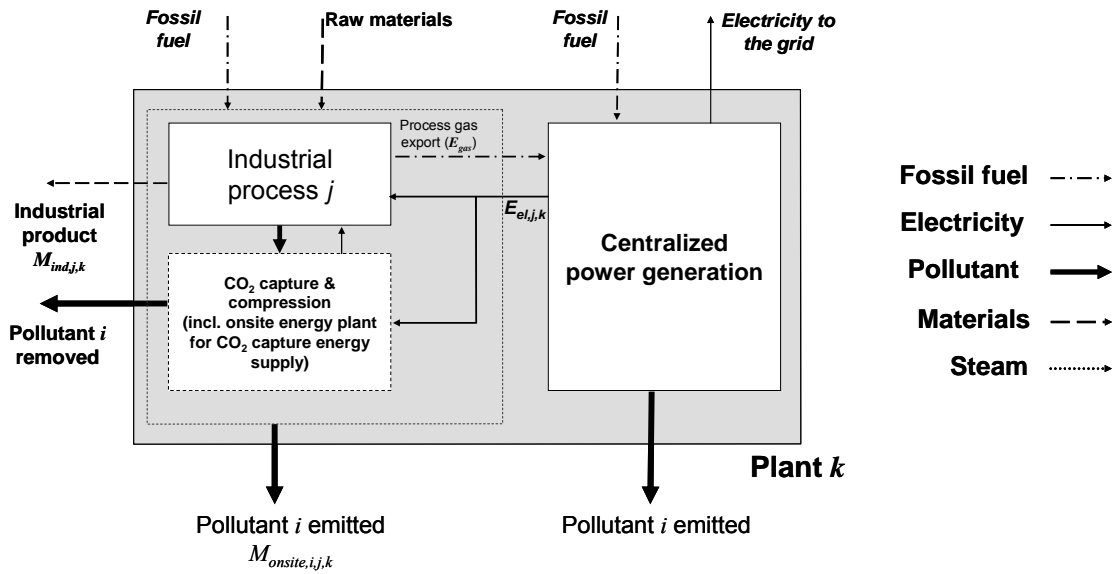


Figure 3-1: System layout and system boundary for the calculation of the emission of pollutants

### 3.3 Methodology

The methodology followed in this study can be summarized in seven main steps. First, specific energy and material consumption and specific CO<sub>2</sub> emissions of a reference plant were defined for each industrial process based on a literature survey on typical plants in Europe. Second, the ranges of specific NECD substance emissions were estimated from the process gas emission/concentration data for existing European plants. Third, short-term (technologies available before or up to 2020) CO<sub>2</sub> capture technologies were selected based on our comparative assessment of CO<sub>2</sub> capture technologies for the industry and petrochemical refineries performed for different timeframes (Chapter 2). In the fourth step, changes in material and energy flows were identified because some CO<sub>2</sub> capture systems are not of an end-of-pipe nature and may alter mass and energy flows in the industrial process, leading to changes in the generation levels of CO<sub>2</sub> and the NECD substances. Fifth, the gas specifications and pollutant removal performance for different CO<sub>2</sub> capture technologies were assessed. Specification for the CO<sub>2</sub> capture unit inlet gas were also looked at because some technologies may require pre-treatment of the process gas before entering the CO<sub>2</sub> capture unit. Sixth, changes in NECD substance emissions were calculated for the reference industrial production for varied NECD substance emission levels obtained in the second step. Lastly, the potential effect of CO<sub>2</sub> capture for the entire EU-27 was calculated using the results from the sixth step and distribution of NECD substance emission levels observed over European plants.

### 3.3.1 System boundary

Figure 3-1 shows the system boundary defined for this study. The system investigated accounts for all onsite emissions, i.e., emissions from the industrial process, onsite power and/or steam generation, CO<sub>2</sub> capture, purification and compression to 110 bar, as well as the emissions associated with electricity purchased from the grid. The CO<sub>2</sub> capture process includes onsite energy plants specifically built to supply energy for CO<sub>2</sub> capture. Therefore, if the onsite energy plant generates excess electricity, then it replaces the centralized electricity generation. There may be cases where CO<sub>2</sub> capture changes the amount of process gas export. In such cases, it was assumed that the change in process gas export affects the natural gas purchase at centralized power plants.

### 3.3.2 Performance indicators and assumptions used throughout the study

#### 3.3.2.1 Specific emission factors

This study uses the specific emission factor of substance  $i$  for the product from industrial process  $j$  (cement:  $Cem$ , refineries:  $Rf$ , and iron and steel:  $IS$ ) as a plant-level technical performance indicator (g/tonne industrial product). For a plant  $k$ , specific emission factors without CO<sub>2</sub> capture (reference case:  $Em_{i,j,k,ref}$ ) and with CO<sub>2</sub> capture ( $Em_{i,j,k,CC}$ ) are defined as follows:

$$Em_{i,j,k,ref} = \frac{M_{onsite,i,j,k,ref}}{M_{ind,j,k}} \quad (3.1)$$

$$Em_{i,j,k,CC} = \frac{M_{onsite,i,j,k,CC} + EF_{i,elec} \Delta E_{el,j,k} - EF_{i,NG} \Delta E_{gas,j,k}}{M_{ind,j,k}} \quad (3.2)$$

where  $M_{onsite,i,j,k,ref}$  is the reference annual onsite emissions from the industrial process  $j$  in the plant  $k$  (g/yr),  $M_{ind,j,k}$  is the annual production from the industrial process  $j$  in the plant  $k$  (tonne industrial product/year),  $M_{onsite,i,j,k,CC}$  is the annual onsite emissions from the industrial process  $j$  in the plant  $k$  equipped with CO<sub>2</sub> capture (g/yr),  $\Delta E_{el,j,k}$  is the change in annual electricity purchase from the grid for the industrial process  $j$  in the plant  $k$  due to CO<sub>2</sub> capture (MJ<sub>e</sub>/yr),  $\Delta E_{gas,j,k}$  is the change in annual net process gas export from the industrial process  $j$  in the plant  $k$  to power plants due to CO<sub>2</sub> capture (MJ<sub>LHV</sub>/yr),  $EF_{i,Elec}$  is the emission factor of substance  $i$  for grid electricity (g/MJ<sub>e</sub>), and  $EF_{i,NG}$  is the emission factor of substance  $i$  for natural gas (g/MJ<sub>LHV</sub>).

#### 3.3.2.2 EU-27 sectoral emissions

The EU-27 sectoral emissions of NECD substances for the industrial process  $j$  with CO<sub>2</sub> capture ( $M_{EU,i,j,CC}$ : kt/yr) were calculated for year 2020 by multiplying the reference EU-27 sectoral emissions in year 2020 by the weighted average relative emission factor for the plants in EU-27 that reported emissions and industrial production data for the recent years in the public literature (e.g. industrial surveys presented in IPPC BREF documents (IPPC, 2010a, b)) (Eq.(3.3)):



$$M_{EU,i,j,CC} = \frac{M_{EU,i,j,ref} \sum_k (Em_{i,j,k,CC} M_{ind,j,k})}{\sum_k (Em_{i,j,k,ref} M_{ind,j,k})} \quad (3.3)$$

where  $M_{EU,i,j,ref}$  is the reference total emissions of NECD substance  $i$  for the industrial process  $j$  in the EU-27 in 2020 (kt/yr).

It was assumed that the reference emissions in year 2020 are equivalent to those in 2007 for all NECD substances and industrial sectors. This assumption can be justified for two reasons. Firstly, a baseline scenario (“National 2010 Baseline”) in the GAINS model developed by the IIASA (2011)<sup>12</sup> indicates that both the industrial activity levels and the related emissions for the three sectors are projected to remain similar for years between 2000 and 2020 (projections for the entire industry can be found in, e.g., Amann et al. (2007); projections for individual industrial sectors can be found in, e.g., IIASA (2011)). The second reason is on the scope of our research which is to provide order-of-magnitude estimates of the effect of CO<sub>2</sub> capture, not exact predictions of emissions in 2020. Therefore, the use of 2007 emissions data would be valid for our study since no significant changes in industrial activity levels and the related emissions are expected towards 2020.

Table 3-2 presents the parameters common for all industrial processes to calculate the impact of CO<sub>2</sub> capture on the emissions of NECD substances. Parameters that are specific to certain CO<sub>2</sub> capture technologies or industrial processes are presented in their respective sections. An important assumption is that no additional pollutant control measures are considered in the case emissions increase as a result of CO<sub>2</sub> capture.

In the following sections, key assumptions and results are presented and discussed per sector.

**Table 3-2:** Emission factors ( $EF$ ) used in this study.

<i>Parameter</i>	<i>Unit</i>	<i>Value (range)</i>
CO <sub>2</sub> emission factor of fuels <sup>1)</sup>		
Coal	g/MJ <sub>LHV</sub>	95
Natural gas	g/MJ <sub>LHV</sub>	56
Grid electricity emission factors ( $EF_{i,Elec}$ ) <sup>2)</sup>		
CO <sub>2</sub> <sup>3)</sup>	g/MJ <sub>e</sub>	130 (90-170)
NO <sub>x</sub> <sup>3)</sup>	g/MJ <sub>e</sub>	0.16 (0.11-0.21)
SO <sub>2</sub> <sup>3)</sup>	g/MJ <sub>e</sub>	0.22 (0.15-0.29)
PM <sup>4)</sup>	g/MJ <sub>e</sub>	$9.9 \times 10^{-3}$ ( $6.9 \times 10^{-3}$ - $1.3 \times 10^{-2}$ )
NH <sub>3</sub> <sup>4)</sup>	g/MJ <sub>e</sub>	$4.9 \times 10^{-4}$ ( $3.5 \times 10^{-4}$ - $6.4 \times 10^{-3}$ )
NECD substance concentration in natural gas ( $EF_{i,NG}$ )		
NO <sub>x</sub> , SO <sub>2</sub> , PM, NH <sub>3</sub>	g/MJ <sub>LHV</sub>	Negligible

<sup>12</sup> The GAINS model deals with costs and potentials of air pollution control and greenhouse gas mitigation and assesses interactions between policies in the medium-term (until 2030).

- <sup>1)</sup> See (IPCC, 2006).
- <sup>2)</sup> High and low values represent  $\pm 30\%$  of the nominal values.
- <sup>3)</sup> The emission intensities of CO<sub>2</sub>, NO<sub>x</sub>, SO<sub>2</sub> in public electricity and heat production in the EEA32 (27 EU member states plus Iceland, Liechtenstein, Norway, Switzerland and Turkey) in 2007 were 620 g/kWh, 0.95 g/kWh (NO<sub>2</sub> equivalent), and 2.2 g/kWh, respectively (EEA, 2011c). In the past decades, significant reductions in the emission intensities of have been observed the three substances and we expect a similar trend between 2007 and 2020. The average annual emission intensity reduction was 1.5%/yr between 1990 and 2007 and 2.3 %/yr between 1990 and 2007 for CO<sub>2</sub>; 4.4%/yr between 1990 and 2007 and 3.8 %/yr between 2000 and 2007 for NO<sub>x</sub>; and 7.0%/yr between 1990 and 2007 and 8.1 %/yr between 2000 and 2007 for SO<sub>2</sub>. We therefore assumed annual reduction rates of 2% in CO<sub>2</sub> emission intensity, 4% in NO<sub>x</sub> emission intensity, and 7.5% in SO<sub>2</sub> emission intensity, respectively (EEA, 2011c).
- <sup>4)</sup> The PM<sub>10</sub> and NH<sub>3</sub> emission intensities of public electricity and heat production in the EU-27 in 2007 were calculated to be  $6.1 \times 10^{-2}$  g/kWh and  $1.9 \times 10^{-3}$  g/kWh, respectively (data on the emissions from public electricity and heat production from EEA (2011a), data on the total outputs from public thermal power stations from EEA (2011c)). Although there is no data available on the trend of PM emission intensity in the previous years, PM emission intensity was assumed to reduce 4% annually, taking into account the annual emission intensity reduction rate for CO<sub>2</sub>, NO<sub>x</sub> and SO<sub>2</sub>. NH<sub>3</sub> emission intensity was assumed to remain unchanged.

### 3.4 Cement production

For the cement sector, clinker kiln is the most CO<sub>2</sub>-emitting process and where CO<sub>2</sub> capture process will likely be fitted to. Therefore, we considered the emission changes in cement plants with clinker kiln.

#### 3.4.1 Reference plant assumptions

Table 3-3 shows the typical production and the emissions data of a reference European cement plant. Cement production can be distinguished by the moisture content of the feed going into the kiln. Today's best available technique (BAT)<sup>13</sup> is based on the dry process (IEA GHG, 2008). Dry process cement plants account for about 90% of European total cement production (IEA GHG, 2008). In this study, a dry process cement plant was assumed. The emissions of NECD substances from cement plants are largely attributable to fuel combustion. Note that cement plants extensively use waste as a fuel, which largely affects the pollutants' emission profiles (IPPC, 2009b).

For NO<sub>x</sub>, SO<sub>2</sub> and PM, flue gas concentration data are available for most European cement plants. Large differences in NECD substance concentrations among plants can be seen. Therefore in this study, the effect of CO<sub>2</sub> capture on the emissions of these substances was assessed for a range of concentration levels instead of an average value.

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<sup>13</sup> The term "best available techniques" is defined by the EU Directive (European Commission, 2008) as "the most effective and advanced stage in the development of activities and their methods of operation which indicate the practical suitability of particular techniques for providing in principle the basis for emission limit values designed to prevent and, where that is not practicable, generally to reduce emissions and the impact on the environment as a whole".

**Table 3-3:** Reference European cement plant specification and the range of NECD substance concentrations in the clinker kiln flue gas investigated in this study. Reference: IPPC (2009b), unless otherwise stated.

Parameter	Unit	Symbol	Value
Reference process specification			
Clinker/cement ratio	kg/kg		0.8
Flue gas CO <sub>2</sub> concentration <sup>1)</sup>	vol%		22.5
Flue gas O <sub>2</sub> concentration <sup>2)</sup>	vol%		10
Flue gas NO <sub>2</sub> /NO <sub>x</sub> ratio <sup>3)</sup>	mol/mol		0.05
Specific flue gas volume	10 <sup>3</sup> Nm <sup>3</sup> /t clinker	$V_{Sp,Cem}$	2.3
Specific CO <sub>2</sub> emission <sup>4)</sup>	tCO <sub>2</sub> /t clinker		0.81
Range of NECD substance concentrations in the flue gas			
NO <sub>x</sub> (as NO <sub>2</sub> ) <sup>5)</sup>	mg/Nm <sup>3</sup>	$C_{NOX,Cem}$	150-1200
SO <sub>2</sub> <sup>6)</sup>	mg/Nm <sup>3</sup>	$C_{SO2,Cem}$	50-400
PM <sup>7)</sup>	mg/Nm <sup>3</sup>	$C_{PM,Cem}$	0.27-30
NH <sub>3</sub> <sup>8)</sup>	mg/Nm <sup>3</sup>	$C_{NH3,Cem}$	10-35

<sup>1)</sup> Flue gas CO<sub>2</sub> concentration ranges around 15-30% (IEA GHG, 2008).

<sup>2)</sup> Based on IEA GHG (2008).

<sup>3)</sup> This value has been reported for the rotary kiln flue gas from a number of existing plants in Europe. The literature values range between 2% (Leibacher et al., 2006) and 10% (IEA GHG, 2008).

<sup>4)</sup> The value is calculated using the presented assumptions and it agrees with the range (0.73-0.93 tCO<sub>2</sub>/t cement) indicated in Mahasen et al. (2005).

<sup>5)</sup> NO<sub>x</sub> emissions are largely attributable to fuel combustion in the clinker furnace. The NO<sub>x</sub> formation level is around 1 g/Nm<sup>3</sup> and the generated NO<sub>x</sub> is reduced by selective non-catalytic reduction (SNCR) in most cement plants. With SNCR, NO<sub>x</sub> is reduced to N<sub>2</sub> by injecting ammonia water (up to 25% NH<sub>3</sub>), ammonia precursor compounds or urea solution into the combustion gas at about 830-1050 °C. About 150 of 258 reported installations in Europe show emissions levels below 0.8 g/Nm<sup>3</sup>.

<sup>6)</sup> SO<sub>2</sub> is the main sulfur compound (99%) emitted from cement plants and its emission levels are primarily determined by the content of the volatile sulfur in the raw materials and possibly by the fuels. SO<sub>2</sub> emissions originate in various outputs of the kiln system: exhaust gases, CaSO<sub>4</sub> and other combined compounds in the clinker, and PM. Around two-thirds of the reported installations in Europe show emission levels below 0.2 g/Nm<sup>3</sup>.

<sup>7)</sup> Continuously measured emissions. Emission levels of below 10 mg/Nm<sup>3</sup> are achieved in 37% of the 253 reported installations in Europe, the dust emissions from most plants were less than 30 mg/Nm<sup>3</sup>.

<sup>8)</sup> NH<sub>3</sub> is emitted via ammonia slip value from de-NO<sub>x</sub> process (SNCR) as well as via processing of raw materials.

### 3.4.2 Short term CO<sub>2</sub> capture technology

Based on an assessment made by the IEA GHG (2008) and Hegerland et al. (2006), post-combustion CO<sub>2</sub> capture using monoethanolamine (MEA) is considered to be the short-term technology to be applied. Post-combustion CO<sub>2</sub> capture is conceptually an end-of-pipe technology that does not require fundamental changes in the clinker production process (Mott MacDonald, 2010). In this study, it was assumed that the cement production process itself is unaffected by chemical absorption CO<sub>2</sub> capture. This assumption agrees with that in previous studies (Hegerland et al., 2006; IEA GHG, 2008). Table 3-4 shows the flue gas specifications and the pollutant removal performance for this technology. It

was assumed that CO<sub>2</sub> is captured from all process gas flows (IEA GHG, 2008). The solvent was assumed to be a 30 wt% MEA solution.

The following three techniques can be used to control SO<sub>2</sub> emissions: absorbent addition, wet scrubbing, and activated carbon filter (IPPC, 2010b). Among them, the wet scrubbing process using CaCO<sub>3</sub> is likely to be used to meet the CO<sub>2</sub> scrubber specifications (Table 3-4). Wet scrubbing also removes other air pollutants, notably particulates. In this study, the conservative assumption was made that for additional desulfurization prior to the CO<sub>2</sub> scrubber, co-removal of other NECD substances is not taking place.

**Table 3-4:** Flue gas specifications and pollutant removal performance for post-combustion CO<sub>2</sub> capture using monoethanolamine (MEA)

<i>Parameter</i>	<i>Unit</i>	<i>Symbol</i>	<i>Value</i>
Flue gas specification			
NO <sub>2</sub> content <sup>1)</sup>	mg/Nm <sup>3</sup>	$C_{NOX,lim}$	20
SO <sub>2</sub> content <sup>2)</sup>	mg/Nm <sup>3</sup>	$C_{SO2,lim}$	29
PM content <sup>3)</sup>	mg/Nm <sup>3</sup>	$C_{PM,lim}$	15
Post-combustion CO <sub>2</sub> capture performance			
CO <sub>2</sub> removal efficiency <sup>4)</sup>	%		85
NO <sub>x</sub> removal efficiency <sup>4)</sup>	%	$r_{NOX}$	NO <sub>2</sub> : 25, NO: 0
SO <sub>2</sub> removal efficiency <sup>4)</sup>	%	$r_{SO2}$	90
PM removal efficiency <sup>4)</sup>	%	$r_{PM}$	50
NH <sub>3</sub> emissions due to additional de-NO <sub>x</sub>			
NH <sub>3</sub> consumption rate <sup>5)</sup>	mol NH <sub>3</sub> /mol NO reduced		1.5
Ammonia slip rate <sup>4)</sup>	%- total NH <sub>3</sub> consumption		1

<sup>1)</sup> Based on IEA GHG (2004a).

<sup>2)</sup> The nominal value has been suggested by several studies for 30 wt% MEA solvents (Chapel et al., 1999; IEA GHG, 2004a). For Kerr McGee/ABB Lummus Crest solvent, which is a 15-20 wt% MEA solution, it has been suggested that SO<sub>2</sub> removal is economically unjustifiable for concentrations below 50 ppm (142 mg/Nm<sup>3</sup>) (Barchas and Davis, 1992).

<sup>3)</sup> Based on Chapel et al. (1999).

<sup>4)</sup> Based on Koornneef et al. (2008).

<sup>5)</sup> For SNCR, the ratio depends on the initial NO<sub>x</sub> concentration and the degree of NO<sub>x</sub> reduction. Two Swedish plants reduce NO<sub>x</sub> emission levels to below 200 mg/Nm<sup>3</sup>, which is consistent with the post-combustion CO<sub>2</sub> capture requirements, at NH<sub>3</sub>/NO ratios of 1.2-1.8 (IPPC, 2009b).

### 3.4.2.1 Post-combustion CO<sub>2</sub> capture: energy requirements

The reboiler duty of chemical absorption capture process is affected by the flue gas CO<sub>2</sub> concentration. A modeling study by Kothandaraman et al. (2009) reports 4.5 GJ<sub>th</sub>/tCO<sub>2</sub> for natural gas combustion gas and 4.2 GJ<sub>th</sub>/tCO<sub>2</sub> for coal combustion gas. The difference is due to the rich loading that can be achieved for coal combustion gas. Assuming 3% CO<sub>2</sub> for natural gas combustion gas and 15% CO<sub>2</sub> for coal combustion gas, the following linear equation was derived:

$$H_{Sp} = -2.5x + 4.58 \quad (0.03 \leq x \leq 0.15) \quad (3.4)$$

where  $H_{Sp}$  is the specific reboiler duty (GJ<sub>th</sub>/tCO<sub>2</sub> captured) and  $x$  is the CO<sub>2</sub> concentration (v/v). The assumption of a linear relation between specific reboiler duty and flue gas CO<sub>2</sub> concentration can be justified for the range of CO<sub>2</sub> concentration considered in this study (3-15 vol%) according to literature (e.g., Alie et al., 2005; Egberts et al., 2003)<sup>14</sup>.

The specific reboiler duty derived from Eq.(3.4) is higher than what is proven by technology providers. The referenced study (Kothandaraman et al. (2009)) also stated that the specific reboiler duty calculated by their model is not among the lowest observed in the literature for commercial chemical solvents. While specific reboiler duty will likely be much lower for new plants, it would be somewhat higher for retrofits. It is suggested that the specific reboiler duty would be 30% higher for retrofits compared to new plants (Egberts et al., 2003). Taking into account that our study assumes retrofit CO<sub>2</sub> capture, we consider the assumed range for specific reboiler duty to be valid.

The equation for calculating specific power requirement ( $P_{Sp}$ : GJ<sub>e</sub>/tCO<sub>2</sub> captured) is taken from Chapel et al. (1999):

$$P_{Sp} = 0.014/x + 0.035 \quad (3.5)$$

For  $H_{Sp}$  and  $P_{Sp}$ , an uncertainty of ±10% was assumed and their influence is presented with the results.

#### **3.4.2.2 NH<sub>3</sub> formation rate from MEA degradation**

Besides the slip from de-NO<sub>x</sub> installation, NH<sub>3</sub> can be formed as a result of oxidative degradation of MEA by flue gas impurities such as SO<sub>2</sub>, NO<sub>2</sub> and O<sub>2</sub>. In this study, the oxidative degradation rate of MEA is defined as follows (adapted from Veltman et al., 2010):

$$[\text{NH}_3] = [\text{MEA}]_{\text{ox-deg}} = 2f_{\text{SO}_2} [\text{SO}_2] + 2f_{\text{NO}_2} * [\text{NO}_2] + f_{\text{O}_2} [\text{O}_2] \quad (3.6)$$

where  $[\text{NH}_3]$  is the NH<sub>3</sub> formation rate (mol/s),  $[\text{MEA}]_{\text{ox-deg}}$  is the oxidative degradation rate of MEA (mol/s),  $f_{\text{SO}_2}$  is the reaction efficiency of SO<sub>2</sub> with MEA (= SO<sub>2</sub> removal efficiency: 90%),  $f_{\text{NO}_2}$  is the reaction efficiency of NO<sub>2</sub> with MEA (= NO<sub>2</sub> removal efficiency: 25%),  $[\text{SO}_2]$  is the SO<sub>2</sub> flow rate (mol/s),  $[\text{NO}_2]$  is the NO<sub>2</sub> flow rate (mol/s),  $f_{\text{O}_2}$  is the reaction rate of MEA with O<sub>2</sub> ( $1.8 \times 10^{-4}$  mol MEA/mol O<sub>2</sub>) and  $[\text{O}_2]$  is the O<sub>2</sub> flow rate (mol/s). The underlying assumption is that each mole of MEA degraded produces one mole of NH<sub>3</sub> (Rao et al., 2004; Veltman et al., 2010).

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<sup>14</sup> The linear relation is not valid for CO<sub>2</sub> concentrations above 15 vol%, where specific reboiler duty is reported to become nearly constant (Alie et al., 2005; Egberts et al., 2003).

The  $f_{O_2}$  value was calculated based on an experimental study by Goff and Rochelle (2004), which reports the  $O_2$ -induced degradation rate of MEA to be 0.29-0.73 kg/t  $CO_2$  captured for a flue gas containing 3%  $CO_2$  and 5%  $O_2$ , and 30 wt% MEA solution at 90% capture efficiency. Using the geometric mean value of 0.46 kg MEA/t  $CO_2$  captured as suggested by Veltman et al. (2010), we assumed the  $O_2$ -induced oxidative degradation rate to be 0.34 kg MEA per tonne of  $O_2$  in the flue gas, which corresponds to the  $f_{O_2}$  value presented above. Note that the MEA degradation rate due to  $O_2$  is mass transfer dependent rather than kinetic rate dependent for solvents with MEA concentration above 30 wt% (Goff and Rochelle, 2004). It is also reported that the  $NH_3$  formation due to the degradation of MEA is linear with flue gas  $O_2$  concentration up to 17 vol% and increases at larger  $O_2$  concentrations (Goff and Rochelle, 2004).

### 3.4.2.3 Heat supply options for post-combustion $CO_2$ capture in cement plants

Chemical absorption  $CO_2$  capture using MEA requires significant amounts of heat to regenerate the solvent. Cement plants often do not have a usable heat source of significant scale in their vicinity. Consequently, solvent regeneration needs heat either generated onsite by a boiler or a combined heat and power (CHP) plant, or imported from external sources. Because of its high energy efficiency, the use of CHP plants for  $CO_2$  capture heat generation was investigated. In this study, we considered natural gas and coal as possible CHP fuels. Moreover, we considered two possibilities with regard to  $CO_2$  generated by the CHP plant: 1) vented; 2) captured together with the  $CO_2$  from the industrial process. Table 3-5 presents three heat supply options and their respective emissions profiles. In this study, we assumed that CHP plants installed for post-combustion  $CO_2$  capture were considered to be equipped with BAT.

**Table 3-5:** Three heat supply options and their emissions profile. Source: IPCC (2006), unless otherwise stated.

<i>Heat supply configuration</i>	<i>I-PC</i>	<i>I-NG</i>	<i>S-NG</i>
Configuration	Integrated PC-CHP	Integrated NG-CHP	Stand-alone NG-CHP
Treatment of $CO_2$ from CHP plant	Capture together with $CO_2$ from industrial process	Capture together with $CO_2$ from industrial process	Vented
CHP plant type	Pulverized coal combustion boiler + steam turbine	Combined cycle	Combined cycle
Steam generation efficiency <sup>1)</sup>	50%	50%	50%
Electricity generation efficiency <sup>1)</sup>	30%	40%	40%
Concentration of pollutants in the CHP flue gas			
$NO_x$ (as $NO_2$ ; mg/Nm <sup>3</sup> ) <sup>2)</sup>	100	50	50
$NO_2/NO_x$ (vol%) <sup>3)</sup>	5	10	10
$SO_2$ (mg/Nm <sup>3</sup> ) <sup>4)</sup>	200	0	0
$NH_3$ (mg/Nm <sup>3</sup> ) <sup>5)</sup>	5	0	0
PM (mg/Nm <sup>3</sup> ) <sup>6)</sup>	10	0	0
Flue gas $O_2$ concentration (vol%)	6	15	15

- 1) For PC-CHP plants, an exergetic efficiency of 45-55% and an overall energy efficiency of 75-90% have been suggested for BAT coal fired CHP plants (IPPC, 2006).
- 2) For coal-fired PC-CHP plants, emission levels of 90-300 mg/Nm<sup>3</sup> for 50-200 MW<sub>th</sub> plants, 90-200 mg/Nm<sup>3</sup> for 100-300 MW<sub>th</sub> plants, and 50-200 mg/Nm<sup>3</sup> for >300 MW<sub>th</sub> plants have been suggested. For NGCC-CHP plants, emission levels of 20-50 mg/Nm<sup>3</sup> have been suggested for new plants. For both coal and gas-fired plants, we made the conservative assumption of taking high-end values (IPPC, 2006).
- 3) Based on Tzimas et al. (2007).
- 4) For PC-CHP plants, emission levels of 150-400 mg/Nm<sup>3</sup> for 50-100 MW<sub>th</sub> plants, 100-200 mg/Nm<sup>3</sup> for 100-200MW<sub>th</sub> plants, and 20-200 mg/Nm<sup>3</sup> for >300MW<sub>th</sub> plants have been suggested (IPPC, 2006). No SO<sub>2</sub> emissions were assumed for NGCC-CHP plants (IPPC, 2006).
- 5) No NH<sub>3</sub> emissions are expected for NGCC-CHP plants as a dry low-NO<sub>x</sub> burner was assumed to be used.
- 6) For PC-CHP plants, emission levels of 5-20 mg/Nm<sup>3</sup> for 50-300 MW<sub>th</sub> plants and 5-10 mg/Nm<sup>3</sup> for >300 MW<sub>th</sub> plants have been suggested. No PM emissions are expected for NGCC-CHP plants (IPPC, 2006).

### 3.4.2.4 Additional flue gas treatment required for post-combustion CO<sub>2</sub> capture

The flue gas from a cement plant and a CHP plant may require pre-treatment before entering the CO<sub>2</sub> capture unit to assure good technical performance and to enable economical operation. Figure 3-2 shows flue gas flows to the CO<sub>2</sub> scrubber and the assessment made of whether additional flue gas treatment prior to CO<sub>2</sub> scrubbing is needed.

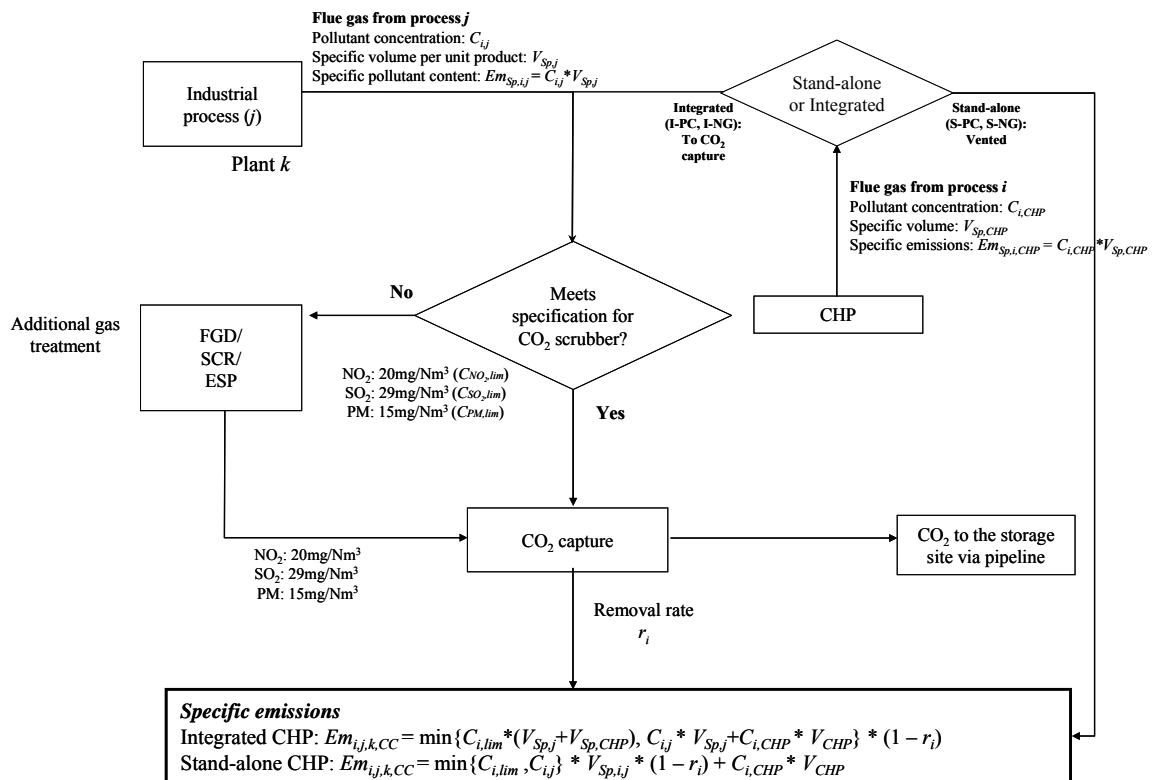


Figure 3-2: Flue gas flows to the post-combustion CO<sub>2</sub> capture unit and the need for additional flue gas treatment prior to CO<sub>2</sub> scrubbing.

### 3.4.3 Results

#### 3.4.3.1 Energy and CO<sub>2</sub> balance

Table 3-6 shows the CO<sub>2</sub> concentration of the process flue gas at the CO<sub>2</sub> scrubber inlet, the CO<sub>2</sub> capture rate and the specific heat requirement for CO<sub>2</sub> capture solvent regeneration. CO<sub>2</sub> concentrations are significantly lower for the integrated CHP option because the CHP flue gas dilutes the cement kiln flue gas.

**Table 3-6:** Technical results of post-combustion CO<sub>2</sub> capture from the reference cement plant for three different CO<sub>2</sub> capture heat supply options

CHP configuration	Flue gas CO <sub>2</sub> concentration [v/v]	CO <sub>2</sub> capture rate [t/t clinker]	CO <sub>2</sub> capture heat requirement		Cogenerated power [GJ <sub>e</sub> /t clinker]	Excess power generation [GJ <sub>e</sub> /t clinker]	CO <sub>2</sub> avoidance rate	
			[GJ/tCO <sub>2</sub> captured]	[GJ/t clinker]			Onsite only	Incl. electricity export credits
I-PC	14.2%	2.7	4.2	11.4	6.9	5.4	53%	122%
I-NG	5.7%	1.49	4.4	6.6	5.3	4.3	74%	143%
S-NG	22.5%	0.86	4.0	3.5	2.8	2.3	47%	68%

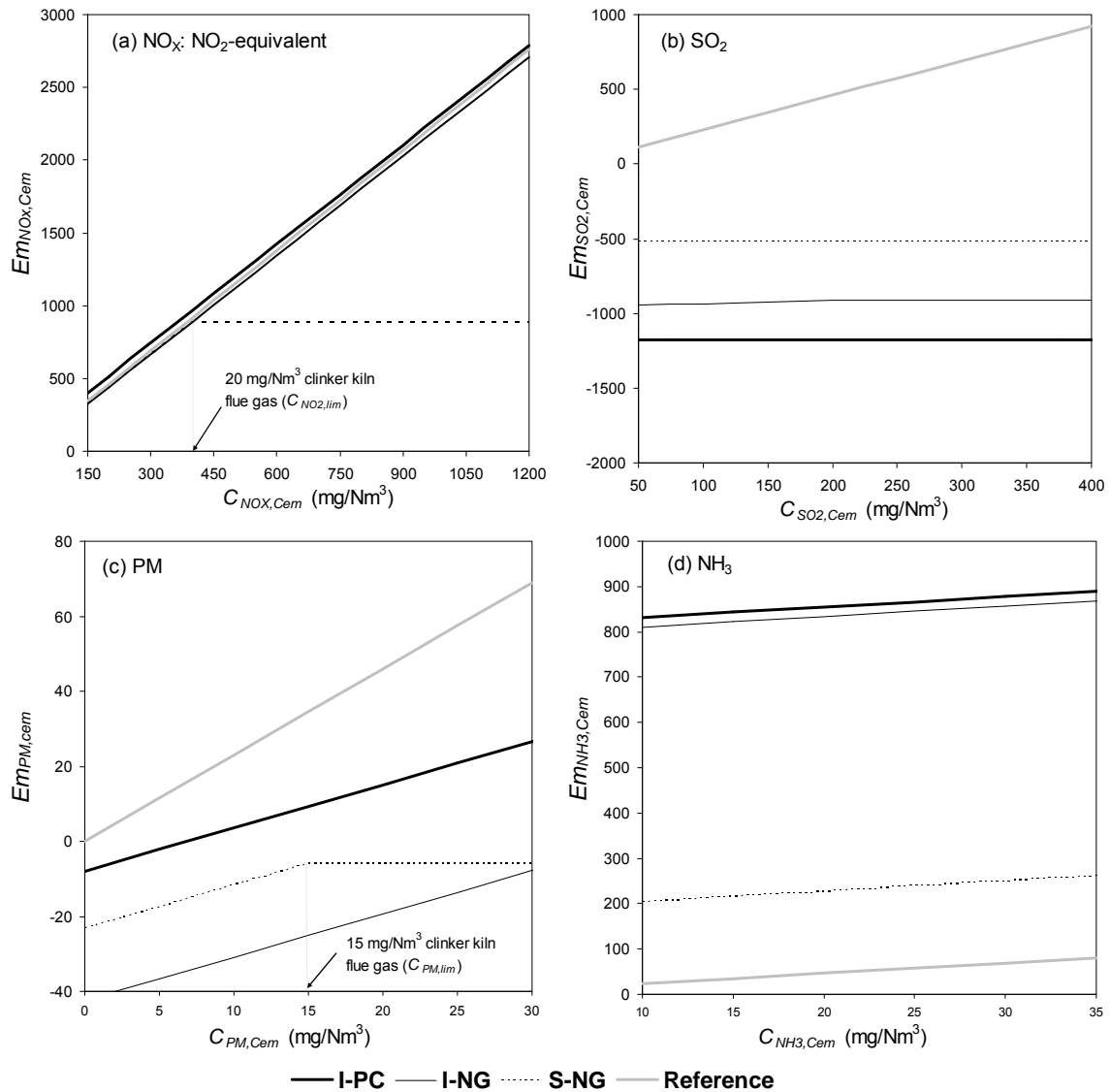
#### 3.4.3.2 Changes in NECD substance emissions by flue gas concentration level

Figure 3-3 shows specific emissions ( $Em_{i,Cem}$ : g/t clinker) of NECD substances from cement plants with and without CO<sub>2</sub> capture as a function of clinker kiln flue gas concentration ( $C_{i,cem}$ ), and Figure 3-4 shows the breakdown of specific emissions by onsite emissions and emission credits for avoided centralized electricity production when equipped with CO<sub>2</sub> capture.

Figure 3-3 shows that the heat supply configuration affects the results significantly. Specific NO<sub>x</sub> emissions ( $Em_{NO_x,Cem}$ ) remain constant for S-NG for  $C_{NO_x,Cem}$  above 400 mg/Nm<sup>3</sup>, at which NO<sub>2</sub> concentration reaches the CO<sub>2</sub> scrubber limit ( $C_{NO_x,lim}$ : 20 mg/Nm<sup>3</sup>). For integrated heat supply options (I-PC and I-NG), in contrast,  $Em_{NO_x,Cem}$  does not curb for the range of  $C_{NO_x,Cem}$  investigated because CHP flue gas dilutes the cement plant flue gas with regard to NO<sub>2</sub>, thus additional de-NO<sub>x</sub> prior to the CO<sub>2</sub> scrubber is not required. Both I-PC and I-NG shows emission levels similar to the reference case because the increase in onsite emissions is offset by the avoided emissions in centralized power plants.

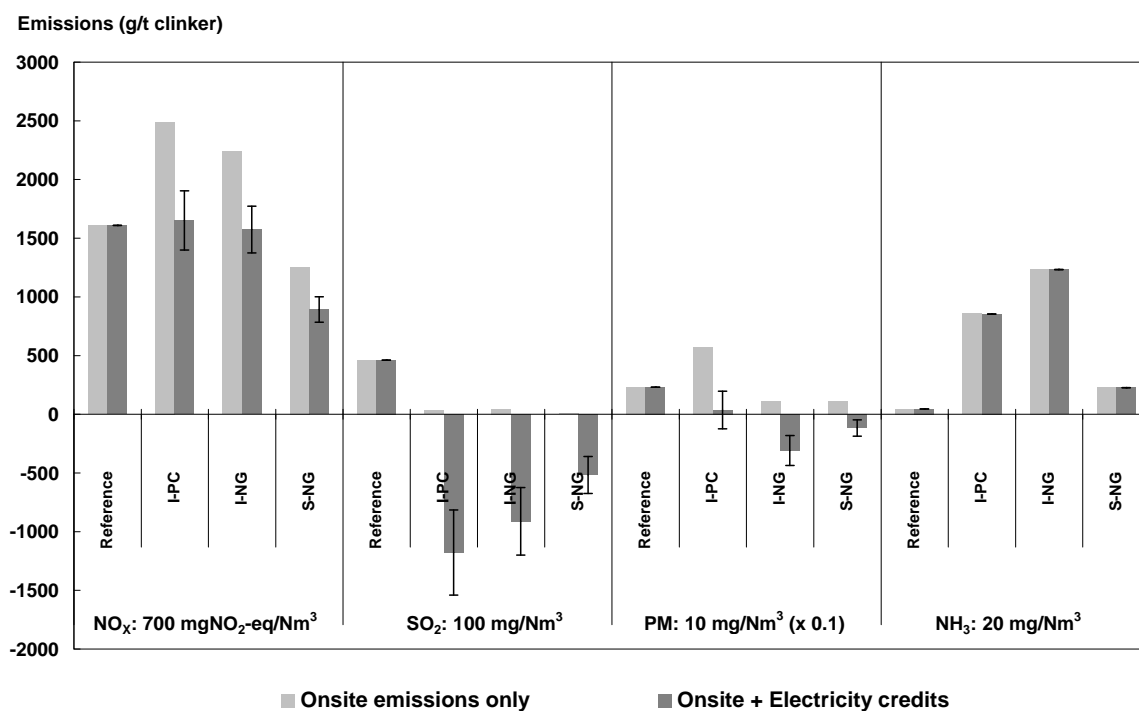
Specific SO<sub>2</sub> emissions ( $Em_{SO_2,Cem}$ ) were found to become negative for all heat supply options and  $C_{SO_2,Cem}$  levels investigated. Onsite SO<sub>2</sub> emissions are reduced by about 90% or more by the CO<sub>2</sub> capture process together with possible flue gas pre-treatment. Moreover, emissions reduction is credited for avoided centralized electricity production due to excess electricity production by the CHP plant.  $Em_{SO_2,Cem}$  is therefore sensitive to the grid electricity SO<sub>2</sub> emission factor (Figure 3-4).





**Figure 3-3:** Specific emissions of NECD substances for cement production ( $Em_{i,Cem}$ : g/t clinker) with post-combustion CO<sub>2</sub> capture as a function of pollutant concentration of each NECD substance cement flue gas. The flue gas specifications for the CO<sub>2</sub> scrubber are 15 mg/Nm<sup>3</sup> for PM, 20 mg/Nm<sup>3</sup> for NO<sub>2</sub>, and 29 mg/Nm<sup>3</sup> for SO<sub>2</sub>. NH<sub>3</sub> emissions are calculated for 100 mg/Nm<sup>3</sup> SO<sub>2</sub> and 700 mg-NO<sub>2</sub> eq./Nm<sup>3</sup> NO<sub>x</sub>.

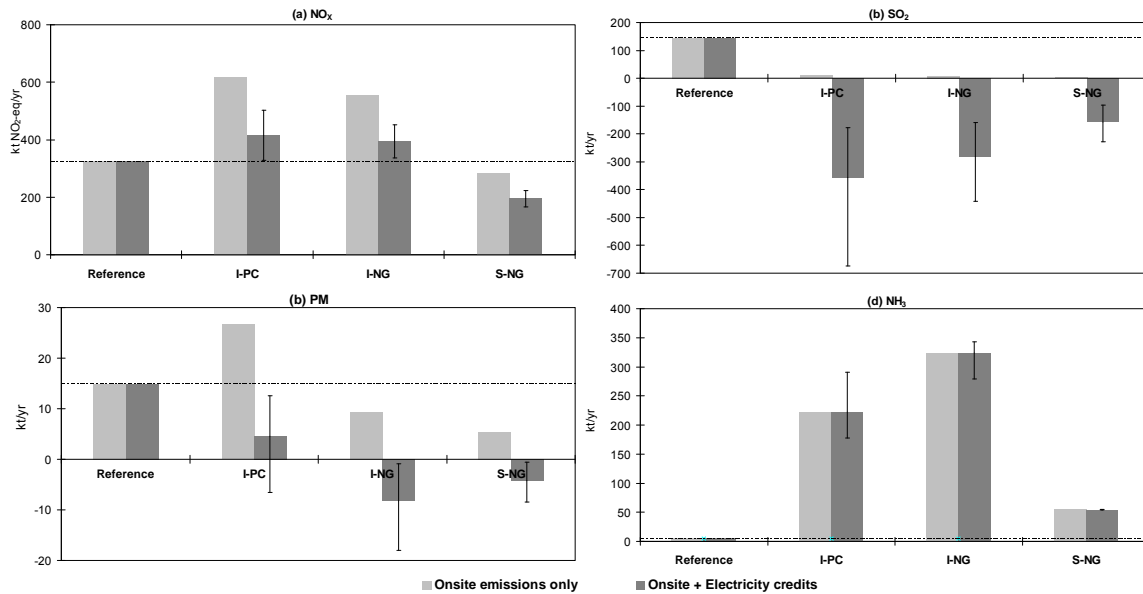
Specific PM emissions ( $Em_{PM,Cem}$ ) are reduced significantly for all heat supply options. For integrated heat supply options (I-PC and I-NG), cement plant flue gas is diluted by the CHP flue gas. The PM concentration of the combined flue gas therefore reaches  $C_{PM,lim}$  at much higher  $C_{PM,Cem}$  than for stand-alone heat supply options (outside of the figure). Negative  $Em_{PM,Cem}$  are observed for I-NG and S-NG because of emission credits for avoided centralized electricity production (Figure 3-4).



**Figure 3-4:** Breakdown of specific emissions of NECD substances for cement production (g/t clinker) with post-combustion CO<sub>2</sub> capture. The high and low range bars represent the uncertainty in grid electricity emission factors.

Finally, NH<sub>3</sub> emissions increase for all heat supply options. Avoided NH<sub>3</sub> emissions due to electricity export are negligible (Figure 3-4). The emission levels for S-NG is factor 3-4 lower than the other two options mainly because of the lower O<sub>2</sub>-induced MEA degradation as CO<sub>2</sub> is captured only from cement process flue gas.

In order to examine the approach of NH<sub>3</sub> calculations applied in this study, we compared the NH<sub>3</sub> formation rate per tonne of CO<sub>2</sub> captured (without water wash) obtained in this study with values reported in the literature. For higher flue gas CO<sub>2</sub> concentrations, for instance, Koornneef et al., (2008) report 0.21 kgNH<sub>3</sub>/tCO<sub>2</sub> captured for PC power plants (about 12-14 vol% CO<sub>2</sub>). In this study, the I-PC case showed a NH<sub>3</sub> formation rate of 0.32 kgNH<sub>3</sub>/tCO<sub>2</sub> captured for 14.2 vol% flue gas CO<sub>2</sub> concentration, which is similar to that of PC power plants. For lower CO<sub>2</sub> concentrations, Statoil (2005) reports a NH<sub>3</sub> formation rate of about 0.4 kgNH<sub>3</sub>/tCO<sub>2</sub> captured for NGCC plants (generally 3-4 vol% CO<sub>2</sub>). For the I-NG case in this study with zero SO<sub>2</sub> content in the cement flue gas (to enable a fair comparison with natural gas combustion flue gas, which generally contains negligible amount of SO<sub>2</sub>), the NH<sub>3</sub> formation rate was calculated to be 0.28 kgNH<sub>3</sub>/tCO<sub>2</sub> captured for 5.7 vol% CO<sub>2</sub> concentration. These results strongly indicate that the approach applied for the calculation of NH<sub>3</sub> formation is valid.



**Figure 3-5:** Annual total emissions of NECD substances from the cement sector in the EU-27 countries with and without post-combustion CO<sub>2</sub> capture using MEA. The error bars represent the high and low values observed for varied heat supply options, CO<sub>2</sub> capture energy requirement and grid electricity emission factors.

### 3.4.3.3 Changes in European sectoral emissions

Regarding EU-27 total sectoral NECD emissions from cement production process ( $M_{EU,i,Cem,ref}$ ), it was assumed that the cement plants with clinker kiln, from which CO<sub>2</sub> is captured, are responsible for all emissions from the sector. We therefore refer to Table 3-1 to estimate total emissions for NO<sub>x</sub>, SO<sub>2</sub> and NH<sub>3</sub>;  $M_{EU,NOX,Cem,ref}$ ,  $M_{EU,SO2,Cem,ref}$  and  $M_{EU,PM,Cem,ref}$  are 320 kt/yr, 140 kt/yr and 15 kt/yr, respectively.  $M_{EU,NH3,Cem,ref}$  was calculated bottom-up using plant-level data reported in IPCC (2010b)<sup>15</sup>. To calculate EU-27 total sectoral emissions with CO<sub>2</sub> capture ( $M_{EU,i,Cem,CC}$ ), data on the pollutant concentrations in the flue gas (continuous measurement results) reported from 253 clinker kilns (of a total of 268 plants in EU-27; IPCC, 2010b) was used. We made two important assumptions for the calculations. Firstly, it was assumed that all the NECD substances are emitted through the clinker kiln flue gas, from which CO<sub>2</sub> is captured. Secondly, we assumed an identical clinker production scale per plant ( $M_{ind,Cem,k}$ ) and flue gas CO<sub>2</sub> concentration for all plants (22.5%; Table 3-3).

<sup>15</sup> The GAINS model does not provide any data on NH<sub>3</sub> emissions from the cement sector. We assumed an emission factor of 20 mg/Nm<sup>3</sup> flue gas, which is within the range observed in European plants (1-26 mg/Nm<sup>3</sup>) (IPCC, 2010b). For PM emissions, the share of clinker kiln gas in total emissions is limited. Using the PM concentration data in IPCC (IPCC, 2010b), clinker kiln gas only accounts for 10 kt/yr, which is less than 20% of sectoral total emissions. We assumed that CO<sub>2</sub> capture only affects the emissions through clinker kiln gas, and emissions from other cement plant processes are unaffected.

The results are presented in Figure 3-5. The changes in net NO<sub>x</sub> emissions due to CO<sub>2</sub> capture ranged between a 40% decrease and a 30% increase compared to the reference situation. The range largely depends on the CO<sub>2</sub> capture heat supply plant option. Net SO<sub>2</sub> emissions were found to decrease by 210-350%. The uncertainty range is large mainly because the results depend on the grid electricity emission factor. As with SO<sub>2</sub>, PM emissions were also found to reduce significantly due to both onsite emissions reduction and avoided emissions from centralized power plants. Regarding NH<sub>3</sub> emissions, our bottom-up estimate on  $M_{EU,NH_3,Cem,ref}$  was about 5 kt/yr, which is considered to be in the right order of magnitude when compared to process-related emissions reported in EEA (2011; see Table 3-1). When equipped with post-combustion CO<sub>2</sub> capture, the EU-27 sector-wide NH<sub>3</sub> emissions were calculated to increase by 50-320 kt/yr as a consequence of the degradation of MEA.

### 3.5 Refinery process furnaces and boilers

In this study, CO<sub>2</sub> capture from petroleum refinery furnaces and boilers were investigated. CO<sub>2</sub> capture from existing onsite CHP and gas turbine plants was assumed to be performed separately from furnaces and heaters because CHP plants and gas turbine plants will be large enough in scale to have its own CO<sub>2</sub> capture unit. Therefore, flue gases from these existing CHP and gas turbine plants were not combined with the flue gases from furnaces and boilers and consequently not taken into account in this study.

#### 3.5.1 Reference plant specification and range of emission levels

Table 3-7 shows the reference refinery combined stack flue gas specification and the ranges of NECD substance concentrations in the flue gas investigated in this study. For refinery furnaces and boilers, specific emissions ( $Em_{i,Rf}$ ) were calculated for one tonne of CO<sub>2</sub> emitted from the reference plant (no CO<sub>2</sub> capture) instead of one tonne of crude oil processed because the fuel combustion rate in a refinery depends strongly on the degree of energy integration and complexity of the refinery (IPPC, 2010a). The combustion plant capacity in a refinery is reported to range widely from 1.7-5.4 GJ per tonne crude oil processed (IPPC, 2010a). The ranges of flue gas NECD substance concentrations represent the 5<sup>th</sup>-95<sup>th</sup> percentile values reported in IPPC (2010a). As with the cement sector, we assessed the effect of CO<sub>2</sub> capture on the NECD substance emissions for a range of concentration levels.

**Table 3-7:** Reference refinery combined stack flue gas specification and the range of NECD substance concentrations in the flue gas investigated in this study. Source: IPPC (2010a), unless otherwise stated.

<i>Parameters</i>	<i>Unit</i>	<i>Symbol</i>	<i>Value</i>
Reference plant specification			
Gas pressure	Bar		1.01
CO <sub>2</sub> concentration <sup>1)</sup>	vol%		10
O <sub>2</sub> concentration	vol%		3
Specific flue gas volume <sup>2)</sup>	10 <sup>3</sup> Nm <sup>3</sup> /tCO <sub>2</sub>	$V_{Sp,Rf}$	5.06
Range of NECD substance concentrations in the flue gas			
NO <sub>x</sub> (as NO <sub>2</sub> ) <sup>3, 4)</sup>	mg/Nm <sup>3</sup>	$C_{NOx,Rf}$	100-500
SO <sub>2</sub> <sup>5)</sup>	mg/Nm <sup>3</sup>	$C_{SO2,Rf}$	20- 1900
PM <sup>6)</sup>	mg/Nm <sup>3</sup>	$C_{PM,Rf}$	20- 700
NH <sub>3</sub> <sup>7)</sup>	mg/Nm <sup>3</sup>	$C_{NH3,Rf}$	0.1- 22

<sup>1)</sup> Based on van Straelen et al. (2010), which indicates 8-12%.

<sup>2)</sup> Based on the CO<sub>2</sub> density of 1.977kg/Nm<sup>3</sup>.

<sup>3)</sup> Based on data reported by 31 European refineries (2010a).

<sup>4)</sup> The fraction of NO<sub>2</sub> in total NO<sub>x</sub> was assumed to be 5%. NO contributes over 90% of the total NO<sub>x</sub> in most combustion processes (IPPC, 2010a).

<sup>5)</sup> Based on data reported by 34 European refineries (2010a). A survey on 67 European refineries in 2006 suggests that about 4% of total sulfur intake in these refineries has been emitted to the air in the form of SO<sub>2</sub>. Note that sulfur is sometimes added to furnaces feed to minimize coke formation in the furnace so that the operating cycles of the unit can be extended.

<sup>6)</sup> Based on the data reported by 43 European refineries (2010a). Includes all particulate sizes. The reported data in grams per tonne of crude oil processed was converted by applying specific flue gas volume of 1.0 × 10<sup>3</sup>Nm<sup>3</sup> per tonne of crude oil processed, which was taken from IPPC (2003a).

<sup>7)</sup> Based on data reported by 17 European refineries (2010a). NH<sub>3</sub> may rise from various sour water and sour gas streams or from gasifiers.

### 3.5.2 Short term CO<sub>2</sub> capture technology

The CO<sub>2</sub> concentration of flue gases from petroleum refinery combined stacks is about 8-12% (van Straelen et al., 2010). Because of the low CO<sub>2</sub> partial pressure, chemical absorption using MEA is considered to be the only feasible CO<sub>2</sub> capture option for flue gases from petroleum refinery combined stack gas in the short term (see Chapter 2). As with the cement sector (clinker production process), it was assumed that the refining process itself is unaffected by chemical absorption capture. The flue gas specifications and the technical performance of post-combustion capture are presented in Table 3-4. The need for additional flue gas pre-treatment prior to CO<sub>2</sub> capture was assessed using the scheme presented in Figure 3-2. With regard to CO<sub>2</sub> capture heat supply, the following two options were considered: 1) integrated NG-CHP (I-NG); and 2) stand-alone NG-CHP (S-NG). These CHP plants were assumed to be newly built plants and not the existing ones in the refinery because it was expected that the CO<sub>2</sub> capture energy demand would be larger than the existing plants could supply.

### 3.5.3 Results

#### 3.5.3.1 Energy and CO<sub>2</sub> balance

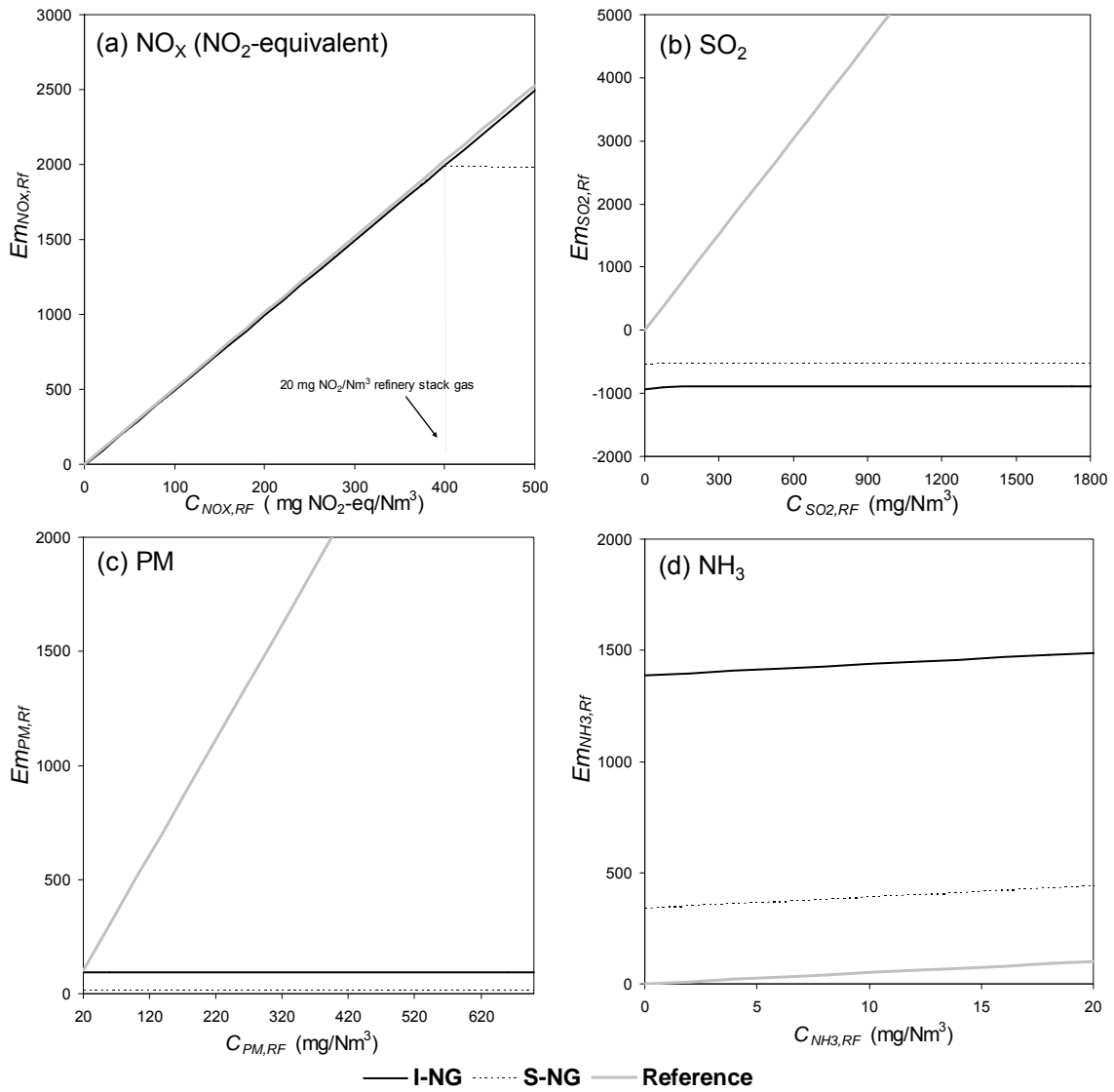
Table 3-8 shows the CO<sub>2</sub> concentration of the process flue gas at the CO<sub>2</sub> scrubber inlet, the volume of CO<sub>2</sub> captured and the specific heat requirement for CO<sub>2</sub> capture. CO<sub>2</sub> concentrations are significantly lower for the integrated CHP options because the CHP flue gas dilutes the refinery flue gas. Note that CO<sub>2</sub> from the CHP plants nearly doubles the amount of CO<sub>2</sub> captured.

**Table 3-8:** Technical results of post-combustion CO<sub>2</sub> capture from the refinery combined stack with two natural gas heat supply options.

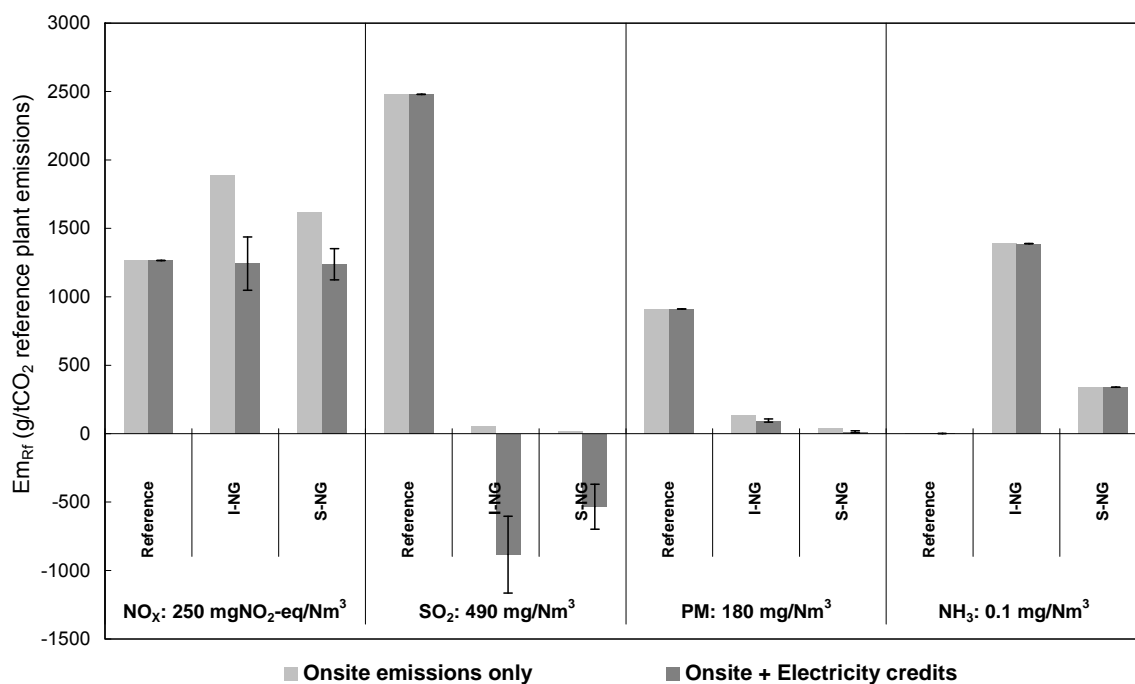
CHP configuration	Flue gas CO <sub>2</sub> concentration [v/v]	CO <sub>2</sub> capture rate [t/t reference plant emissions]	CO <sub>2</sub> capture heat requirement [GJ/tCO <sub>2</sub> captured]	Cogenerated power [GJ/tCO <sub>2</sub> reference emissions]	Excess power generation [GJ/tCO <sub>2</sub> reference emissions]	CO <sub>2</sub> avoidance rate	
						Onsite only	Incl. excess electricity credits
I-NG	4.8%	148%	4.5	5.3	4.2	74%	120%
S-NG	10%	85%	4.3	2.9	2.5	44%	71%

#### 3.5.3.2 Changes in NECD substance emissions by flue gas concentration level

Figure 3-6 shows specific emissions (g/tCO<sub>2</sub> reference emissions) of NECD substances from petroleum refinery combined stacks with and without chemical absorption CO<sub>2</sub> capture as a function of stack flue gas concentration. Figure 3-7 shows the breakdown of specific emissions by onsite emissions and emission credits for avoided centralized electricity production. The results are similar to those for the cement sector because similar CO<sub>2</sub> capture process is applied. The exception is PM emissions, which resulted in constant emission level for the range of  $C_{PM,Rf}$  assessed because additional control measures prior to CO<sub>2</sub> capture was required at all concentration levels. However, the impact of credits for avoided centralized electricity production is relatively smaller for SO<sub>2</sub> and PM, and larger for NO<sub>x</sub> compared to the case for cement plants because of the differences in pollutant concentrations observed for cement and refinery flue gases.



**Figure 3-6:** Specific emissions of NECD substances for petroleum refinery furnaces and boilers (g/tCO<sub>2</sub> reference plant emissions) with post-combustion CO<sub>2</sub> capture as a function of pollutant concentration of each NECD substance in the combined stack flue gas. NH<sub>3</sub> emissions were calculated for 490 mg/Nm<sup>3</sup> SO<sub>2</sub> and 250 mg-NO<sub>2</sub> eq./Nm<sup>3</sup> NO<sub>x</sub>. The high and low range bars show the sensitivity of the results to the solvent reboiler duty.



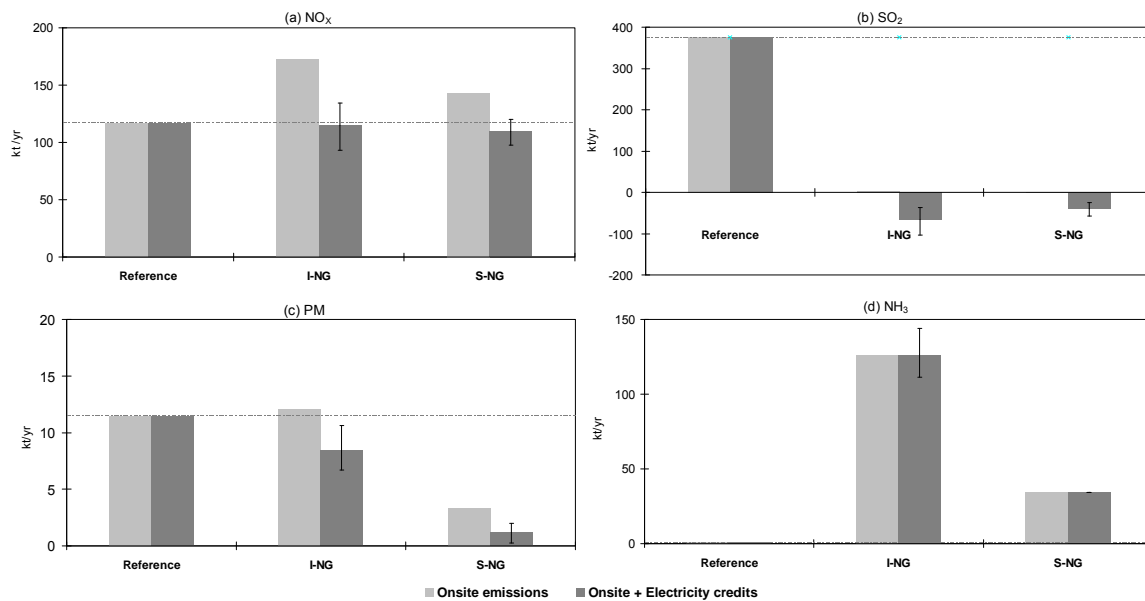
**Figure 3-7:** Breakdown of specific emissions of NECD substances for petroleum refinery furnaces and boilers (g/t reference plant CO<sub>2</sub> emissions) equipped with post-combustion CO<sub>2</sub> capture. The high and low range bars represent the uncertainty in grid electricity emission factors.

### 3.5.3.3 Changes in European sectoral emissions

The share of NECD substance emissions from furnaces and boilers in EU-27 total sectoral emissions were assumed to be 65%, equivalent to the share for CO<sub>2</sub> emissions (IEA GHG, 1999b). EU-27 sectoral emissions presented in Table 3-1 were multiplied by 65% to calculate  $M_{EU,NOX,Rf,ref}$ ,  $M_{EU,SO2,Rf,ref}$ ,  $M_{EU,PM,Rf,ref}$  and  $M_{EU,NH3,Rf,ref}$ , which are 120 kt/yr, 380 kt/yr, 11 kt/yr and 0.6 kt/yr, respectively. To calculate EU-27 sectoral NO<sub>x</sub> and SO<sub>2</sub> emissions with CO<sub>2</sub> capture ( $M_{EU,Rf,CC}$ ), we used data from around 30 European refineries (IPPC, 2010a)<sup>16</sup> that reported flue gas NECD substance concentration and total emissions for all processes within a refinery. The results obtained in the previous section were applied to the reported data on the flue gas NO<sub>x</sub> and SO<sub>2</sub> concentrations and total emissions from furnaces and boilers. For PM and NH<sub>3</sub>, there are no plant-specific data on flue gas concentration levels available. These emissions were therefore estimated by assuming for all refineries a typical CO<sub>2</sub> concentration of 10 vol%, an average PM concentration of 180 mg/Nm<sup>3</sup> (a median value observed in European refineries (IPPC, 2010a)), and an NH<sub>3</sub> concentration of 10 mg/Nm<sup>3</sup> (in agreement with the data observed in European refineries (IPPC, 2010a)).

<sup>16</sup> There are no data for total PM emissions from furnaces and boilers. Therefore, we estimated the emissions based on the CO<sub>2</sub> emissions from furnaces and boilers, assuming typical CO<sub>2</sub> concentration of 10 % (v/v) and an average PM concentration of 180 mg/Nm<sup>3</sup>, which is a median value observed for European refineries (IPPC, 2010a).





**Figure 3-8:** Annual total emissions of NECD substances from the oil refining sector in the EU-27 countries with and without CO<sub>2</sub> capture (post-combustion: MEA). The error bars represent the high and low values observed for varied heat supply options, CO<sub>2</sub> capture energy requirement and grid electricity emission factors.

The results are presented in Figure 3-8. For NO<sub>x</sub>, onsite emissions increase by 20-50%. However, this increase is offset by the emission credits for avoided centralized electricity production that total emissions reduce by up to 5%. SO<sub>2</sub> emissions decrease by more than 100% compared to the reference situation when the emission credits for avoided centralized electricity production is taken into account. PM emissions were also found to reduce, but the degree of reduction differs largely by the heat supply option. NH<sub>3</sub> emissions were found to increase by 50-120 kt/yr.

### 3.6 Iron and steel production

#### 3.6.1 Reference plant: activity levels and associated emissions

Energy and material flows in integrated iron and steel making plants are complex, making bottom-up estimates on non-CO<sub>2</sub> emissions for a reference iron and steel plant more difficult than for other sectors. For this research, we used the emissions profile from the Tata Steel IJmuiden plant in the Netherlands. Figure 3-9 shows the breakdown of annual NECD substance emissions from this plant in 2008 by process. The emissions of NECD substance from the Tata Steel IJmuiden plant were calculated based on the breakdown of emissions by process presented in Figure 3-9. Table 3-9 describes the details of NECD substance emission sources in the iron and steel making process.

**Table 3-9:** Description of emissions of NECD substances from integrated steel making plants with plant specific information from Tata Steel IJmuiden plant. Source: Corus (2009), unless otherwise stated.

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*Main emission sources and notes*

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**NO<sub>x</sub>**

More than half of total NO<sub>x</sub> emissions originate in fuel combustion in pelletizing and sintering processes. Emissions from blast furnace (BF) are mostly attributable to BF stove (IPPC, 2009a). In these processes, various mix of fuels are used (IPPC, 2009a).

**SO<sub>2</sub>**

Sulfur is emitted mainly as SO<sub>2</sub> and H<sub>2</sub>S (IPPC, 2009a). With regard to SO<sub>2</sub>, pelletizing and sintering processes account for nearly half of total plant emissions. SO<sub>2</sub> emissions from both pelletizing and sintering processes are primarily from fuel combustion (coal and oil for pelletizing, coke breeze for sintering), and to a lesser extent from iron ore (IPPC, 2009a). For BF, about half of the SO<sub>2</sub> emissions is attributable to hot stove fuel combustion and the other half to casting process and slag granulation (IPPC, 2001). Large amount of sulfur is input to BF through coal and coke, but slag absorbs the vast majority of it and most of the rest is contained in the hot metal. Gases containing H<sub>2</sub>S and SO<sub>2</sub> are generated when hot metal and slag are removed from the BF hearth and exposed to air (slag pit process) or water vapor (granulating process) (IPPC, 2009a; Pelton et al., 1974). The SO<sub>2</sub> emissions from a hot rolling plant are due to the combustion of coke oven gas (COG) and BF gas.

Note that at the Tata Steel IJmuiden plant, H<sub>2</sub>S emissions amount up to 55 g per tonne of crude steel (tcs). Moreover, BF gas, which is exported to power plants, also contains significant amount of H<sub>2</sub>S. The SO<sub>2</sub> emissions resulting from the combustion of BF gas exported to power plants amounted up to 80 g/tcs in 2007 (Corus, 2008). H<sub>2</sub>S emissions are also observed for coke ovens. H<sub>2</sub>S emissions were, however, not investigated in this study.

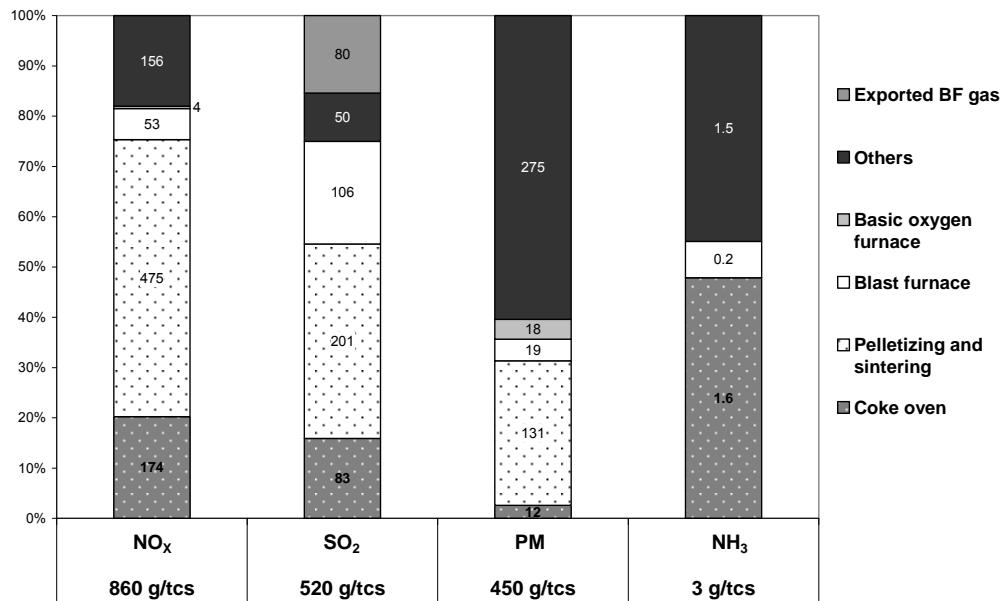
**PM**

Open source emissions account more than half of the total PM emissions from the Tata Steel IJmuiden plant. Open sources include, e.g. conveyer belts, stockpiles, and resuspension from paved roads.

**NH<sub>3</sub>**

Most of the NH<sub>3</sub> emissions from iron and steel plants are from the coke oven and hot roll mill. For coke oven, the formed NH<sub>3</sub> appears in both COG and the condensate from COG (weak liquor), with typically 20-30% of the NH<sub>3</sub> being found in the weak liquor (IPPC, 2009a). Assuming a specific coke consumption of 360 kg/t hot metal, NH<sub>3</sub> emissions from the Tata Steel plant corresponds to about 4.4 gNH<sub>3</sub>/t coke, which is on the lower side of the range observed in European plants (0.5-24.7 g/tcs. IPPC, 2009a). The NH<sub>3</sub> emissions from hot roll mill are due to NH<sub>3</sub> slip from the de-NO<sub>x</sub> unit.

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**Figure 3-9:** Breakdown of specific emissions of NECD substances per tonne of crude steel (tcs) from Tata Steel IJmuiden plant in 2008 by process (Corus, 2009)

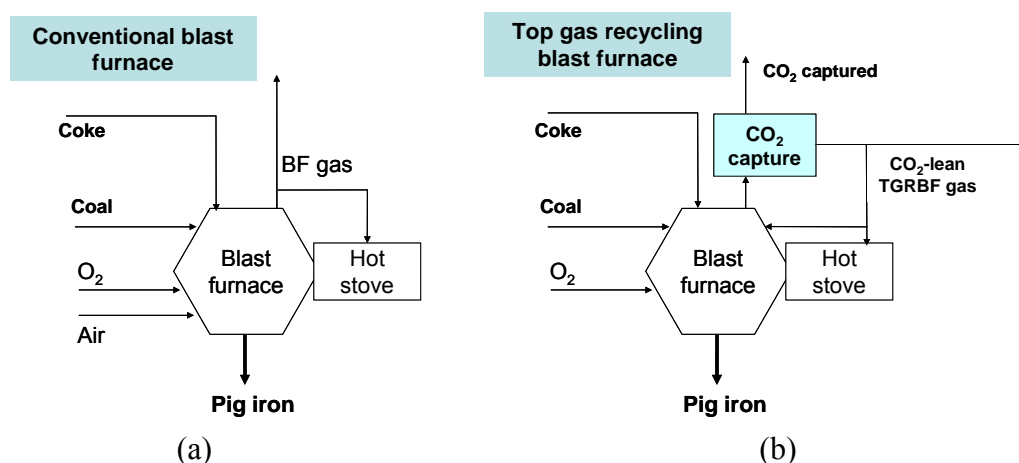
To evaluate whether the emission profile of the Tata Steel plant are representative for the EU-27 iron and steel sector, it was compared with the average emission levels in the EU-27, which were calculated based on the data presented in Section 3.2. The EU-27 average emission figures per tonne of crude steel (tcs) in 2007 were: 740 mgNO<sub>x</sub> /tonne crude steel (tcs), 820 mg SO<sub>2</sub>, 440 mg PM, and 1.3 mg NH<sub>3</sub>. Tata Steel plant data emission levels are very close to the European average for NO<sub>x</sub> and PM. For SO<sub>2</sub>, the emissions data were found to be significantly lower for the Tata Steel IJmuiden plant, while NH<sub>3</sub> emissions were higher. This is likely due to the fact that the Tata Steel IJmuiden plant is far more efficient than the benchmark energy efficiency in the iron and steel sector (Corus, 2007). Coke oven is responsible for large share for both SO<sub>2</sub> and NH<sub>3</sub> emissions and thus, the reduction in coke consumption may have lead the SO<sub>2</sub> and NH<sub>3</sub> emissions from Tata Steel IJmuiden plant to lower levels than the European average.

### 3.6.2 Short-term CO<sub>2</sub> capture technology

In a conventional integrated steelmaking plant (Figure 3-10(a)), generally around 85% of the carbon introduced into the process is present in three gas flows: approximately 70% in the blast furnace (BF) gas, 9% in the coke oven gas and 7% in the blast oxygen furnace gas (Farla et al., 1995). Therefore much research has been performed on CO<sub>2</sub> capture from the blast furnace or from the BF gas (see for example, Chapter 2 of this thesis)

In the short-mid term, Top Gas Recycling Blast Furnace (TGRBF, presented in Figure 3-10(b)) is considered to be a promising option for cost-effective CO<sub>2</sub> capture (Birat and Lorrain, 2009; Torp, 2005). The main difference of TGRBF from conventional BF is that it is oxygen-blown. Its process gas therefore contains little nitrogen and has CO<sub>2</sub> and CO

concentrations, which are reported to be 38% CO<sub>2</sub> and 47-48% CO in a modeling study (Danloy et al., 2008). These values are significantly higher than the values for conventional BF gas (17-25% CO<sub>2</sub> and 20-28% CO (IPPC, 2001)). The CO<sub>2</sub> contained in TGRBF gas can be removed by various technologies such as vacuum pressure swing adsorption (VPSA), physical absorption, or chemical absorption. TGRBF gas after CO<sub>2</sub> removal is CO-rich and is recycled back to the furnace as a reducing agent, leading to an improvement in BF performance.



**Figure 3-10:** Schematics of (a) conventional BF-based steelmaking process and (b) Top Gas Recycling Blast Furnace (TGRBF) steelmaking process

In this study, TGRBF using VPSA was considered for the analysis. This combination of technologies was tested successfully at an iron production level of 1.5 t/h (Danloy et al., 2009). Table 3-10 presents the key material inputs and outputs of a conventional BF (air-blown) and a TGRBF. The major differences in energy and material flows are the reduction in coke input and the consequent reduction in the amount of BF gas that can be used in or exported to other processes. For other processes, the volumes of intermediate products, i.e. iron ore sinter and pellets, and BOF and hot roll mill, are unaffected by TGRBF CO<sub>2</sub> capture.

TGRBF gas needs to be de-dusted for operating VPSA, but no additional equipment is necessary because BF gas is usually wet-cleaned anyway before it can be redistributed (de Beer et al., 1998). BF gas contains pollutants such as PM, cyanides, NH<sub>3</sub> and H<sub>2</sub>S, and is usually cleaned up in two stages before being used in other processes (IPPC, 2009a)<sup>17</sup>. VPSA only uses electricity for CO<sub>2</sub> removal, so there are no onsite emissions of the NECD substances from this process. Moreover, VPSA does not remove any of the NECD substances nor H<sub>2</sub>S.

<sup>17</sup> Coarse PM is removed in the first stage, and PM including zinc oxide and carbon, cyanide and NH<sub>3</sub> are removed in the second stage by wet scrubbing or wet electrostatic precipitation (IPPC, 2009a).

**Table 3-10:** Key material and energy flows per tonne of crude steel (tcs) for an integrated steel making process, with CO<sub>2</sub> capture (TGRBF with VPSA) and without CO<sub>2</sub> capture (air-blown BF). Note that the TGRBF requires a CO<sub>2</sub> capture unit for the operation. Source: Chapter 2.

<i>Input</i>	<i>Unit</i>	<i>Air blown BF</i>	<i>TGRBF+ VPSA</i>
Coke consumption	t/tcs	0.34	0.24
Coal injection to BF	t/tcs	0.18	0.18
Gas consumption in BF stove	GJ/tcs	1.5	0.72
Net gas consumption <sup>1)</sup>	GJ/tcs	-3.2	0.02
Electricity consumption	GJ/tcs	0.85	1.7
CO <sub>2</sub> capture rate	t/tcs	0	0.72
CO <sub>2</sub> avoidance rate	%-total carbon input to the reference integrated steel plant	-	45

<sup>1)</sup> Includes BF gas, coke oven gas and basic oxygen furnace gas.

Regarding the emissions from BF gas combustion, it was assumed that PM and NH<sub>3</sub> emissions from the BF stove are proportional to the stove firing capacity. The assumption implies that PM and NH<sub>3</sub> concentrations (in mg/MJ<sub>LHV</sub> heat content) are the same for the conventional BF gas and the TGRBF gas. Moreover, it was assumed that the mix of process fuel gases for internal consumption is unchanged because the process fuel gas production is barely sufficient for internal gas consumption (shown in Table 3-10). Table 3-11 summarizes the assumptions made for the calculation of non-CO<sub>2</sub> emissions from the iron and steel plant with CO<sub>2</sub> capture.

**Table 3-11:** Summary of assumptions for quantifying the effect of CO<sub>2</sub> capture (TGRBF with VPSA) on the emissions of NECD substances from an iron and steel plant.

	<i>NO<sub>x</sub></i> <sup>1)</sup>	<i>SO<sub>2</sub></i>	<i>PM</i> <sup>3)</sup>	<i>NH<sub>3</sub></i>
Coke oven	Proportional to coke consumption.			
Blast furnace	Proportional to BF stove fuel combustion <sup>1)</sup>	Hot stove <sup>4)</sup> : Proportional to stove fuel combustion. Casting process and slag granulation: Proportional to total S input to BF <sup>5)</sup>	Unaffected	Proportional to BF stove fuel consumption <sup>2)</sup>
Sinter/pellet plant	Proportional to sinter and pellet consumption <sup>3)</sup>			
BOF	Proportional to crude steel production			
Combustion of exported BF gas	Unaffected	Proportional to the amount of exported BF gas (in heat content) <sup>6)</sup>	Unaffected	Unaffected
Others	Unaffected			

<sup>1)</sup> NO<sub>x</sub> formation depends largely on the flame temperature. Although BF gas composition changes, BF gas can be used in a mix of various fuels. It was therefore assumed that the NO<sub>x</sub> formation rate per unit firing capacity is unaffected.

<sup>2)</sup> NH<sub>3</sub> emissions were assumed to be entirely from de-NO<sub>x</sub> units.

<sup>3)</sup> The assumptions for PM emissions are conservative when the open source emissions are concerned. Regarding BF and coke oven, coal and coke consumption reduces and it is therefore likely that some

fugitive emissions related to coal coke handling from BF and coke oven will be reduced. However, there is no literature that suggests to what extent these emissions will reduce.

For the emissions from BF, half of them originate in cast house, and most of the rest in raw material charging and coal preparation (IPPC, 2001). Since sinter and pellet inputs are unchanged, the assumption is reasonable.

- 4) The shares of stove fuel combustion and other processes in SO<sub>2</sub> emissions from BF were assumed to be 50% each. For the hot stove SO<sub>2</sub>, there is no information in the literature indicating a change in sulfur concentration in BFG due to the nitrogen-free operation of the BF. It was therefore assumed that the sulfur concentration of TGRBF gas per unit heat content is identical to that for conventional BF gas.
- 5) Since the vast majority of the sulfur input is absorbed by either slag or hot metal, this is a reasonable assumption. There is no information in the literature indicating a change in sulfur concentration in BF gas due to a nitrogen-free operation of the BF. Sulfur content (wt%) in coke is estimated to be 85% of that in coking coal, based on the information that generally 60-70% of total sulfur in coking coal remains in coke (Hu et al., 2008), and that 1 tonne of coking coal is converted into 0.8 tonne of coke (Diemer et al., 2004). Moreover, sulfur contents (wt%) in coking coal and non-coking coal were assumed to be the same.
- 6) There is no information in the literature indicating a change in H<sub>2</sub>S concentration in BF gas due to a nitrogen-free operation of the BF. We therefore assumed that the H<sub>2</sub>S concentration of the TGRBF gas per unit heat content is identical to that of conventional BF gas.

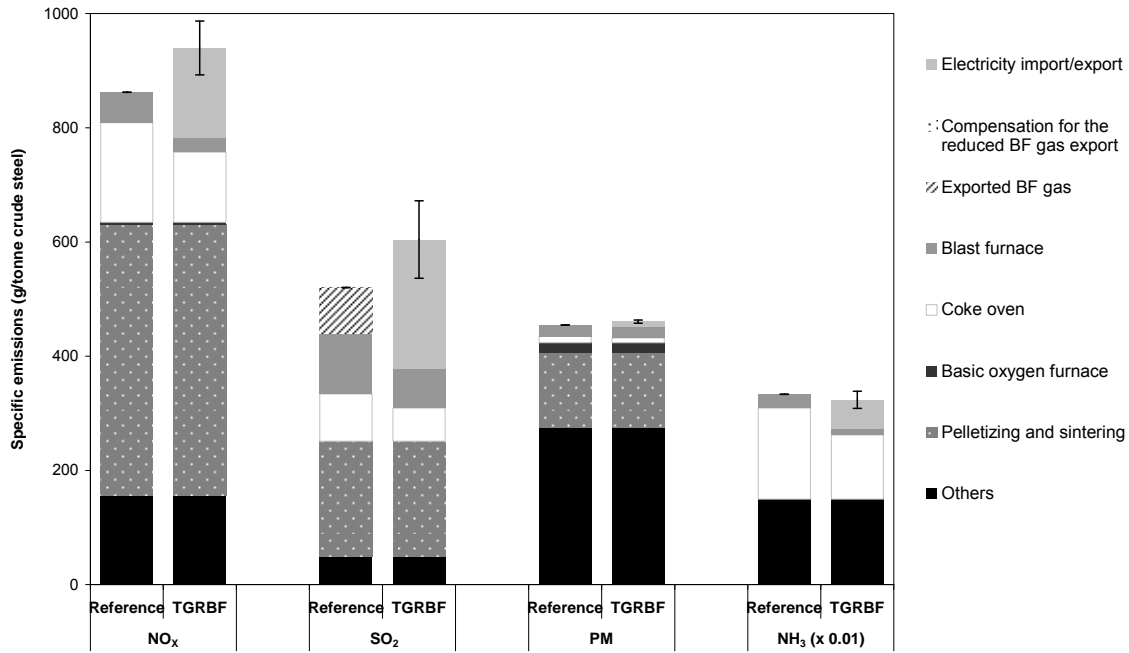
### 3.6.3 Results

Figure 3-11 shows the relative emission level of NECD substances from an iron and steel plant when CO<sub>2</sub> is captured from the blast furnace. The reduction in onsite SO<sub>2</sub> emissions is mainly due to the substitution of reduced BF gas combustion, but this is outweighed by the additional emissions due to increased electricity consumption. Onsite NO<sub>x</sub> emissions also reduce due to reduced coke consumption, but this reduction is also outweighed by the grid electricity-associated emissions. PM and NH<sub>3</sub> emissions were found to be unaffected by CO<sub>2</sub> capture.

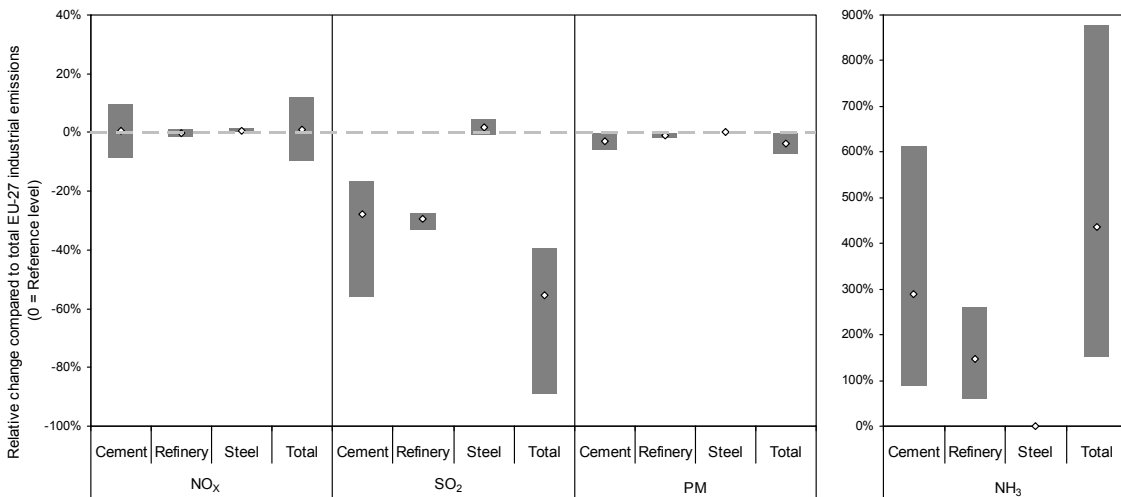
With regard to the changes in sector-wide emissions, we extrapolated the case study results for the Tata Steel IJmuiden plant to the entire EU-27 iron and steel sector. NO<sub>x</sub> and SO<sub>2</sub> emissions were found to increase by about 10 kt/yr and 30 kt/yr, respectively. The changes in PM and NH<sub>3</sub> emissions were negligible.

## 3.7 Potential impact on European industrial emissions

Figure 3-12 shows the ranges of relative changes in the emissions of NECD substances with relation to total industrial emissions in the EU-27. The bars represent the highest and lowest values observed among different heat supply options, varied grid electricity emission factors (the range is presented in Table 3-4) and varied CO<sub>2</sub> capture energy requirement ( $\pm 10\%$ ). When all three industrial sectors in the EU-27 investigated in this study were fully equipped with CO<sub>2</sub> capture, total industrial SO<sub>2</sub> emissions may decrease by 100% while total industrial NH<sub>3</sub> emissions would increase by 900%, which is about 12% of total European NH<sub>3</sub> emissions. NO<sub>x</sub> emissions may increase by up to 10% due to the cement sector when coal-fired CHP plants are used to supply CO<sub>2</sub> capture energy. The results also show that the changes in NECD substance emissions due to CO<sub>2</sub> capture will be attributable mainly to the cement and the petroleum refinery sectors.



**Figure 3-11:** NEC emissions from an iron and steel plant with top gas recycling and CO<sub>2</sub> capture using VPSA compared with a reference plant. The error bars show the sensitivity of the results to grid electricity emission factors.



**Figure 3-12:** Relative changes in NECD substance emissions from key industrial processes as a result of CO<sub>2</sub> capture compared to total industrial emissions in EU-27 in 2007. The dots represent the average values under nominal parameter sets (average values for different CO<sub>2</sub> capture energy supply options are presented for the cement and the petroleum refining sectors). The bars represent the high and low values observed for varied heat supply options, CO<sub>2</sub> capture energy requirement and grid electricity emission factors.

### 3.8 Limitations of the study

One of the main limitations of this study concerns the availability of the plant-specific emissions data for existing European plants. For the petroleum refineries, we assumed that the distribution of flue gas emission levels observed for the surveyed plants can be applied to the entire European petroleum refining sector. The uncertainty due to this assumption may affect the SO<sub>2</sub> emission results.

Data availability was particularly an issue for the iron and steel sector. Due to the specificity of process integration and the lack of plant-specific emissions figures for European steel plants, the results for the Tata Steel IJmuiden plant were extrapolated to the entire European iron and steel sector. Therefore, a detailed process analysis is necessary to obtain more reliable results for CO<sub>2</sub> capture technologies that require modification of the core of the industrial process like the TGRBF technology. Although the uncertainty of the results due to a lack of data would be limited at European sectoral level, it would be significant at plant level.

For the cement sector, there were sufficient emissions data for plants with clinker kiln (data from 258 plants of a total of 268). It was assumed that all European cement plants are of identical scale, but this does not affect the results on SO<sub>2</sub> and NH<sub>3</sub> emissions because specific emissions for these two substances are nearly constant for the range of flue gas concentrations observed for European plants. The results for NO<sub>x</sub> and PM, however, may be more sensitive than those for SO<sub>2</sub> and NH<sub>3</sub>.

Regarding system boundaries, this study took into account the NECD substance emissions associated with changes in centralized power generation due to CO<sub>2</sub> capture in the industrial sectors investigated. The results have shown that the emissions associated with grid electricity have a significant impact on total NO<sub>x</sub> and SO<sub>2</sub> emissions. However, literature indicates that other indirect emissions of NECD substances, which are outside the system boundary of this study, may also be significant. Indirect emissions from coal fuel preparation and transport is found to be as important as the direct emissions for the case of coal-fired plants with CO<sub>2</sub> capture (Koornneef et al., 2008; van Harmelen et al., 2008). These emissions should therefore be taken into account to fully understand the consequences of CO<sub>2</sub> capture on the NECD substance emissions. From the results presented in van Harmelen et al. (2008), we expect that considering secondary and tertiary emissions from coal use would result in increased emissions for chemical absorption CO<sub>2</sub> capture process using coal-fired CHP plants and reduced emissions for TGRBF steel making process.

With regard to NH<sub>3</sub> emissions results obtained in this study, it is important to note that a water-wash of CO<sub>2</sub> scrubber outlet gas was not considered as discussed previously in the introduction. For instance, a water wash can remove 95% of the NH<sub>3</sub> as well as other substances such as evaporated MEA (Veltman et al., 2010).



### 3.9 Conclusions

The objective of this study was to quantitatively assess the potential impact of CO<sub>2</sub> capture on the emission of NECD substances (PM, SO<sub>2</sub>, NO<sub>x</sub> and NH<sub>3</sub>) in Europe (EU-27) from key industrial sectors when applying CO<sub>2</sub> capture technology in the short term (timeframe: 2020). The following industrial processes were investigated: cement, petroleum refineries (furnaces and boilers), and iron and steel (integrated steel plants). We analyzed the impact on onsite emissions as well as emissions associated with the changes in centralized electricity production due to CO<sub>2</sub> capture. We investigated the changes in emissions at the plant level as well as at the EU-27 sector level. For the cement sector and for furnaces and heaters used in petroleum refineries, post-combustion CO<sub>2</sub> capture using MEA was considered. For the iron and steel sector, oxygen-blown top gas recycling blast furnace (TGRBF) with CO<sub>2</sub> capture using vacuum pressure swing adsorption (VPSA) was selected in our analysis. For post-combustion CO<sub>2</sub> capture, different solvent regeneration heat supply options were investigated. The main differences were: (1) fuel type (natural gas or coal) and (2) the treatment of CO<sub>2</sub> from the heat supply plant (vented or captured together with CO<sub>2</sub> from the industrial process).

Table 3-12 shows the summary of the results obtained in the current study. It is shown that CO<sub>2</sub> capture short-term CO<sub>2</sub> capture may have considerable impact on SO<sub>2</sub> and NH<sub>3</sub> emissions. When all three industrial processes in the EU-27 are fully equipped with CO<sub>2</sub> capture, total industrial SO<sub>2</sub> emissions in EU-27 may decrease by 90%. About half of the decrease in SO<sub>2</sub> emissions is attributable to onsite emissions reduction and the rest to avoided emissions from centralized power plants by exporting electricity to the grid. Total industrial NH<sub>3</sub> emissions may increase by 900% (about 12% of total European NH<sub>3</sub> emissions). It is, however, important to note that a water-wash of CO<sub>2</sub> scrubber outlet gas was not considered; it can remove 95% of the NH<sub>3</sub>. NO<sub>x</sub> emissions ranged between a 10% increase and a 10% decrease. The increase is mainly attributable to the cement sector when coal-fired CHP plants are used to supply CO<sub>2</sub> capture energy. The total changes in NECD substance emissions due to CO<sub>2</sub> capture in the short-term were found to be attributable to the cement and the petroleum refinery sectors; contribution of the iron and steel sector was negligible.

Regarding individual industrial processes, in the cement sector the changes in NO<sub>x</sub> emissions ranges between a 10% decrease and 10% increase compared to total EU-27 industrial emissions, largely depending on the CO<sub>2</sub> capture heat supply plant option. SO<sub>2</sub> emissions were found to decrease by 20-50% compared to total EU-27 industrial emissions. For the petroleum refineries, the reduction in SO<sub>2</sub> emissions amounts up to about 20-30% of the total EU-27 industrial SO<sub>2</sub> emissions. The changes in PM emissions were found to be marginal compared to total EU-27 industrial emissions for both sectors. The study also found that in the case of post-combustion capture for the cement sector and the petroleum refineries, avoided centralized power plant emissions, in particular SO<sub>2</sub> and PM, due to electricity export from the CHP plant are as important as changes in onsite emissions.

**Table 3-12:** Summary of the potential changes in emissions due to CO<sub>2</sub> capture from three key industrial processes relative to reference emissions (0% = reference level). The increased emissions are marked ↑ and reduced emissions are marked ↓.

<i>Sector</i>	<i>CO<sub>2</sub></i>	<i>NO<sub>x</sub></i>	<i>SO<sub>2</sub></i>	<i>PM</i>	<i>NH<sub>3</sub></i>
Cement (vs EU-27 sectoral)	60-140% ↓	40% ↓ – 30% ↑	210-350% ↓	70-160% ↓	factor 10-70 ↑
Refinery (vs EU-27 sectoral)	70-130% ↓	~5% ↓	>100% ↓	25-90% ↓	factor 50-200 ↑
Steel (vs EU-27 sectoral)	~50% ↓	10% ↑	15% ↑	No significant change	No significant change
<b>Three sectors combined (vs total EU-27 industrial)</b>	<b>25-50% ↓</b>	<b>10%↑–10%↓</b>	<b>40-90% ↓</b>	<b>0-10% ↓</b>	<b>150-880% ↑</b>

For the iron and steel sector, the effect of CO<sub>2</sub> capture was found to be limited. SO<sub>2</sub> emissions may increase by about 5% of total EU-27 industrial emissions, but the changes in other NECD substance emission levels compared to total EU-27 industrial emissions were marginal. CO<sub>2</sub> capture from blast furnace does not largely affect the most emitting processes within the integrated steel plant such as pelletizing, sintering and coking processes.

The results obtained in the current study are for specific CO<sub>2</sub> capture technologies and therefore do not necessarily apply to other CO<sub>2</sub> capture technologies. Therefore, the results presented in this paper should be handled carefully. Recommendations for future research are: (1) detailed flowsheet modeling for CO<sub>2</sub> capture technologies that integrated into the core of industrial processes for a more accurate estimation of changes in pollutant emission levels, (2) similar assessment for applying other CO<sub>2</sub> capture technologies such as oxyfuel combustion capture, post-combustion capture using advanced solvents and membranes, and (3) assessment of the economic impact of the changes in non-CO<sub>2</sub> emissions due to CO<sub>2</sub> capture.

### Acknowledgements

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## Prospects for cost-effective post-combustion CO<sub>2</sub> capture from industrial CHPs

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### Abstract

Industrial Combined Heat and Power plants (CHPs) are often operated at partial load conditions. If CO<sub>2</sub> from these plants is to be captured, additional energy requirements can be fully or partly met by increasing the load. Load increase improves plant efficiency and, consequently, part of the additional energy consumption due to CO<sub>2</sub> capture would be offset. If these advantages are large enough, industrial CHPs may become an attractive option for CCS.

The techno-economic performance of post-combustion CO<sub>2</sub> capture from small-to-medium-scale (50-200 MW<sub>e</sub> maximum electrical capacity) industrial Natural Gas Combined Cycle- (NGCC-) CHPs was compared with that from large-scale (400 MW<sub>e</sub>) NGCCs in the short term (2010) and the mid-term future (2020-2025). NGCC, CO<sub>2</sub> capture, compression, and branch CO<sub>2</sub> pipeline were investigated.

The technical results showed that the CO<sub>2</sub> capture energy requirement for industrial NGCC-CHPs is significantly lower than that for 400 MW<sub>e</sub> NGCCs: up to 16% in the short term and up to 12% in the mid-term future. The economic results showed that at low heat-to-power ratio operations, CO<sub>2</sub> capture from industrial NGCC-CHPs at 100 MW<sub>e</sub> in the short term (45-49 €/t CO<sub>2</sub>) and 200 MW<sub>e</sub> in the mid-term future (29-34 €/t CO<sub>2</sub>) may compete with CO<sub>2</sub> capture from 400 MW<sub>e</sub> NGCCs.

## 4.1 Introduction

CO<sub>2</sub> capture and storage (CCS) is attracting increasing attention as an option to reduce greenhouse gas (GHG) emissions. CCS may, in fact, play a substantial role in the smooth and cost-effective transition to a sustainable, low-carbon energy future (e.g. IPCC, 2005; Rubin et al., 2007; Turkenburg, 1992). A variety of CCS options is, in principle, already available (see e.g. Damen et al., 2006; Damen et al., 2007; IPCC, 2005). However, many integrated assessment models indicate that CCS is unlikely to be deployed in the absence of stringent policy measures that limit GHG emissions to the atmosphere (IPCC, 2005). Even under stringent policy measures, conventional low-emission technologies such as combined heat and power plants (CHP), wind turbines, and nuclear power plants may remain competitive options to gas- or coal-fired power plants with CCS when generating electricity (van den Broek et al., 2008).

CCS is currently expected to be deployed at large-scale power plants mainly due to the economies of scale. Large-scale CCS, however, may not be an effective strategy if electricity markets shift from centralized to decentralized power generation, i.e. CHPs, on-site renewable energy systems, and energy-recycling systems (WADE, 2006). Between 2004 and 2005, decentralized generation accounted for 25% of the increase in the world's total electricity production. This is a significant rise from the 13% of the increase in the world's total electricity production seen between 2001 and 2002. Moreover, this share of decentralized generation is expected to increase further in the future (WADE, 2006).

CHP is one of the most popular options for decentralized electricity generation. Because CHP is more energy efficient than the separate generation of electricity and heat, the deployment of CHP leads to lower primary energy consumption and lower CO<sub>2</sub> emissions. In recent years, most OECD countries have introduced policies to encourage decentralized CHPs (COGEN EUROPE, 2001a). Nevertheless, if society becomes further constrained by GHG emissions reduction, even CHP technologies may have to consider advanced options to reduce CO<sub>2</sub> emissions, in particular CCS. To date, only a few studies on CO<sub>2</sub> capture from CHPs have been reported in the literature. One study on CO<sub>2</sub> capture by a 5 MW<sub>e</sub> district-heating CHP plant showed that the recovery of low-grade heat from flue gas after the heat recovery steam generator (HRSG) may significantly reduce the energy penalty of CO<sub>2</sub> capture at low cost (Desideri and Corbelli, 1998). Möllersten et al. (Möllersten et al., 2006; Möllersten et al., 2004) studied a number of CO<sub>2</sub> capture options for black liquor-powered CHPs (340-510MW thermal input) in paper mills and concluded that pre-combustion CO<sub>2</sub> capture in the gasification stage of black liquor is more economical than capturing CO<sub>2</sub> from flue gas using chemical absorption techniques.

At first sight, CCS from industrial CHPs seems unattractive because these plants are often smaller in scale than centralized power plants. Moreover, industries usually require higher returns on investment costs than utilities. Nevertheless, there are potential advantages for capturing CO<sub>2</sub> from industrial CHPs. CO<sub>2</sub> capture energy requirements for industrial

CHPs may be lower than for centralized power plants. Industrial CHPs generally operate more than 8000 hours/year because of the continuous demands for process steam, but they are usually not operated at full load because the plant must be able to follow the fluctuations of the process-steam and electricity demands (Khrushch et al., 1999). The electrical conversion efficiency of a gas-fired power plant decreases at partial load operation compared to full-load operation mainly because of a decrease in the gas turbine efficiency, which is caused by the lower combustion temperature. If industrial CHPs are operated at partial load conditions, then the CO<sub>2</sub> capture energy requirements can be fully or partly met by increasing the load. Thus, an increase in fuel use due to CO<sub>2</sub> capture would improve plant efficiency and, consequently, part of the additional energy consumption due to the CO<sub>2</sub> capture would be offset. If such an advantage outweighs the aforementioned disadvantages, then industrial CHPs may become an attractive option for CCS. To our knowledge, however, no study that takes into account operational conditions has assessed the feasibility of post-combustion CO<sub>2</sub> capture from industrial CHPs.

The present study investigated the techno-economic feasibility of post-combustion CO<sub>2</sub> capture by small-to-medium-scale (50-200 MW<sub>e</sub><sup>18</sup>) industrial Natural Gas Combined Cycle-Combined Heat and Power plants (NGCC-CHPs). The main objectives of this study were: (1) to quantify CO<sub>2</sub> avoidance costs for industrial NGCC-CHPs of various scales in the short term (2010) and the mid-term future (2020-2025); and (2) to compare this system with post-combustion CO<sub>2</sub> capture by a centralized large-scale NGCC plant. The system boundaries of our study included the following components: NGCC-CHP, CO<sub>2</sub> capture, CO<sub>2</sub> compression, and branch pipeline CO<sub>2</sub> transport that connects the emission site and the trunk CO<sub>2</sub> pipeline.

This study aims to obtain comprehensive insights into the techno-economic competitiveness of post-combustion CO<sub>2</sub> capture for medium-scale industrial NGCC-CHPs. Therefore, a number of generalized relationships between plant scale and the technical and economic performance of NGCC-CHPs and CO<sub>2</sub> capture systems are used. Please note that in reality, industrial NGCC-CHPs are “tailor-made”, so that each industrial plant can meet plant-specific demands and conditions. In addition, CO<sub>2</sub> trunk pipelines are likely to be ready for short- to mid-term application only for very limited locations, e.g. industrial areas with very large CO<sub>2</sub> emissions.

Section 4.2 of this article presents an overview of industrial NGCC-CHP installations and their operational conditions, while Section 4.3 describes the methodology and assumptions used in the study. Section 4.4 presents the results of the cost calculations and compares them with the results obtained for centralized large-scale NGCC plants. The influence of various assumptions on the results is also examined. The results, research methodology,

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<sup>18</sup> Hereafter, the industrial NGCC-CHP scales are expressed in terms of maximum power generation capacity in the power-only production mode.

and uncertainties involved are discussed in Section 4.5. Lastly, the main conclusions are reported in Section 4.6.

## 4.2 Status of NGCC-CHPs in the industrial sector

On average, industrial CHPs account for 5.5% of the total electricity production of countries that together account for around 80% of the total global electricity production (IEA, 2007a). Industrial CHPs are mostly located at sites with a high demand for process heat and electricity throughout the year. In 2006, the industrial sector accounted for 75GW (or 88%) of the total CHP electrical capacity in the USA (Hedman, 2005), and in 2002, for 36 GW (or 39%) of capacity in the EU-25 (Lösönen and Danko, 2006). Typical examples can be found in the refining, pulp and paper, chemical, oil, greenhouse, and textile sectors (COGEN EUROPE, 2001b). The first three dominate the industrial CHP market and account for around 70% of the total industrial CHP electrical capacity in both the USA and the EU-25 (Hedman, 2005; Lösönen and Danko, 2006).

The global capacity of industrial NGCC-CHPs in various sectors is difficult to estimate because very little disaggregated data on CHP installations are available. The only fairly well-documented data that are public come from the USA and the Netherlands. In both countries, around half of the industrial CHP electrical capacity comes from NGCC-CHPs (Table 4-1).

**Table 4-1:** Installed capacity of industrial CHPs in the Netherlands and the USA by generator type (in GW<sub>e</sub>).

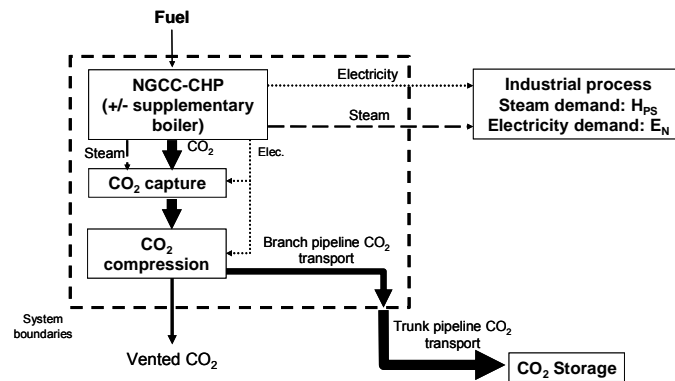
<i>Country and reference</i>	<i>Combined Cycle</i>	<i>Combustion Turbine</i>	<i>Steam Cycle</i>	<i>Gas Engine</i>	<i>Others</i>	<i>Total</i>
The Netherlands; 2005 (CBS, 2007)	2.0	0.8	0.3	0.1	0	3.2
The United States; 2004 (Hedman, 2005)	40.4	12.1	26.7	1.6	0.1	80.9

In the USA, nearly half (49%) of the industrial NGCC-CHP capacity can be found in the chemical sector and the average electrical capacity of the NGCC-CHPs in the chemical sector is around 230 MW<sub>e</sub>. In the other industrial sectors, the average electrical capacity of NGCC-CHPs are: 4 MW<sub>e</sub> in the agricultural sector, 300 MW<sub>e</sub> in primary metals production, 170 MW<sub>e</sub> in refining, 110 MW<sub>e</sub> in food sector, and 70 MW<sub>e</sub> in the paper sector (ONSITE SYCOM, 2000). The chemical industry is also the largest NGCC-CHP user in the Netherlands. In 2005, the Dutch chemical industry accounted for around 60% (1327 MW<sub>e</sub>) of the total decentralized NGCC-CHP capacity (2201 MW<sub>e</sub>) (CBS, 2007). The average electrical capacities are smaller than those in the USA, ranging between 25 and 100MW<sub>e</sub>.

### 4.3 Methodology and assumptions

#### 4.3.1 Timeframe, system description, and general assumptions

A block diagram of the NGCC-CHP system with CO<sub>2</sub> capture is shown in Figure 4-1, and included the following components: NGCC-CHP, CO<sub>2</sub> capture, CO<sub>2</sub> compression, and branch pipeline CO<sub>2</sub> transport. This study assumed that a large-scale trunk pipeline network was available to transport CO<sub>2</sub> to the storage sites. There may be cases that the industrial CHP cannot supply sufficient amount of steam for CO<sub>2</sub> capture solvent regeneration by increasing load. In such cases, a supplementary boiler was used for additional steam generation. All economic costs presented in this study are expressed in 2008 Euros. The cost data that are expressed in U.S. dollars collected from the literature are first standardized to 2008 constant U.S. dollar using Chemical Engineering Plant Cost Index (Chemical Engineering, 2010), then a €/€ currency conversion rate for the year 2008 (1€ = 1.43\$) (OANDA, 2011) was applied. The annual CEPCIs for years between 1998 through 2008 are presented in Appendix A.



**Figure 4-1:** Boundaries of the NGCC-CHP system with CO<sub>2</sub> capture, compression and branch pipeline transport investigated in this study.

#### 4.3.2 Operational cases considered in this study

Table 4-2 shows all of the operational cases for industrial NGCC-CHPs that were investigated in this study. Since one of the study’s aims was to evaluate the effect of load increase to meet the CO<sub>2</sub> capture energy requirement on the plant efficiency, the study was performed on two different cases of heat-to-power ratio (HPR) and three different cases of partial load operations. It also investigated the case in which CO<sub>2</sub> is captured from flue gas of the boiler of the supplementary boiler.

**Table 4-2:** Operational cases for industrial NGCC-CHPs without CO<sub>2</sub> capture investigated in this study.

		Fuel input rate					
		No CO <sub>2</sub> capture for boiler flue gas			CO <sub>2</sub> capture for boiler flue gas		
		60%	75%	90%	60%	75%	90%
Heat efficiency	25%	1A	1B	1C	1A*	1B*	1B*
	40%	2A	2B	2C	2A*	2B*	2B*

### 4.3.3 Component data

A literature review was performed to obtain data on the potential performance of system components in the short term (2010) and the mid-term future (2020-2025).

#### 4.3.3.1 NGCC-CHP

The main components of an NGCC-CHP are: gas turbine, HRSG, and steam turbine. In an NGCC, natural gas is first combusted with compressed air. This high-pressure combustion gas, often having a temperature above 1000°C, is then expanded in a gas turbine to generate electricity. The gas turbine exhaust gas, well above 500°C, is subsequently used to generate steam in the HRSG; this generated steam is then expanded in the steam turbine to generate additional electricity. A technical description of an NGCC can be found in, e.g., Kehlhofer et al. (1999). When an NGCC is used as an industrial CHP plant, process steam is extracted from the steam turbine.

#### *Full-load NGCC electrical efficiency*

We constructed an up-to-date curve for the relationship between plant scale and electrical conversion efficiency based on data from the Gas Turbine World (GTW) Handbook 2007-2008 (GTW, 2007; see Figure 4-2). The derived regression curve can be calculated as follows (Eq.(4.1)):

$$\eta_{CC,full-load} = 0.384 * X^{0.0619} \quad (R^2 = 0.69) \quad (4.1)$$

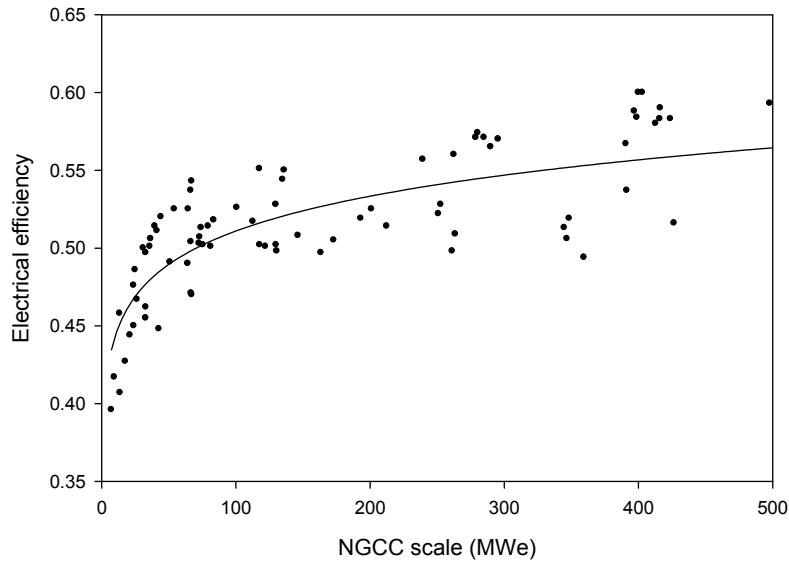
where  $\eta_{CC,full-load}$  is the full-load electrical efficiency of the NGCC and  $X$  is the capacity of the NGCC plant (MW<sub>e</sub>). The decrease in electrical conversion efficiency associated with the decrease in plant capacity is due to a lowering of the combustion temperature (Rodrigues et al., 2003).

The electrical conversion efficiency of an NGCC is expected to increase in the future. Based on an extensive literature review, Peeters et al. (2007) concluded that the electrical efficiency of large-scale NGCCs may increase by 3%-points in the mid-term future (2020) compared to 2005 NGCCs. This study assumed a 3%-point increase in NGCC electrical efficiency for all plant scales in the mid-term future compared to the short term.

#### *Partial load NGCC electrical efficiency*

The drop in electrical efficiency of an NGCC at partial load operation is mainly due to a decrease in the gas turbine efficiency, which is caused by the lower combustion temperature. The efficiency of the bottoming steam turbine, on the contrary, is hardly affected for loads above 50%, because the exhaust temperature can be maintained well above 500 °C. In order to calculate the partial load efficiency of an NGCC, a number of assumptions were made in this study. First, the gas turbine was assumed to generate two-thirds of the NGCC electricity at full load. This assumption agrees with the general trend that can be derived from data in GTW Handbook 2007-2008 (GTW, 2007). Second,





**Figure 4-2:** Electrical conversion efficiency of an NGCC as a function of capacity. Data (from GTW, 2007) as well as the regression curve that fits the data are presented.

partial load gas-turbine electrical conversion efficiency was calculated using the formula presented by Vuorinen (2007):

$$\eta_{GT,part-load} = \eta_{GT,full-load} * \frac{R_{fuel} - c}{R_{fuel} * (1 - c)} \quad (4.2)$$

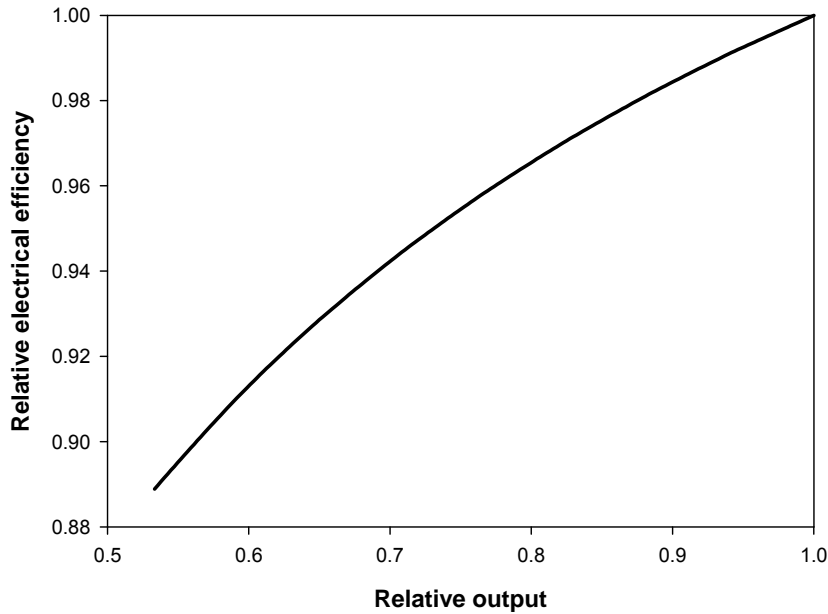
where  $\eta_{GT}$  is the electrical conversion efficiency of the gas turbine,  $R_{fuel}$  the fuel input rate ( $0.6 \leq R_{fuel} \leq 1$ ), and  $c$  the correction constant. This constant is 0.2 for a gas turbine (Vuorinen, 2007).

Third, the steam temperature at the turbine inlet is kept constant for all fuel input rates, meaning that the steam turbine efficiency is unaffected by the plant load. Please note that gas turbine CHPs in the future may have significantly higher inlet steam temperature due to improved gas turbine efficiency, leading to an increase in overall CHP efficiency. The NGCC efficiency at partial load can be determined using Eq.(4.3). The efficiency curve derived from this equation is presented in Figure 4-3. The derived partial load efficiencies were within the range found in literature (KEMA, 2006; Naqvi et al., 2007; Rolls, 2005).

$$\eta_{NGCC,part-load} = \eta_{NGCC,full-load} * \left[ 1 - \frac{2c * (1 - R_{fuel})}{3R_{fuel} * (1 - c)} \right] \quad (4.3)$$

#### Process-steam production

Process steam is extracted from the steam turbine of the NGCC-CHP. Process steam properties largely differ depending on the processes concerned. This study assumed that NGCC-CHPs supply dry, saturated steam at 10 bar (absolute) and 180°C.



**Figure 4-3:** Relative electrical efficiency of an NGCC as a function of relative electrical output derived from Equation (4.3).

This figure agrees with values used in a case study on CHPs in the Brazilian chemical industry (Szklo et al., 2004) and in a study on various industrial CHPs in the USA (ONSITE SYCOM, 2000).

The reduction in the production of electricity due to process-steam (PS) extraction from the steam turbine ( $\gamma_{PS}$ ) was assumed to be  $0.28J_e/J_{th}$  for an NGCC with 58% electrical efficiency (Bolland and Undrum, 2003).  $\gamma_{PS}$  was assumed to be proportional to the efficiency of the steam turbine of the plant. As the steam turbine accounts for one-third of the electrical conversion efficiency of the NGCC when operated at full load (and this holds for all fuel-input rates considered in this study),  $\gamma_{PS}$  can also be considered proportional to the full-load combined-cycle efficiency.

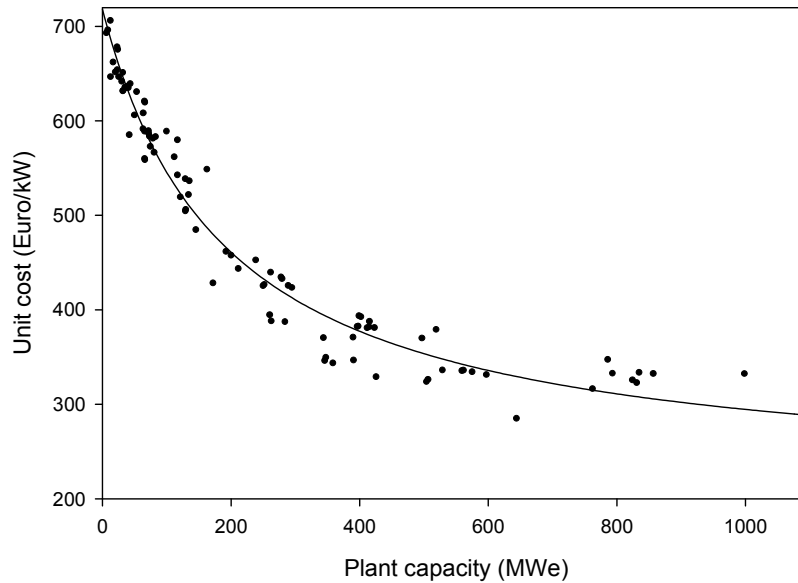
#### *Economic costs of NGCC-CHPs*

A general function describing the total equipment cost per  $\text{kW}_e$  installed as a function of NGCC capacity can be derived from the GTW Handbook 2007-2008 (GTW, 2007), using the approach of Rodrigues et al. (2003). The result is shown in Figure 4-4 and the regression curve is expressed as:

$$C_{NGCC} = \frac{718 + 1.1X}{1 + 0.0052X} \quad (R^2 = 0.97) \quad (4.4)$$

where  $C_{NGCC}$  is the total NGCC equipment cost per installed  $\text{kW}_e$  ( $\text{€}/\text{kW}_e$ ) and  $X$  is the capacity of the NGCC plant ( $\text{MW}_e$ ).

Based on an extensive literature review of future NGCC cost projections, Peeters et al. (2007) assumed a 15% cost reduction for NGCCs in 2020 compared to 2010 as a result of



**Figure 4-4:** NGCC plant equipment cost (€/kW) as a function of capacity (MW<sub>e</sub>). The regression curve is a result of the dataset in GTW (2007).

technological learning. Similarly, the present study assumed a 15% cost reduction in the mid-term future (2020-2025) compared to the short term (2010) due to technological learning. The installed cost was assumed to be 120% of the equipment cost, and the total capital requirement (TCR) was calculated using the factors presented in Peeters et al. (2007).

#### *Operation and maintenance (O&M) costs*

Total O&M costs for industrial NGCC-CHPs were assumed to be 4% of the TCR, which is in agreement with figures found in the literature (ONSITE SYCOM, 2000; Simbeck, 2001; van Oostvoorn, 2003). The O&M costs as a percentage of the TCR were assumed to be constant for all plant scales, although in practice they can be 10% lower for 200-250 MW<sub>e</sub> plants than for 40-50 MW<sub>e</sub> plants (van Oostvoorn, 2003). This percentage was within the uncertainty of the present study.

#### **4.3.3.2 Post-combustion CO<sub>2</sub> capture**

Flue gas of an NGCC is emitted at an atmospheric pressure with a CO<sub>2</sub> concentration of around 3-4%. For low CO<sub>2</sub> partial pressures, chemical absorption is the preferred method for CO<sub>2</sub> capture. Chemical absorption CO<sub>2</sub> capture systems are considered to have less operational issues compared to other CO<sub>2</sub> capture routes (pre-combustion and oxyfuel combustion), even when the steam extraction from the steam cycle is taken into account (Kvamsdal et al., 2006). This study therefore investigated the application of post-combustion CO<sub>2</sub> capture based on chemical absorption technology.

A general description of the chemical absorption of CO<sub>2</sub> can be found in, among others, the IPCC special report on CCS (IPCC, 2005). Absorption is based on the chemical reaction of CO<sub>2</sub> with an aqueous alkaline solvent. In the absorption process, flue gas is

cooled down to temperatures around 40-60°C before it comes in contact with the solvent. After absorption, the CO<sub>2</sub>-rich solvent is pumped to a stripper to remove the CO<sub>2</sub> and regenerate the solvent by heating it to 100-140°C.

#### *Energy requirements and plant scale*

The values for the reduction of electricity production per unit regeneration heat (RH) extraction ( $\gamma_{RH}$ ) in the short term and the mid-term future were extracted from Peeters et al. (2007) for an NGCC with 58% electrical efficiency (Table 4-3). Energy requirements for CO<sub>2</sub> capture using chemical solvents are assumed to be constant per unit CO<sub>2</sub> captured for all plant scales. The extraction of steam from an NGCC-CHP steam turbine for solvent regeneration leads to a decrease in electrical efficiency. The amount of decrease depends on factors such as steam temperature and pressure, and steam turbine type. Since detailed thermodynamic analysis is beyond the scope of this study, a generic approach was taken to this problem.  $\gamma_{RH}$  was assumed to be proportional to the steam-turbine efficiency of the plant and to the electrical conversion efficiency of an NGCC at full load.

**Table 4-3:** Technical performance of post-combustion CO<sub>2</sub> capture from 400 MW<sub>e</sub> NGCC of 58% electrical efficiency in the short term and the mid-term future based on Peeters et al. (2007).

<i>Parameter</i>	<i>Short-term (2010)</i>	<i>Mid-term (2020 - 2025)</i>
CO <sub>2</sub> capture efficiency (%)	90	90
Heat for regeneration (GJ/t CO <sub>2</sub> captured)	4.4	2.6
Required heat temperature (°C)	130	120
Equivalent electrical penalty (kW <sub>e</sub> /kW <sub>th</sub> )	0.203	0.186
Electricity for absorption (GJ/t CO <sub>2</sub> captured)	0.21	0.084
Electricity consumption for compression to 110 bar (GJ/t CO <sub>2</sub> captured)	0.4	0.4

#### *NGCC-CHP output limitations*

Industrial NGCC-CHPs were assumed to supply steam for CO<sub>2</sub> capture solvent regeneration by increasing the fuel input. For the efficiency calculations, it was assumed that the HPR of an NGCC-CHP does not exceed 1.5 and that the overall CHP efficiency does not exceed 90% (Bolland, 1993). If the regeneration heat requirement was not met, additional steam was generated by a boiler at 90% efficiency. This study assumed that some of the produced electricity would be sold back to the grid. Then, if there was a shortage in electricity production, consequently there would be a loss in electricity sales.

#### *Future technology development*

Several studies have been conducted on the potential future improvement of particular CO<sub>2</sub> capture technologies based on different approaches. IEA GHG (2004a) and Choi et al. (2005) assessed the performance improvement by modeling highly integrated post-

combustion CO<sub>2</sub> capture systems, while Rao et al. (2004) evaluated the performance of future amine-based CO<sub>2</sub> capture systems based on expert judgments. Others assessed the technological learning of CO<sub>2</sub> capture technologies by reviewing past experience in controlling sulfur dioxide (SO<sub>2</sub>) emissions (Riahi et al., 2004) or individual process that composes a power plant with CO<sub>2</sub> capture (Rubin et al., 2004; van den Broek et al., 2009). Moreover, a few studies investigated the cost-reduction potentials for post-combustion CO<sub>2</sub> capture as a function of time (Peeters et al., 2007; Rao et al., 2004).

For the short term and mid-term future performances of post-combustion CO<sub>2</sub> capture technology, the figures from Peeters et al. (2007) were used. In that study, various amine solvents and piperazine-potassium blend solvents were considered as possible future absorbents. It also investigated the development potential of the technology at the component level. Mid-term future technologies were defined as technologies currently in a pilot plant stadium (Peeters et al., 2007). Technical parameters on CO<sub>2</sub> capture and the values used in the present study are presented in Table 4-3. Please note that these technical parameter values are for a hypothetical solvent based on an extensive literature review, and do not represent any of the existing solvents. The ratio of electrical production reduction from regeneration heat extraction was assumed to be proportional to the maximum electrical capacity of a plant.

#### *CO<sub>2</sub> capture and compression capital costs and effects of scale*

A CO<sub>2</sub> capture system mainly consists of the following components: absorber, stripper, heat exchanger and flue gas cooler. A CO<sub>2</sub> capture system consists of one or more parallel process trains, with each process train containing all components necessary for CO<sub>2</sub> capture. The number of process trains per CO<sub>2</sub> capture system is largely influenced by the number of absorbers, and the scale of a single absorber is largely limited by the diameter of the absorber column. For an NGCC of 400 MW<sub>e</sub> scale, many studies assume two parallel absorber trains. Chapel et al. (1999) suggests a maximum single absorber capacity of around 2400 t/d recovered, which is based on a column diameter of 12.8 m. Such an absorber capacity cannot process all the flue gas from a 400 MW<sub>e</sub> NGCC, which emits more than 3500 t/d. Choi et al. (2005) also states two parallel absorber trains will be necessary for 400 MW<sub>e</sub> NGCC, if the commercial absorber diameter is to be kept below 11.5-13.1 meters to minimize potential flow maldistribution problems. Although disadvantageous with regard to economies of scale, Yokoyama (2002) argues that it is not necessarily a bad thing as multiple trains provide flexibility in the case of varying plant load.

An alternative system configuration to multiple parallel process trains consists of multiple absorbers and a single stripper, because a single stripper can process much more flue gas than an absorber of identical scale. Using the equation provided by Chapel et al. (1999), a single stripper with column diameter of 12.8 m can process nearly 10000 t/d. Such a system configuration has been proposed by Fluor and Mitsubishi Heavy Industries (MHI) for advanced post-combustion CO<sub>2</sub> capture systems for 800 MW<sub>e</sub> scale NGCC comprised of three absorbers and a single stripper (IEA GHG, 2004a).

Literature suggests there are a number of advantages for having multiple process trains for CO<sub>2</sub> capture. Aboudheir et al. (2007) argues that the concept of utilizing two CO<sub>2</sub> capture process trains is recommended for 400 MW<sub>e</sub> scale NGCC for a number of reasons:

- Two process train concept enables the construction of smaller columns that can be shop-fabricated rather than on-site fabrication of larger columns in a single process train;
- Smaller size equipment, e.g. orifice flow measuring devices, pumps, exchangers, is typically more readily available and at much lower costs as compared to custom designed equipment;
- Equipment start-up of smaller blowers, pumps and compressors in the two-train configuration puts less power demand on the NGCC power plant;
- Higher flexibility with higher turndown capacity and the ability to turn off one train if this is required by electrical demands.

In the short term, it is reasonable to assume two parallel process trains for CO<sub>2</sub> capture from the reference 400 MW<sub>e</sub> NGCC. CO<sub>2</sub> capture systems for NGCC-CHPs of 50-200 MW<sub>e</sub> scale were assumed to consist of a single process train. In a longer term, a single absorber will be able to process up to 5000-8000 t/d, although depending on the inlet flue gas CO<sub>2</sub> concentration (IEA GHG, 2004a; Rao et al., 2004; Yagi et al., 2004). MHI claims that they are ready to build a 3000 t/d single-train plant for CO<sub>2</sub> capture from natural gas combustion flue gas on a commercial basis (MHI, 2009). We therefore assumed that the CO<sub>2</sub> capture system for the reference NGCC in the mid-term future consists of a single process train. CO<sub>2</sub> capture systems for NGCC-CHPs were, as for the short term, assumed to consist of a single process train. The summary of the number of parallel process trains for CO<sub>2</sub> capture from the reference NGCC and the NGCC-CHPs are presented in Table 4-4.

**Table 4-4:** Number of parallel process trains comprising post-combustion CO<sub>2</sub> capture system (chemical absorption based) used in this study.

<i>Generator type</i>	<i>Net output (MW<sub>e</sub>)</i>	<i>Number of CO<sub>2</sub> capture trains</i>	
		Short term	Mid term
Reference NGCC	400	2	1
Industrial NGCC-CHP	50-200	1	1

#### *Maximum capacity of a single CO<sub>2</sub> compression train*

With regard to CO<sub>2</sub> compression, the maximum size of a single compressor train is suggested to be around 40 MW<sub>e</sub> (compression power). Assuming a specific power consumption of 0.4GJe/t CO<sub>2</sub>, this is equivalent to more than 8600 t/d. A single CO<sub>2</sub> compression train is therefore sufficient for NGCCs up to 400 MW<sub>e</sub> scale. Equipment cost

data were extracted from Kreutz et al.(2005). Calculation of the TCR from equipment costs was based on Peeters et al. (2007).

To calculate the capital cost of a CO<sub>2</sub> capture system for NGCC-CHPs, the following steps were taken. Firstly, the capital cost of the CO<sub>2</sub> capture and CO<sub>2</sub> compression systems for the reference 400 MW<sub>e</sub> NGCC was calculated using the spreadsheet model (with minor modifications) developed by Peeters et al. (2007). Secondly, the capital cost of CO<sub>2</sub> capture for an NGCC-CHP of  $X$  MW<sub>e</sub> scale was calculated using Eq. (4.5).

$$C_X = C_{Ref} * \frac{N_{Units,X}}{N_{Units,Ref}} * \left( \frac{S_X}{S_{Ref}} \right)^{SF} \quad (4.5)$$

where  $C$  is capital cost of the process system (CO<sub>2</sub> capture or compression),  $N$  is the number of parallel process trains per system,  $S$  is the capacity of a single process train (t/d captured) and SF is the scaling factor. An SF of 0.7 has been used for all equipment when it is needed. The influence of scaling factor will be assessment in the sensitivity analysis.

#### *Branch pipeline CO<sub>2</sub> transport*

Although transport and underground storage of CO<sub>2</sub> during enhanced oil recovery has been practised for a few decades, knowledge about its economics is still uncertain (McCollum and Ogden, 2006). Data from a number of theoretical studies on CO<sub>2</sub> transport costs differ widely because each model used its own assumptions with regard to factors like terrain and labour costs, population density, and steel prices. By standardizing key parameters, McCollum and Ogden (2006) were able to compare several techno-economic models estimating pipeline sizes and costs for CO<sub>2</sub> transport. They modified the models in such a way that the pipeline inlet and pressures, and the installation of booster stations could be selected manually. Their results showed that the variability in costs remained high, even after standardizing key parameters.

Most techno-economic models on CO<sub>2</sub> transport pipelines have been developed for trunk pipelines and not for smaller scale CO<sub>2</sub> transport volumes and pipeline lengths. The IEA-GHG PH4/6 (2002) is the only model that also enables techno-economic calculations for small branch pipelines. This model deals with CO<sub>2</sub> from small emission sources (down to 10 MW<sub>e</sub> power plants). The model outputs, however, remain to be verified.

The present study therefore compared the cost estimates from various pipeline techno-economic models for CO<sub>2</sub> transport volumes of 150-1500 t/d. These values are roughly equal the emission levels of 20 MW<sub>e</sub> and 200 MW<sub>e</sub> NGCCs, respectively. For a fair comparison, the spreadsheet model developed by McCollum and Ogden (2006) was used with some modifications. Assumptions for the calculations are presented in Table 4-5.

The terrain through which a pipeline is built is an important factor for cost estimations. According to Damen et al. (2007), construction costs in urban areas can be up to 10 times higher than in rural areas. The geographic terrain factor  $F_T$  can be used to explain the

impact of the terrain on construction costs. The  $F_T$  value that was selected in the current study was based on a terrain likely to be filled with numerous pipelines. The nominal pipeline length was set at 30 km, assuming that a trunk pipeline runs relatively near an industrial plant. The cost equation for the ANSI 900# class (gas pressure up to 140 bar absolute) in the IEA-GHG PH4/6 model was used. Table 4-5 shows the comparisons of branch CO<sub>2</sub> transport cost estimates from various models for transport mass flow rates between 150-1500 t/d. While the figure shows significant variation in cost estimates for different CO<sub>2</sub> flow rates, it also indicates that the cost estimates of IEA-GHG PH4/6 do not deviate from the cost-estimate range observed. The Chemical Engineering Plant Cost Index (Chemical Engineering, 2010) was used to include recent increases in material and manufacturing costs.

**Table 4-5:** Parameter values for CO<sub>2</sub> branch pipeline cost calculations.

<i>Parameter</i>	<i>Value</i>	<i>Parameter</i>	<i>Value</i>
Plant Capacity Factor	0.80	Gas CO <sub>2</sub> Density ( kg/Nm <sup>3</sup> )	1.965
Pipeline Inlet Pressure (bar)	110	Pipeline CO <sub>2</sub> viscosity (25 °C, 10 <sup>5</sup> N s m <sup>-2</sup> )	6.06
Pipeline Outlet Pressure (bar)	100	Economic	
Pipeline CO <sub>2</sub> Temperature (°C)	25	Electricity price (€/MWh)	60
Pipeline CO <sub>2</sub> density (25 °C, kg m <sup>-3</sup> )	884	Geographic terrain factor ( $F_T$ )	2.0

Based on these results, the approach of the IEA-GHG PH4/6 model (IEA GHG, 2002) for the branch pipeline cost calculations was used. In this model, the diameter of a pipeline is calculated using the following equation (Eq.(4.6)):

$$\Delta P = 2.252 \frac{f L \rho Q^2}{D^5} \quad (4.6)$$

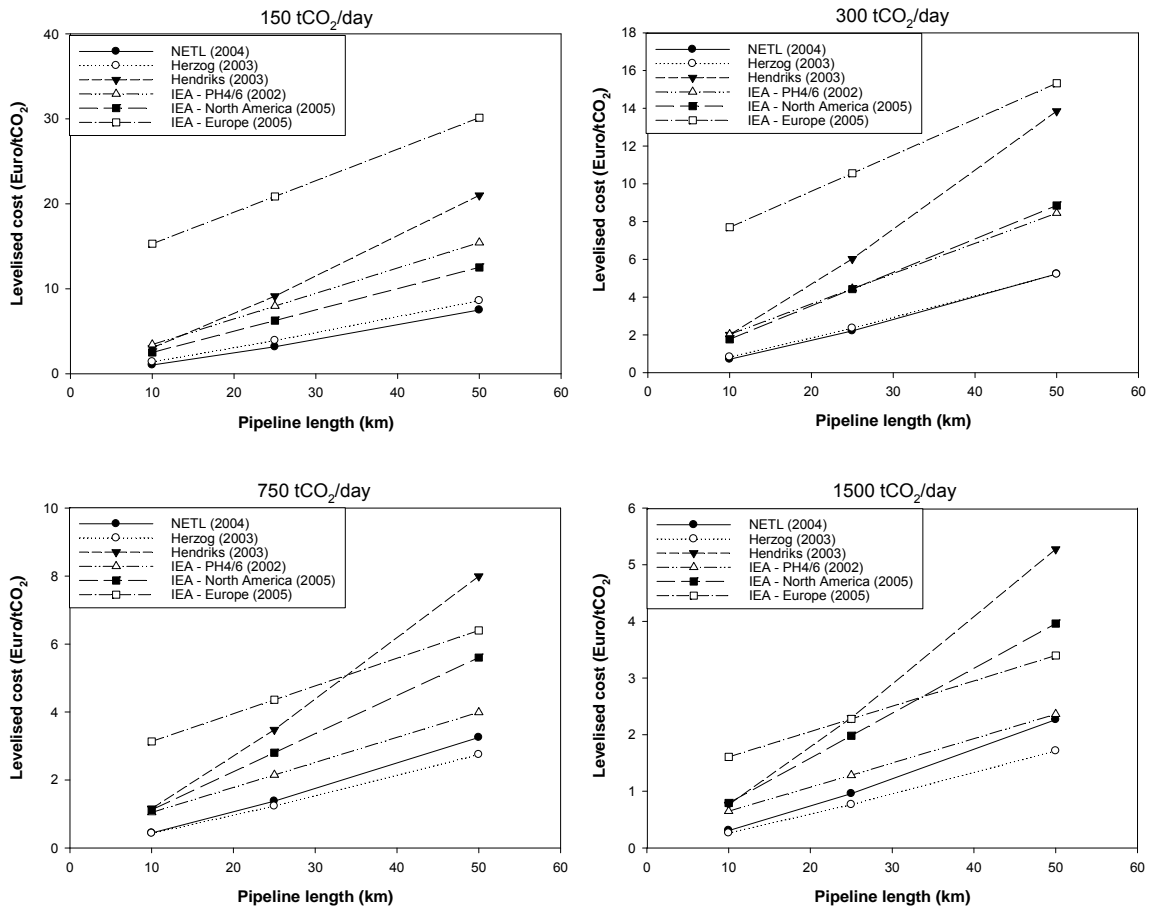
where  $D$  is the pipeline internal diameter (mm),  $\Delta P$  the pressure drop (bar),  $f$  the friction factor (0.015),  $L$  the pipeline length (km),  $\rho$  the CO<sub>2</sub> density (884 kg/m<sup>3</sup>), and  $Q$  the CO<sub>2</sub> flow rate (l/min).

The pipeline capital cost was calculated as follows:

$$C_{PL} = F_L * F_T * [(63.2L + 871) + (0.0462L + 4.02 * 10^{-4}) * D + (4.73 * 10^{-4} L + 4.75 * 10^{-5}) * D^2] \quad (4.7)$$

where  $C_{PL}$  is the pipeline capital cost (k€),  $F_L$  the location factor (1 for Europe),  $F_T$  the geographic terrain factor, and  $L$  the pipeline length (km).





**Figure 4-5:** CO<sub>2</sub> transport cost estimates from various models versus pipeline length for CO<sub>2</sub> mass flow rates 150, 750, and 1500 t/d.

#### 4.3.4 Technical and economic parameters used in the study

Table 4-6 shows the value of the other parameters used in this study. There has been a significant increase and variation in natural gas price in recent years and energy-economic models predict that the price will remain high in the timeframe considered in this study (EIA, 2009; IEA, 2008d).

##### 4.3.4.1 Performance indicators

###### *Technical indicator*

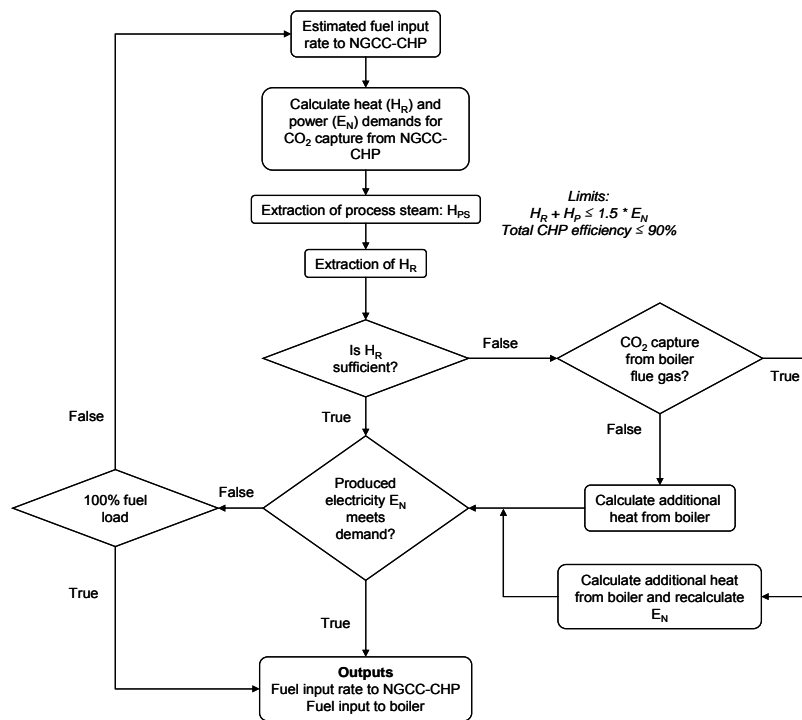
Based on the assumptions described above, it is now possible to evaluate the technical performance of industrial NGCC-CHPs with CO<sub>2</sub> capture. The hierarchy of the technical calculations is shown in Figure 4-6.

**Table 4-6:** Parameters and their values used for the calculation of technical and economic performance of post-combustion CO<sub>2</sub> capture from the reference 400 MWe NGCC and industrial NGCC-CHPs.

Parameter	Value	Range
Technical parameters		
Emission factor: Natural gas (kgCO <sub>2</sub> /GJ)	56	
Annual operation time: NGCC-CHP (hours/year)	8000	7600-8400
Distance between CHP site and CO <sub>2</sub> trunk pipeline (km)	30	-
Economic parameters		
Plant lifetime: NGCC-CHP (years)	20	-
Plant lifetime: NGCC (years)	30	-
Real interest rate: NGCC-CHP	15%	-
Real interest rate: Reference NGCC	10%	-
O&M cost: CO <sub>2</sub> pipeline (% - capital cost) <sup>(1)</sup>	2%	0.6-4%
O&M cost: other investments NGCC, CO <sub>2</sub> capture and compression (% - capital cost)	4%	4-6%
Average grid electricity carbon factor (t CO <sub>2</sub> /MWh)	0.5	-
Average electricity price (large industrial consumer, €/MWh)	40	-
Average natural gas price (incl. tax, €/GJ) <sup>(2)</sup>	7.5	3.8-11.3

<sup>(1)</sup> The range is obtained from McCollum and Ogden (2006).

<sup>(2)</sup> Our nominal value lies between the projections from IEA (2008d) and EIA (2009).



**Figure 4-6:** Hierarchy for calculating fuel consumption of industrial NGCC-CHPs with CO<sub>2</sub> capture.

Electrical efficiency of an industrial NGCC-CHP with and without CO<sub>2</sub> capture was calculated as follows (Eq.(4.8)):

$$\eta_{el,CHP,X,R_{fuel}} = \eta_{el,NGCC,X,R_{fuel}} - (\eta_{th,PS} * \gamma_{PS} - \eta_{th,RH} * \gamma_{RH}) * \frac{\eta_{el,X,full-load}}{\eta_{el,ref}} \quad (4.8)$$

where  $\eta_{el,CHP,X,R_{fuel}}$  is the electrical efficiency of  $X$  MW<sub>e</sub> NGCC-CHP at a fuel input rate of  $R_{fuel}$ ,  $\eta_{el,NGCC,X,R_{fuel}}$  the electrical efficiency of  $X$  MW<sub>e</sub> NGCC with CO<sub>2</sub> capture (%) at a fuel input rate of  $R_{fuel}$ ,  $\eta_{th,PS}$  the process steam output efficiency (%),  $\eta_{th,RH}$  the regeneration heat output efficiency (%) for CO<sub>2</sub> capture,  $\eta_{el,X,full-load}$  the electrical efficiency of  $X$  MW<sub>e</sub> NGCC at full load (%), and  $\eta_{el,ref}$  the electrical efficiency of a 400 MW<sub>e</sub> NGCC (58%).

This study compared the technical performance of CO<sub>2</sub> capture between industrial NGCC-CHPs and the reference NGCC based on the extra fuel consumption rate ( $R_x$ ).  $R_x$  for industrial NGCC-CHPs for CO<sub>2</sub> capture was calculated as:

$$R_x = \frac{F_{CHPCC} - F_{CHP} + F_B + \Delta E / \eta_{el,cent}}{F_{CHP}} \quad (4.9)$$

where  $F_{CHPCC}$  the fuel input to the NGCC-CHP with CO<sub>2</sub> capture,  $F_{CHP}$  the fuel input to the NGCC-CHP without CO<sub>2</sub> capture,  $F_B$  the fuel input to the boiler,  $\Delta E$  the amount of electricity production loss, and  $\eta_{el,cent}$  the average electrical efficiency for centralized power plants (45%).

#### Economic indicator

There are several indicators to estimate the cost performance of CO<sub>2</sub> capture. In the case of power plants, commonly used economic indicators are cost of electricity (COE) and CO<sub>2</sub> avoidance cost. In the case of CHPs, the COE largely depends on how costs are allocated to the electricity and heat that are produced. CO<sub>2</sub> avoidance cost in this study, therefore, was used as the economic indicator of post-combustion CO<sub>2</sub> capture from NGCC-CHPs. The CO<sub>2</sub> avoidance cost was calculated as follows:

$$C_{CO_2} = \frac{\alpha * (I_{CHPCC} - I_{CHP}) + (C_{e,pen} + C_{O\&M,CC})}{Em_{av}} \quad (4.10)$$

where  $C_{CO_2}$  is CO<sub>2</sub> avoidance cost (€/t CO<sub>2</sub>),  $\alpha$  the annuity factor (/year),  $I_{CHPCC}$  the initial investment for NGCC-CHP with CO<sub>2</sub> capture (incl. transport to the trunk pipeline, €),  $I_{CHP}$  the initial investment for NGCC-CHP (€),  $C_{e,pen}$  the electricity sales loss due to energy penalty of CO<sub>2</sub> capture (€),  $C_{O\&M,CC}$  the additional O&M costs for the CO<sub>2</sub> capture system (€/year), and  $Em_{av}$  the annual avoided CO<sub>2</sub> emissions (t CO<sub>2</sub>/year).

## 4.4 Results

### 4.4.1 Short term (2010)

The technical performance of post-combustion CO<sub>2</sub> capture from industrial NGCC-CHPs for low and high HPR operations in the short term are presented in Table 4-7 and Table 4-8, respectively. For low HPR operations, CO<sub>2</sub> capture energy requirements could be fully met simply by increasing fuel input unless the load rate is as high as 90%. The extra fuel input for CO<sub>2</sub> capture ( $R_x$ ) in the studied NGCC-CHPs was 10-16% lower than for the reference NGCC, with lower values for larger-scale plants and no CO<sub>2</sub> capture from boiler flue gas.

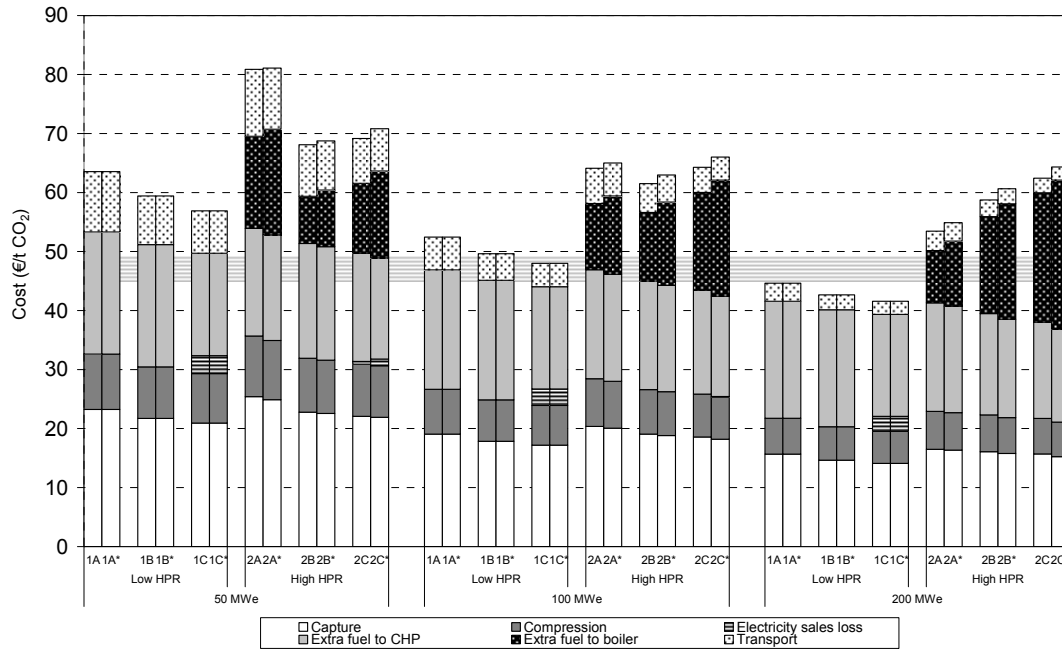
In contrast, the NGCC-CHPs for high HPR operations could only meet a small fraction of the heat demand for CO<sub>2</sub> capture because the heat output for industrial processes was already very high (Table 4-8). Because some solvent regeneration heat was produced by the boiler,  $R_x$  values were 110-170% higher for the NGCC-CHPs than for the reference NGCC. The values were especially high when CO<sub>2</sub> was captured from boiler flue gas.

CO<sub>2</sub> avoidance costs for industrial NGCC-CHPs in the short term are presented in Figure 4-7. The gray band shows the CO<sub>2</sub> avoidance cost range for the reference NGCC (400 MW<sub>e</sub>) with operation times between 6500 hours/year (49 €/t CO<sub>2</sub>) and 8500 hours/year (45 €/t CO<sub>2</sub>). The figure clearly shows that, for low HPR cases, CO<sub>2</sub> capture from NGCC-CHPs above 200 MW<sub>e</sub> can compete economically with CO<sub>2</sub> capture from the reference 400 MW<sub>e</sub> NGCC. In addition, CO<sub>2</sub> avoidance costs for these NGCC-CHPs are lower than the CO<sub>2</sub> avoidance costs for the reference NGCC operated at 8000 hours/year. Finally, 100 MW<sub>e</sub> CHPs can compete with the reference NGCC if the branch CO<sub>2</sub> transport pipeline costs can be reduced.



**Table 4-8:** Technical performance of industrial NGCC-CHPs with post-combustion CO<sub>2</sub> capture in the short term (2010) for high heat-to-power ratio (HPR) operations (operational cases 2A, 2B, and 2C).

<i>Maximum power capacity (MW<sub>e</sub>)</i>		<i>50</i>			<i>100</i>			<i>200</i>			<i>Reference</i>
<b>Operation case</b>		<i>2A</i>	<i>2B</i>	<i>2C</i>	<i>2A</i>	<i>2B</i>	<i>2C</i>	<i>2A</i>	<i>2B</i>	<i>2C</i>	<i>400MW<sub>e</sub></i>
<b>Without CO<sub>2</sub> capture</b>											
Fuel input to CHP (MW)		61	77	92	117	147	176	224	280	336	718
Electrical efficiency		34%	37%	39%	36%	38%	40%	37%	40%	42%	55.7%
Total CHP efficiency		74%	77%	79%	76%	78%	80%	77%	80%	82%	---
HPR		1.17	1.09	1.04	1.12	1.04	0.99	1.08	1.00	1.0	---
<b>With CO<sub>2</sub> capture</b>											
Fuel input to CHP (MW)		68	86	102	131	163	196	251	310	375	718
Fuel input rate		67%	84%	100%	67%	83%	100%	67%	83%	100%	100%
Heat efficiency		51%	54%	52%	53%	52%	50%	54%	51%	48%	---
Electrical efficiency		34%	36%	38%	35%	38%	40%	36%	39%	42%	48.1%
Total CHP efficiency		85%	90%	90%	88%	90%	90%	90%	90%	90%	---
HPR		1.50	1.50	1.39	1.50	1.39	1.27	1.47	1.28	1.15	---
Electricity production shortage (MW <sub>e</sub> )		0	0	0.2	0	0	0	0	0	0	---
Regeneration heat shortage (MW <sub>th</sub> )		5.1	3.5	5.9	7.3	9.4	15.3	11.2	24.6	37.4	---
Extra fuel Consumption (R <sub>x</sub> )	No CO <sub>2</sub> capture from boiler flue gas	20.1%	17.2%	18.7%	18.2%	15.8%	20.0%	17.0%	19.9%	21.3%	15.8%
	CO <sub>2</sub> capture from boiler flue gas	23.4%	19.0%	21.7%	20.6%	21.1%	24.1%	19.1%	23.7%	26.4%	



**Figure 4-7:** Breakdown of CO<sub>2</sub> avoidance costs for industrial NGCC-CHPs in the short term (2010). The gray band shows the CO<sub>2</sub> avoidance cost range for the reference 400 MW<sub>e</sub> NGCC with annual operation hours between 6500 hours/year (49 €/t CO<sub>2</sub>) and 8500 hours/year (45 €/t CO<sub>2</sub>).

The regeneration heat and electricity production shortages were found to increase CO<sub>2</sub> avoidance costs considerably. For high HPR cases, the  $R_x$  values were 20-67% higher for NGCC-CHPs than for the reference NGCC. Consequently, CO<sub>2</sub> avoidance costs for high HPR cases were 10-20 €/t CO<sub>2</sub> higher than those for low HPR cases. This makes CO<sub>2</sub> capture from industrial NGCC-CHPs with high HPR incapable of competing economically with CO<sub>2</sub> capture from the reference NGCC.

For both low and high HPR cases, CO<sub>2</sub> capture from industrial NGCC-CHPs below 50 MW<sub>e</sub> is unlikely to become economically competitive, even if the branch CO<sub>2</sub> transport pipeline cost is minimized. The results also showed that CO<sub>2</sub> capture from flue gas of the supplementary boiler did not considerably affect the CO<sub>2</sub> avoidance costs. There are some cases with electricity production losses, but the economic consequences are found to be small.

#### 4.4.2 Mid-term future (2020-2025)

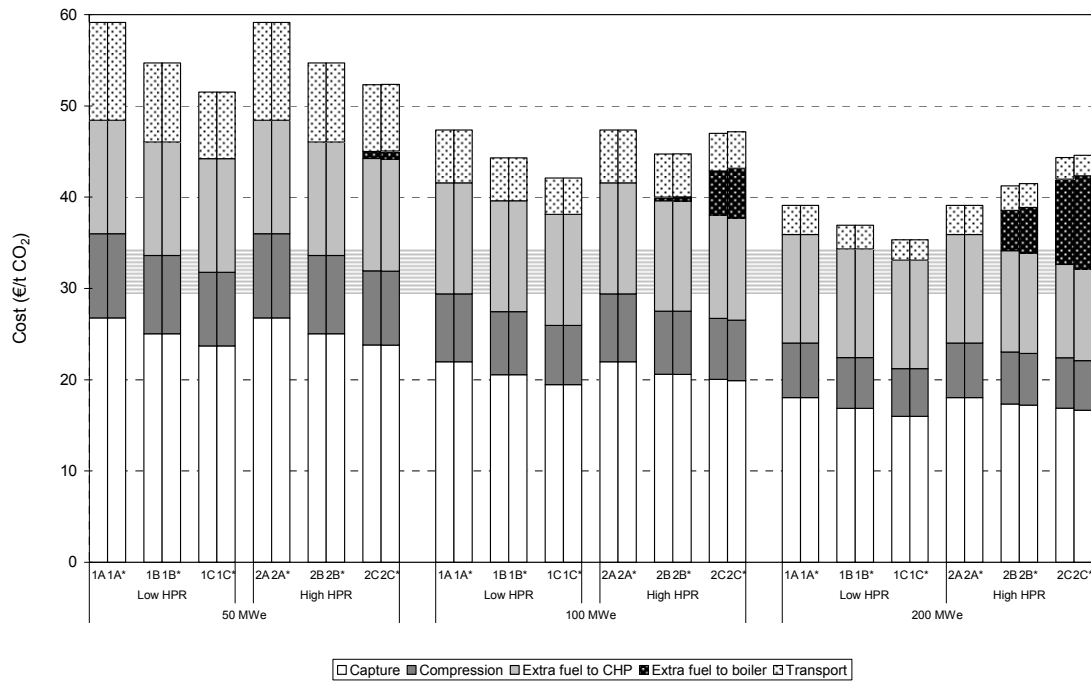
Due to reduced CO<sub>2</sub> capture energy requirements, energy demands for CO<sub>2</sub> capture can be met by an increased fuel input for most cases in the mid-term future. For low HPR operations (cases 1A, 1B, and 1C in Table 4-9), all CO<sub>2</sub> capture energy demands were covered by the NGCC-CHPs themselves.  $R_x$  values were, therefore, 8-13% lower than for the reference NGCC. The reductions in  $R_x$  were smaller than those for the short term due to lower solvent regeneration heat demand, resulting in a lower fuel input increase and a smaller partial load efficiency increase.





**Table 4-10:** Technical performance of industrial NGCC-CHPs with post-combustion CO<sub>2</sub> capture in the mid-term future (2020-2025) for high heat-to-power ratio (HPR) operations (operational cases 2A, 2B, and 2C).

Maximum power capacity (MW <sub>e</sub> )	50			100			200			Reference 400MW <sub>e</sub>	
Operation case	2A	2B	2C	2A	2B	2C	2A	2B	2C		
Without CO <sub>2</sub> capture											
Fuel input to CHP (MW)	58	72	87	111	138	166	212	265	318	682	
Electrical efficiency	36%	39%	41%	38%	41%	43%	39%	43%	45%	58.7%	
Total efficiency	76%	79%	81%	78%	81%	83%	79%	83%	85%	---	
HPR	1.17	1.02	0.98	1.06	0.98	0.94	1.02	0.94	1.17	---	
With CO <sub>2</sub> capture											
Fuel input to CHP (MW)	63	78	94	120	150	178	229	284	338	682	
Fuel input rate	65%	81%	97%	65%	81%	97%	65%	80%	96%	100%	
Heat efficiency	50%	50%	50%	50%	50%	48%	50%	48%	46%	---	
Electrical efficiency	36%	38%	40%	37%	40%	42%	39%	42%	44%	53.8%	
Total efficiency	86%	88%	90%	87%	90%	90%	89%	90%	90%	---	
HPR	1.40	1.30	1.23	1.34	1.24	1.13	1.29	1.14	1.04	---	
Electricity production shortage (MW <sub>e</sub> )	0	0	0	0	0	0	0	0	0	---	
Regeneration heat shortage (MWth)	0	0	0.2	0	0.4	4.8	0	7.6	17.6	---	
Extra fuel Consumption (R <sub>x</sub> )	No CO <sub>2</sub> capture from boiler flue gas	8.3%	8.3%	8.7%	8.1%	8.3%	10.4%	7.9%	10.0%	12.2%	9.1%
	CO <sub>2</sub> capture from boiler flue gas	---	---	8.8%	---	8.3%	11.0%	---	10.6%	13.4%	---



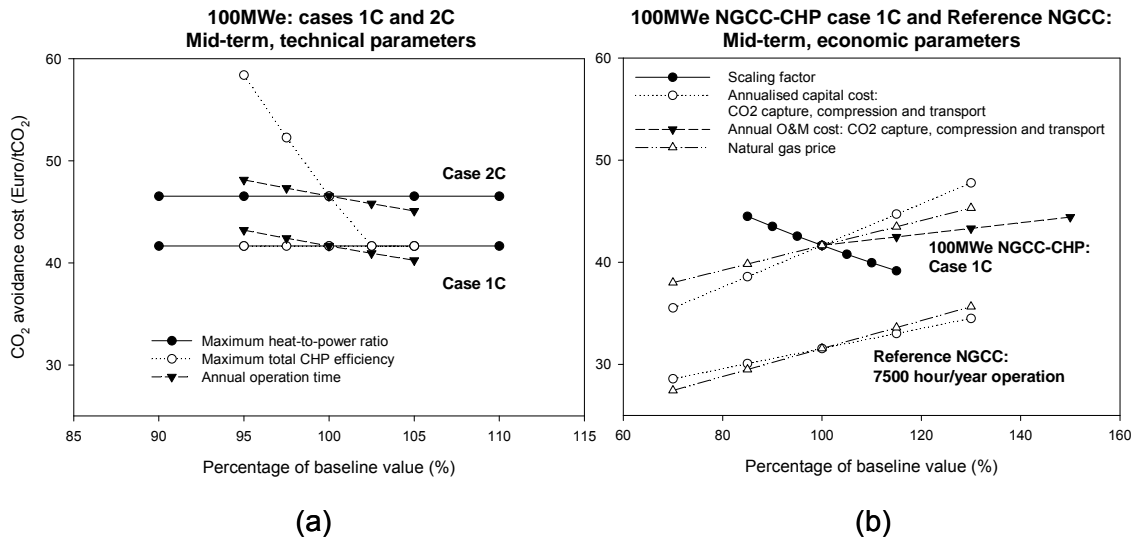
**Figure 4-8:** Breakdown of CO<sub>2</sub> avoidance costs for industrial NGCC-CHPs in the mid-term future (2020-2025). The gray band shows the CO<sub>2</sub> avoidance cost range for the reference 400 MWe NGCC with annual operation hours between 6500 hours/year (34 €/t CO<sub>2</sub>) and 8500 hours/year (29 €/t CO<sub>2</sub>).

The reduced CO<sub>2</sub> capture energy requirements may also enable industrial NGCC-CHPs operating at high HPR to fully meet the CO<sub>2</sub> capture energy requirements when the load rate is low (case 2A in Table 4-10). For case 2A, the  $R_x$  values for NGCC-CHPs running at high HPRs were 9-13% lower than for CO<sub>2</sub> capture from the reference NGCC.

The mid-term economic performance of CO<sub>2</sub> capture for industrial NGCC-CHPs is presented in Figure 4-8. The gray band shows the CO<sub>2</sub> avoidance cost range for the reference NGCC with an operation time between 6500 hours/year (34 €/t CO<sub>2</sub>) and 8500 hours/year (29 €/t CO<sub>2</sub>). The figure shows that for plants with scales above 200 MWe, cases 1A, 1B, 1C, and 2A could compete economically with CO<sub>2</sub> capture from the reference NGCC. CO<sub>2</sub> capture from industrial NGCC-CHPs for scales of 100 MWe and below is less likely to become economically competitive even if the branch pipeline length is zero ( $\geq 38$  €/t CO<sub>2</sub>). General trends in CO<sub>2</sub> avoidance costs in relation to HPR and plant scale are similar to those observed for the short term (2010).

#### 4.4.3 Sensitivity analysis

A number of parameters were used only for industrial NGCC-CHPs and not for the reference NGCC, e.g. maximum CHP total efficiency and maximum HPR parameters. Because these parameters are the main limiting factors for industrial NGCC-CHPs to meet CO<sub>2</sub> capture energy requirement, their influence on the calculated CO<sub>2</sub> avoidance costs was analyzed. The analysis was performed on the mid-term future results for 100 MWe



**Figure 4-9:** Sensitivity analysis on CO<sub>2</sub> avoidance costs of industrial NGCC-CHPs in the mid-term future (2020-2025). (a): Influence of technical parameters on 100 MW<sub>e</sub> industrial NGCC-CHPs (operational cases 1C and 2C). (b): Influence of economic parameters on 100MW<sub>e</sub> industrial NGCC-CHP (operational case 1C) and the reference 400MW<sub>e</sub> NGCC at 7500 hour/year operation time.

NGCC-CHPs for operational cases 1C and 2C. The 100 MW<sub>e</sub> scale was chosen to evaluate whether post-combustion CO<sub>2</sub> capture from NGCC-CHPs as small as 100 MW<sub>e</sub> should really not be considered competitive with the reference NGCC.

The results of the sensitivity analysis are presented in Figure 4-9. Figure 4-9(a) shows the influence of technical parameters: maximum heat-to-power ratio (HPR), maximum total CHP efficiency and the annual operation time, on CO<sub>2</sub> avoidance costs for 100 MW<sub>e</sub> NGCC-CHP with operational cases 1C and 2C. An interesting difference in sensitivity against the maximum total CHP efficiency parameter can be seen between cases 1C and 2C. This parameter had no influence on CO<sub>2</sub> avoidance costs in case 1C, because the NGCC-CHP had enough capacity to produce solvent regeneration heat for CO<sub>2</sub> capture. In contrast, the maximum total CHP efficiency parameter significantly influenced the costs in case 2C. The plotted line shows a sudden break at 105% of the baseline value, where the regeneration heat export becomes zero.

Figure 4-9(b) shows the influence of economic parameters on CO<sub>2</sub> avoidance costs for 100 MW<sub>e</sub> NGCC-CHP (operational case 1C) and for the reference NGCC. For 100 MW<sub>e</sub> NGCC-CHP, the figure shows that the influence of O&M costs for CO<sub>2</sub> capture, compression and transport is limited. Other parameters were found to influence CO<sub>2</sub> avoidance cost by up to +/- 5 €/t CO<sub>2</sub>. This result suggests that the post-combustion CO<sub>2</sub> capture from 100 MW<sub>e</sub> industrial NGCC-CHP may well be economically competitive to that from the reference NGCC.

The figure also shows that the higher the natural gas price, the smaller the CO<sub>2</sub> avoidance cost difference between 100 MW<sub>e</sub> NGCC-CHP and the reference NGCC, due to lower CO<sub>2</sub> capture energy penalty for the NGCC-CHP. The increase in annualized capital costs due to, e.g. increased initial capital investment, shorter economic lifetime and higher interest rate, will not be favorable for industrial NGCC-CHPs in comparison to the reference NGCC. This is because the fraction of annualized capital cost in CO<sub>2</sub> avoidance cost is much larger for the NGCC-CHPs than for the reference NGCC because of smaller plant scale.

## 4.5 Discussion

This study investigated the future prospects for post-combustion CO<sub>2</sub> capture from small-to-medium-scale industrial NGCC-CHPs. It is the first study in CCS research that quantifies the potential benefits of making better use of industrial CHPs at partial load operations for the purpose of CO<sub>2</sub> capture. The results show that the improvement in efficiency caused by the better use of the CHP capacity to meet CO<sub>2</sub> capture energy demands in case of long operation times can potentially outweigh the disadvantage of higher capital costs.

The study was, however, based on a number of generalized relationships between plant scale and the technical and economic performance of NGCC-CHPs and CO<sub>2</sub> capture systems. In reality, industrial NGCC-CHPs are “tailor-made” so that each industrial plant can meet plant-specific demands and conditions. Several issues concerning the limitations and uncertainties of the study are presented in this section.

### 4.5.1 Limitations to post-combustion CO<sub>2</sub> capture from industrial NGCC-CHPs

The results suggest that post-combustion CO<sub>2</sub> capture from industrial NGCC-CHPs operating at high HPR is less attractive economically than those operating at low HPR, mainly due to the limitation of the HPR flexibility of NGCCs. NGCC plants are inherently unsuitable for CHP operations with high HPR because they are built to achieve high electrical, i.e. exergetic, efficiency. The capacity for solvent regeneration is also limited because industrial NGCC-CHPs need to continuously supply industrial process steam. If an NGCC-CHP cannot supply sufficient heat for solvent regeneration, then extra fuel needs to be combusted just to generate heat. This supplementary firing leads to higher CO<sub>2</sub> capture costs.

### 4.5.2 Scale factor for CO<sub>2</sub> capture unit components

An extensive literature survey was performed on the scale factor for post-combustion CO<sub>2</sub> capture system components. Although it was assumed that a large-scale CO<sub>2</sub> capture unit consisted of multiple process trains for all time frames, it is definitely possible that a large-scale CO<sub>2</sub> capture unit with a single process train will become commercial by 2020. Literature based on interviews with experts indicates that the single process train CO<sub>2</sub> capture should be able to process 5000 t/d, which is equivalent to the amount produced by a 600-700 MW<sub>e</sub> NGCC (Rao et al., 2004; Yagi et al., 2004). Current investment costs for

single process train CO<sub>2</sub> capture systems are estimated to be no cheaper than those for multiple process train CO<sub>2</sub> capture systems (Undrum et al., 2000), but the costs may decline faster and steeper than the costs for multiple process train CO<sub>2</sub> capture systems in the future.

#### **4.5.3 Other post-combustion capture technologies**

Of the various post-combustion CO<sub>2</sub> capture technologies, this study only considered an amine-based chemical absorption method. It did not consider a number of potentially promising technologies, such as the chilled ammonia process (Rhudy, 2006), the flue gas recycling method, membrane gas absorption, and the split flow process. These technologies may result in significantly lower economic costs than amine-based solvent technology. Moreover, other post-combustion capture technologies that were considered less competitive in this study may show better heat integration than the chemical absorption method when installed in a CHP. Therefore, further study on these technologies is needed to provide more robust estimates on the techno-economic performance of post-combustion CO<sub>2</sub> capture from industrial NGCC-CHPs.

#### **4.5.4 Branch pipeline CO<sub>2</sub> transport**

This study used standardized key parameter values in order to compare small- and medium-scale pipeline CO<sub>2</sub> transport cost estimates from various models. The differences among the considered models with regard to the cost estimates for small-scale and short-distance CO<sub>2</sub> transport were significant. Besides the branch pipeline CO<sub>2</sub> transport costs, other factors can also have an impact on the uncertainty. One such factor is the terrain factor, which was set at 2 in this study. It is definitely possible that the terrain on which an industrial plant is built is full of other pipelines for gas, water, etc., which, as argued by Damen et al. (Damen et al., 2007), will increase pipeline installment costs. Another factor could be that the actual distance between an industrial plant site and the closest trunk CO<sub>2</sub> pipeline is much greater than assumed here (30 km).

#### **4.5.5 Prices of electricity exported to/ imported from the grid**

It was assumed in this study that the possible losses in net electrical output were losses in the export of electricity to the grid. It is, however, possible that the losses in net electrical output are actually an extra import of electricity from the grid. In the USA, the average price that industrial consumers pay for electricity is about twice the average price that industrial electricity producers receive when they sell their electricity to the grid (Khrushch et al., 1999). The economic consequences, therefore, may be larger. Moreover, the present study did not take into account the extra costs related to the connection of CHPs to the electricity grid (interconnection fee, stand-by fees, etc). These charges may be significant and often make smaller CHP projects less profitable (Khrushch et al., 1999).

### **4.6 Conclusions**

This study investigated the techno-economic prospects of chemical absorption-based CO<sub>2</sub> capture for industrial NGCC-CHPs of 50-200 MW<sub>e</sub>. In this study, a techno-economic

analysis was performed on NGCC-CHP, post-combustion CO<sub>2</sub> capture, and CO<sub>2</sub> branch pipeline transport for the short term (2010) and the mid-term future (2020-2025).

The technical performance results showed that the extra energy consumption for CO<sub>2</sub> capture from industrial NGCC-CHPs may be significantly lower than that for CO<sub>2</sub> capture from 400 MW<sub>e</sub> NGCCs: up to 16% lower for the short term (2010) and up to 12% for the mid-term future (2020-2025). This is mainly due to the better use of the capacity of the NGCC-CHPs, which are generally operated at partial load conditions. By increasing the fuel input rate to meet energy requirements for CO<sub>2</sub> capture, the plant efficiency is increased and, consequently, offsets part of the CO<sub>2</sub> capture energy requirements.

The economic performance results showed that CO<sub>2</sub> capture from industrial NGCC-CHPs may become economically attractive when the CHPs are operated at a low HPR. CO<sub>2</sub> avoidance costs for industrial NGCC-CHPs may become comparable to those for 400 MW<sub>e</sub> NGCCs at scales as small as 100 MW<sub>e</sub> in the short term (45-49 €/t CO<sub>2</sub>) and as small as 200 MW<sub>e</sub> in the mid-term future (29-34 €/t CO<sub>2</sub>). The results also showed that CO<sub>2</sub> avoidance costs may increase considerably if an industrial NGCC-CHP cannot self-supply the solvent regeneration heat, but needs to import part of it. In such cases, the price of the imported heat is crucial for the economic feasibility of CO<sub>2</sub> capture. One potential low-cost heat-import option may be waste heat recovered within an industrial plant. However, there are limitations in temperature and pressure that needs to be taken into account.

From the results obtained in this study, it can be concluded that post-combustion CO<sub>2</sub> capture from medium-scale industrial CHPs may become more economical than the post-combustion capture from the reference NGCC in the early stages of CCS deployment. To obtain a better understanding of the competitiveness of industrial CHPs with CCS, options for further research are: (1) to investigate other CO<sub>2</sub> capture technologies for industrial CHPs; (2) to make more robust cost estimates for industrial waste heat recovery for the purpose of CO<sub>2</sub> capture; (3) to identify the amount of industrial waste heat exploitable for CO<sub>2</sub> capture solvent regeneration in key industrial areas; and (4) to incorporate medium-scale industrial CHPs with CO<sub>2</sub> capture into national/regional energy economic models in order to gain insights into the market potential of CO<sub>2</sub> capture for industrial CHPs.

### **Acknowledgements**

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## Competitiveness of CO<sub>2</sub> capture from an industrial solid oxide fuel cell combined heat and power system in the early stage of market introduction

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### Abstract

In this article it was investigated whether potentially low-cost CO<sub>2</sub> capture from SOFC systems could enhance the penetration of solid oxide fuel cells (SOFC) in the energy market in a highly carbon-constrained society in the mid-term future (up to year 2025). The application of 5 MW<sub>e</sub> SOFC systems for industrial combined heat and power (CHP) generation was considered. For CO<sub>2</sub> capture, oxyfuel combustion of anode off-gas using commercially available technologies was selected. Gas turbine (GT-) CHP plant was considered to be the reference case.

Technical results showed that despite the energy penalties due to CO<sub>2</sub> capture and compression, net electrical and heat efficiencies were nearly identical with or without CO<sub>2</sub> capture. This was due to higher heat recovery efficiency by separating SOFC off-gas streams for CO<sub>2</sub> capture. However, CO<sub>2</sub> capture significantly increased the required SOFC and heat exchanger areas.

Economic results showed that for above 40-50 \$ t<sup>-1</sup> CO<sub>2</sub> price, SOFC-CHP systems were more economical when equipped with CO<sub>2</sub> capture. CO<sub>2</sub> capture also enabled SOFC-CHP to compete with GT-CHP at higher cell stack production costs. At zero CO<sub>2</sub> price, cell stack production cost had to be as low as 140 \$ kW<sup>-1</sup> for SOFC-CHP to outperform GT-CHP. At 100 \$ t<sup>-1</sup> CO<sub>2</sub> price, the cell stack production cost requirement raised to 350 \$ kW<sup>-1</sup>. With CO<sub>2</sub> capture, SOFC-CHP still outperformed GT-CHP at a significantly higher cell stack production cost above 900 \$ kW<sup>-1</sup>.

## 5.1 Introduction

CO<sub>2</sub> capture and storage (CCS) is an interesting option to reduce greenhouse gas (GHG) emissions that may play a substantial role for a smooth and cost-effective transition to a sustainable, low-carbon energy future (e.g. IPCC, 2005; Rubin et al., 2007; Turkenburg, 1992). CCS is currently expected to be deployed at large-scale power plants mainly due to the economies of scale of CO<sub>2</sub> capture, compression and transport. The deployment of large-scale CCS, however, becomes difficult if electricity markets shift from centralized to decentralized power generation. The share of decentralized power generation in world total power generation is steadily increasing and is expected to further increase in the future (WADE, 2006). If society becomes further constrained with GHG emissions, decentralized power technologies may also have to consider advanced options to reduce CO<sub>2</sub> emissions, especially CCS.

When CO<sub>2</sub> capture from decentralized energy conversion systems is to be considered, Solid Oxide Fuel Cell (SOFC) may become an important future technology. The 20 MW<sub>e</sub> SOFC-gas turbine (GT) hybrid system is suggested to enable CO<sub>2</sub> capture at 20-40 € t<sup>-1</sup> CO<sub>2</sub> avoided (including CO<sub>2</sub> compression and pipeline transport), on condition that the plant is located nearby the trunk CO<sub>2</sub> transport pipeline network (Damen et al., 2006; Damen et al., 2007). IEA-GHG (2007a) also indicated that CO<sub>2</sub> can be captured from a 500 kW<sub>e</sub> SOFC combined heat and power (CHP) system at 37 € t<sup>-1</sup> CO<sub>2</sub> avoided (including CO<sub>2</sub> compression) using Oxygen Conducting Membrane (OCM) afterburner.

While aforementioned advanced technologies seem promising, SOFC technology itself is not commercialized to date. SOFC is currently in the demonstration phase with the estimated system costs well above 10,000 \$ kW<sup>-1</sup> (Gerboni et al., 2008) and the potential cost reduction of SOFC being uncertain. The American Solid Energy Conversion Alliance (SECA) program sets a cell stack production cost (CSPC) target of 400 \$ kW<sup>-1</sup> (DOE, 2008) and an in-depth technical report suggests that the CSPC of 200 \$ kW<sup>-1</sup> is achievable (Thijssen, 2007). Other reports, however, are much more pessimistic. They claim that such a low production cost is too optimistic and a considerable amount of further work is necessary for SOFC systems to economically compete in the energy market (Krewitt and Schmid, 2005; Thijssen, 2007).

In this article, the following research question is addressed: Can CO<sub>2</sub> capture using currently commercial or pre-commercial technologies facilitate the market penetration of SOFC in a carbon-constrained society in the mid-term future (up to year 2025)?

Sub-questions are:

- What are the technical specifications and performances of the first-generation commercial SOFC systems?
- What are the commercially available and potentially cost-effective CO<sub>2</sub> capture options for the SOFC systems within the given timeframe?



- What is the possible CSPC requirement for SOFC without CO<sub>2</sub> capture to compete economically with conventional industrial CHP technologies under probable CO<sub>2</sub> price conditions?
- How does CO<sub>2</sub> capture affect the CSPC requirement under probable CO<sub>2</sub> price conditions?

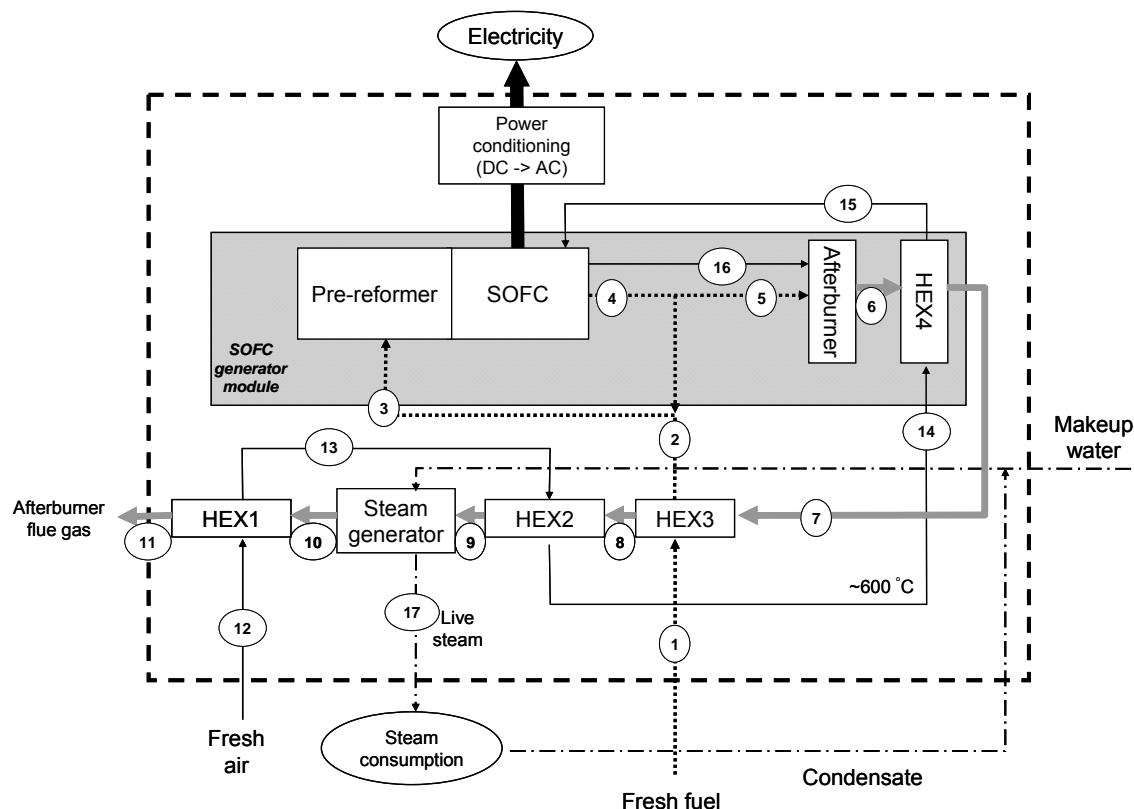
The SOFC technology has a potential for industrial CHP applications. SOFC can produce high quality heat streams because of its high operation temperatures (800-1000 °C). The long operation time of over 8000 h yr<sup>-1</sup> in industries having a continuous process steam demand, is also favorable for the economics of SOFC systems, especially in the early phase of market penetration of SOFC systems. This paper therefore focuses on SOFC systems in such industrial applications.

## 5.2 Methodology

### 5.2.1 Industrial SOFC system and its components

Figure 5-1 shows a simplified process diagram of an industrial SOFC-CHP system without CO<sub>2</sub> capture (ISOFC). The main components of the ISOFC are: fuel pre-reformer, SOFC stack, afterburner, heat recovery steam generator (HRSG), and air pre-heater. For industrial process steam generation, the heat recovery unit configuration is different from that for hot water production; the HRSG is placed between the two fresh air preheaters (van Gerwen, 2003).

Two cell types of different geometries are being pursued to date: tubular cells and planar cells. The tubular cell technology is being developed by a number of companies, e.g. Siemens Power Generation (SPG), Acumentrics, Mitsubishi Heavy Industries, and TOTO (Blum et al., 2005). Earlier designs comprised of cylindrical tubes that contain one cell and are connected either in series or in parallel. The main advantages of tubular cell geometry are easier sealing and structural integrity of the cells, which potentially outweighs the disadvantages such as lower volumetric power density due to the longer current path compared to planar cells (EG&G Technical Services Inc., 2004). The SPG's air-electrode-supported cylindrical cells show a voltage degradation of less than 0.2% per 1,000 hours for over 44,000 hours of operation (EG&G Technical Services Inc., 2004). The longest continuous operation time of the tubular cell systems to date is 69,000 hours (EG&G Technical Services Inc., 2004). Recent tubular cell developments incorporate the advantage of planar cells, i.e. shorter current paths by flattening the tubes and having integral ribs that function as bridges for current flow and also as air channels. SPC also refers to these flat tubular cells as *seal-less planar cells* (Siemens Power Corporation, 2009). The latest triangular tube cell series developed by SPC is expected to increase the power density significantly up to 0.3-0.4 W cm<sup>-2</sup> at 0.7 V (Huang, 2007; Siemens Power Generation Inc., 2008), compared to about 0.13 W cm<sup>-2</sup> at 0.7 V for cylindrical tubular cells (Zhang, 2006).



**Figure 5-1:** Simplified process flow diagram of an industrial SOFC-CHP without CO<sub>2</sub> capture (ISOFC)

The planar cell technology may become the mainstream SOFC geometry in the longer term. Although the tubular cell technology is less advanced than the tubular cells in terms of long-term stability and demonstration of plant technology, it has a significantly larger performance improvement potential (Fontell et al., 2004). The planar cell technology has been researched extensively in recent years (see, e.g., the SECA program (DOE, 2008)). The main advantage of planar cells is the higher volumetric power density compared to tubular cells due to shorter current path. Anode supported planar cells are expected to achieve power densities of 0.6-0.8 W cm<sup>-2</sup> under commercially-relevant conditions (EG&G Technical Services Inc., 2004). The main disadvantage of the planar cell geometry is the difficulty of sealing the oxidant from fuel (EG&G Technical Services Inc., 2004). In the SECA program, Versa Power Systems reported 1.7% voltage degradation per 1,000 hours for over 5,000 hours of continuous operation of a 10 kW stack (Borglum, 2009). The longest operation time achieved for planar SOFC to date is about 20,000 hours at about 1.3% voltage degradation per 1,000 hours (Blum et al., 2005).

Most developers clearly see a clear advantage in the cost reduction potential of the planar cell technology and thus, planar cells will likely be the mainstream design (Blum et al., 2005). Nevertheless, the development status of the planar cell technology is currently behind the tubular (Blum et al., 2005), and considering the timeframe defined for this study, we reckon that the advanced (flat) tubular cell geometry will remain economically competitive to the planar cell geometry. Therefore, tubular cell geometry is considered throughout this study.

### 5.2.1.1 Estimation of first-generation tubular SOFC system performance

#### *Maximum single generator module and plant scales*

The maximum single SOFC generator module developed to date is 250 kW<sub>e</sub> (Siemens Power Corporation, 2009). In the short term and mid-term future, the single generator module scale is unlikely to increase considerably. A techno-economic study on a 20 MW<sub>e</sub> pressurized SOFC-GT hybrid system (Lundberg et al., 2000) assumed that a 14 MW<sub>e</sub> SOFC generator system consists of 9 generator modules, i.e. about 1.5 MW<sub>e</sub> per generator module. Regarding the fact that Siemens has been running several tubular SOFC demonstration projects up to 250 kW<sub>e</sub> scale already, we reckon 500 kW<sub>e</sub> as a reasonable maximum single generator module size in the considered timeframe.

With regard to the maximum single plant scale, very different predictions are found in the literature. On the one hand, the SECA program plans to build multi-MW<sub>e</sub> scale proof-of-concept plants in the coming decade and eventually to build multi-hundred MW<sub>e</sub> systems in the long term (DOE, 2008). On the other hand, Krewitt and Schmid (2005) suggest a typical scale of 250 kW for tubular SOFC systems in 2030. This study subscribes to the prediction of Fontell et al. (2004), that the planar SOFC technology may achieve 0.25 MW to 1 MW by 2015, and 2 MW to 5 MW by 2020. Taking into account the fact that a 250 kW<sub>e</sub> plant has already been demonstrated, it was concluded that a 5 MW<sub>e</sub> scale SOFC systems can be achieved in the early stage of SOFC's introduction to the power market. Therefore, a 5 MW<sub>e</sub> system will be modeled in this paper for further analysis. Considering the plant scale, the SOFC system assessed in this paper is assumed to be natural gas powered.

#### *Cell power density performance*

As discussed earlier, this study assumed that the advanced flat-tube cell technology equivalent to Delta cells will be used for the first-generation commercial systems. This study assumed a conservative cell performance of 0.3 W cm<sup>-2</sup> at 0.7 V, based on experimental data (Huang, 2007; Siemens Power Generation Inc., 2008).

#### *Stack production cost estimates*

The investment costs for SPC's demonstration plants operated to date are estimated between 10,000 \$ kW<sup>-1</sup> and 20,000 \$ kW<sup>-1</sup> (Krewitt and Schmid, 2005). Cost reduction can be achieved not only by cell power density increase but also by, e.g. mass production, cell stack scale increase, and advanced cell manufacturing technologies. Thijssen (2007) reported that the manufacturing cost of a 5 kW flat-tube cell stack with a power density of 0.3 W cm<sup>-2</sup> at 0.7 V can be reduced to less than 400 \$ kW<sup>-1</sup> <sup>19</sup> by increasing the per plant annual production volume up to 250 MW. Further cost reductions were found to be limited (10-20 \$ kW<sup>-1</sup>) for larger production volumes up to 1GW yr<sup>-1</sup>. The report also

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<sup>19</sup> Hereafter all CSFC figures are given for 0.7 V cell voltage and 85% fuel utilization.

suggests that the stack manufacturing costs can be further reduced to around 200 \$ kW<sup>-1</sup> by increasing the cell stack scale from 5 kW to 2 MW. On the other hand, stack production costs can be nearly 2,000 \$ kW<sup>-1</sup> for production volumes as small as 5 MW per year (Thijssen, 2007).

### 5.2.1.2 CO<sub>2</sub> capture for SOFC systems

#### *Overview of commercially available CO<sub>2</sub> capture technologies for natural gas fed SOFC systems*

SOFC systems accept fuels other than hydrogen, i.e. CO and other hydrocarbons such as CH<sub>4</sub>. The water gas shift reaction involving CO and the steam reforming of CH<sub>4</sub> in the high temperature environment of SOFC produce hydrogen that is easily oxidized at the anode (EG&G Technical Services Inc., 2004). This means that SOFC functions as a CO<sub>2</sub> concentrator when it is fed with carbon fuels. SOFC oxidizes 80-90% of the gaseous fuel input to the system, making the anode off-gas comprised mostly of H<sub>2</sub>O and CO<sub>2</sub>, with some unutilized fuel (H<sub>2</sub> and CO) and little nitrogen. To take advantage of this nature of SOFC, two post-fuel cell CO<sub>2</sub> capture options have been attracting more attention. The first option is oxyfuel combustion, i.e. to oxidize the remaining fuel with pure oxygen, and the second option is to remove CO<sub>2</sub> or the remaining fuel from the anode off-gas. After CO<sub>2</sub> removal, the remaining fuel contained in the anode off-gas is recycled for heat or electricity production.

In this section, CO<sub>2</sub> capture from gas-powered SOFC systems is further discussed. Table 5-1 presents an overview of post-fuel cell CO<sub>2</sub> capture technologies proposed for gas-powered SOFC systems in the literature. Because of the modular nature of SOFC, some of the proposed CO<sub>2</sub> capture techniques are based on the application of membrane technologies that may enable good integration with SOFC. Note that coal-powered integrated gasification SOFC systems with CO<sub>2</sub> capture are also being extensively researched (e.g., Grol et al., 2007; Li et al., 2010; Lisbona and Romeo, 2008; Parsons et al., 2002; Słowiński, 2006).

The number of technologies that can potentially offer low-cost CO<sub>2</sub> capture for gas-powered SOFC systems in the mid-term future, however, is limited for several reasons. Firstly, many of the potentially low-cost CO<sub>2</sub> capture technologies for multi-MW<sub>e</sub> scale SOFC systems are not mature. Most of the membrane gas separation technologies are currently in the laboratory phase and it is unlikely that they are commercialized in the considered timeframe. Secondly, many of the commercially available CO<sub>2</sub> capture technologies are generally for large-scale applications, and will likely be prohibitively expensive for the plant scale considered in this study, i.e. 5 MW<sub>e</sub>, due to the economies of scale. Thirdly, some of the CO<sub>2</sub> capture technologies may not be suitable for SOFC systems for CHP applications. An example is to remove CO<sub>2</sub> from anode off-gas after shift reaction. The H<sub>2</sub> separated from CO<sub>2</sub> can be recycled back to the anode (Carson, 1995) or combusted (Campanari, 2002). For CHP plants, the benefit of generating

additional hydrogen may not be maximized because the maximization of electrical efficiency is not necessary.

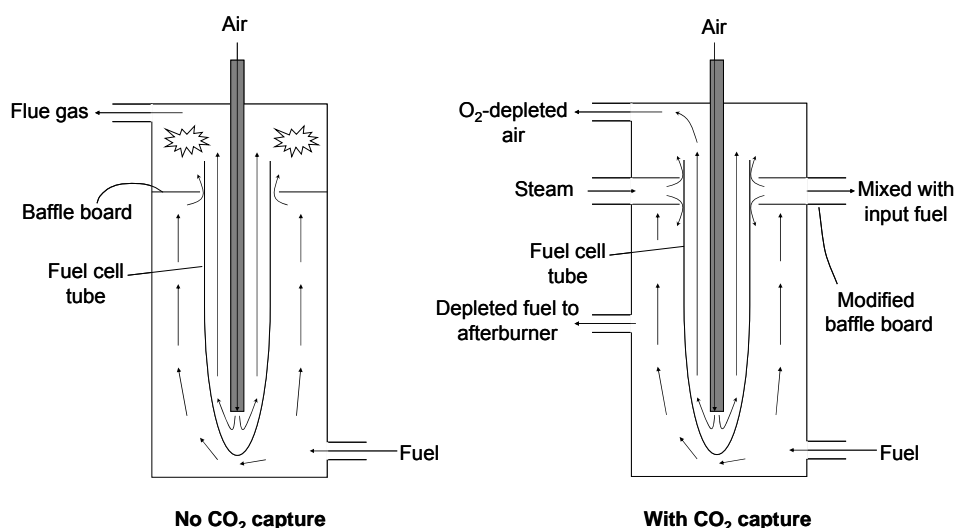
**Table 5-1:** Overview of post-fuel cell CO<sub>2</sub> capture technologies for gas-powered SOFC systems proposed in the literature.

Technology		Separated gas	Maturity of components
Oxyfuel combustion of anode off-gas	Small scale ASU + separate afterburner (Haije et al., 2006; Lygre et al., 2001)	O <sub>2</sub>	PSA/VPSA: commercial OCM and oxyfuel afterburner: near-commercial <sup>1)</sup>
	Additional SOFC <sup>2)</sup> (Maurstad et al., 2004) Oxygen conducting membrane (OCM) integrated afterburner (Maurstad et al., 2004)	O <sub>2</sub>	Laboratory Laboratory
CO <sub>2</sub> separation from anode off-gas	Without water-gas shift (WGS)	CO <sub>2</sub>	Cryogenic separation (Namie et al., 1993) Commercial
			Chemical absorption (Campanari, 2002) Commercial
			Physical absorption Calcium oxide (Vivanpatarakij et al., 2009) Laboratory (Alonso et al., 2010)
	With WGS	H <sub>2</sub>	Pressure swing adsorption (PSA) (Carson, 1995) Conventional WGS: commercial PSA: commercial
		CO <sub>2</sub>	Physical absorption Water-gas shift membrane reactor (WGSMR) (Kvamsdal et al., 2006) Laboratory (Kvamsdal et al., 2006)
Post-combustion CO <sub>2</sub> capture	Chemical absorption (Möller et al., 2004) Commercial	CO <sub>2</sub>	
	Calcium oxide (Vivanpatarakij et al., 2009) Laboratory (Alonso et al., 2010)		

<sup>1)</sup> Air Products report that the testing of a 5 tO<sub>2</sub> d<sup>-1</sup> OCM unit was successful under full driving force conditions (Stiegel, 2008). Oxyfuel combustion of fuel gases is currently in the pilot phase, with 20 MW<sub>th</sub> plant being successfully tested for CH<sub>4</sub> around 2002-2003 (Anderson et al., 2008).

<sup>2)</sup> The additional (or second) SOFC module is located downstream of the main SOFC and it oxidizes all the fuel that were not oxidized by the main SOFC.

This study selected the oxyfuel combustion CO<sub>2</sub> capture option using a conventional ASU and a separate afterburner for further analysis. The technical performance of this option has been investigated by Lygre et al. (2001) for the cogeneration of hot water and power (500 kW<sub>e</sub>). Regarding commercial small-scale oxygen generation technologies, (vacuum-) pressure swing adsorption (PSA/VPSA) technologies continue to reduce costs. They are expected to further increase the scale at which they will breakeven in product cost with cryogenic separation technology for the similar oxygen purity (Castle, 2002). Therefore, this option may enable low-cost CO<sub>2</sub> capture using commercially available technologies.



**Figure 5-2:** SOFC off-gas flow manifolding. Left: SOFC without CO<sub>2</sub> capture; Right: SOFC with CO<sub>2</sub> capture; generator plenum and cathode off-gas plenum is separated by the steam plenum, which has slightly higher pressure than the anode and cathode off-gases (Draper and Veyo, 2002).

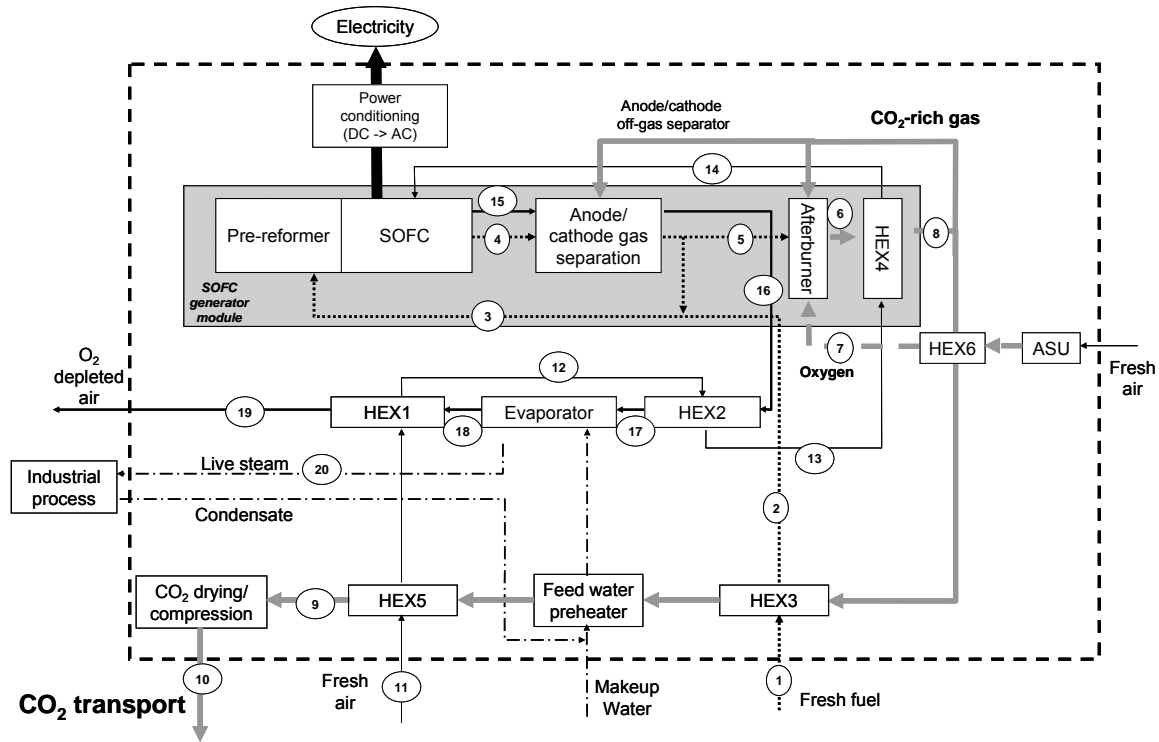
#### *Modifications to the SOFC system for CO<sub>2</sub> capture*

Figure 5-2 depicts off-gas flow manifolding for tubular SOFC systems with and without CO<sub>2</sub> capture. In the Siemens SOFC module, anode and cathode off-gases mix through controlled leakage through baffle boards. Anode off-gas leaks into the cathode off-gas plenum, where the combustion takes place (EG&G Technical Services Inc., 2004). For post-SOFC CO<sub>2</sub> capture, however, anode and cathode off-gases must be well separated. To make the best of the existing concept and yet separating anode and cathode off-gases, a buffer plenum of steam between the anode off-gas stream and the cathode off-gas stream has been proposed as a viable approach to maintaining the separation of those streams (Figure 5-2, right; (Draper and Veyo, 2002)). For CO<sub>2</sub> capture SOFC systems, part of afterburner flue gas, rich in CO<sub>2</sub> and water vapor, can also be used to separate anode and cathode off-gases instead of using part of generated process steam for some configurations.

#### **5.2.1.3 Pure oxygen from a stand-alone ASU**

##### *Air separation unit (ASU)*

The cost-effectiveness of specific air separation techniques depend on the production volume, purity and the required state (liquid or gaseous). For the SOFC systems applied in this study (5 MW<sub>e</sub> scale), the oxygen consumption rate is estimated around 15-30 t d<sup>-1</sup>. For such oxygen production volumes and the timeframe considered in this study, pressure, vacuum and vacuum-pressure swing adsorption (PSA, VSA and VPSA) technologies are likely to be the most economical (Universal Industrial Gases Inc., 2009). Although the purity of oxygen separated by PSA, VSA and VPSA can only reach around 95%, high oxygen purities, e.g. above 99%, are not required for CO<sub>2</sub> capture SOFC systems because the oxygen is used to oxidize only around 15% of the input fuel.



**Figure 5-3:** Simplified diagram of the industrial SOFC-CHP with CO<sub>2</sub> capture (ISOFC-CC).

In the longer term, advanced modular air separation technologies, e.g. solid oxide electrolyzer cell (SOEC) and OCM, may replace adsorption-based air separation technologies. SOEC is a reverse SOFC, generating hydrogen and oxygen via electrolysis of water. Hydrogen can be used as a fuel to generate the electricity required by the electrolysis itself (Iora and Chiesa, 2009). OCM is, like SOFC, made of dense, complex metal oxide ceramics and has both electronic and oxygen ionic conductivity when operated at high temperatures above 700 °C (Anheden et al., 2005a). The main difference between the two oxygen transport membranes is that SOEC is electricity driven, whereas OCM is pressure driven. Both SOEC and OCM produces oxygen of very high purity, above 99% (Iora and Chiesa, 2009; Smith and Klosek, 2001), which is a significant advantage over adsorption-based air separation technologies.

#### *Afterburner design*

The oxyfuel combustion technology has been studied extensively in recent years and is currently in the demonstration phase. The ongoing demonstration projects have applied the flue gas recycling method to control the temperature (Cieutat et al., 2008; Strömberg et al., 2008), because the oxyfuel combustion flame temperature can reach 2500 °C (Praxair, 2009). The recycled flue gas acts as a screen between the afterburner wall and the flame, thus keeping the wall temperature low. Air Liquide's oxyfuel combustion pilot plant was tested for flue gas recycling rates between 48% and 71% (Cieutat et al., 2008).

#### *Overall system configuration*

Figure 5-3 shows a simplified diagram of the industrial SOFC-CHP system with CO<sub>2</sub> capture (ISOFC-CC). The SOFC generator consists of multiple 500kW<sub>e</sub> modules. With regard to fuel reforming, indirect internal reforming technique, which is used by SPC (van Gerwen, 2003) and Wärtsilä (Fontell et al., 2004), were selected for both ISOFC and ISOFC-CC. Anode and cathode off-gases are separated by a small fraction of the recycled CO<sub>2</sub>-rich afterburner flue gas. As a consequence of avoiding the mixing anode and cathode off-gases, some afterburner flue gas leaks into the off-gas chambers. The rest of the flue gas is fed to the pre-reformer.

## 5.2.2 Modeling of the CO<sub>2</sub> capture SOFC-CHP systems

### 5.2.2.1 System modeling methodology

Industrial SOFC systems were modeled using Cycle Tempo software (Delft University of Technology, 2008). This software is built for the thermodynamic analysis and optimization of energy conversion systems. Parameter values for the base case industrial SOFC system model are presented in Table 5-2. It is assumed that desulphurized natural gas of Slochteren quality, i.e. 81.3 mol% CH<sub>4</sub>, 2.9% C<sub>2</sub>H<sub>6</sub>, 0.38% C<sub>3</sub>H<sub>8</sub>, 0.15% C<sub>4</sub>H<sub>10</sub>, 0.05% C<sub>5</sub>H<sub>12</sub>, 14.3% N<sub>2</sub>, 0.01% O<sub>2</sub> and 0.89% CO<sub>2</sub>, is available. The overall cell resistance is calculated for ISOFC with the cell performance of 0.3W cm<sup>-2</sup> at 0.7V. The calculated cell resistance value is applied to all operations of the industrial SOFC systems investigated in this paper. Such an approach can be justified because constant cell operation temperature is assumed for all the industrial SOFC systems assessed in this paper. De Groot concluded from experimental results that it is acceptable to assume the cell resistance as a function of operation temperature only (de Groot, 2004).

### 5.2.2.2 Competing technologies

Taking into account the plant scale an SOFC system can achieve within the timeframe considered in this study, the net electrical capacity of the reference industrial CHP was assumed to be 5 MW. For plant scales between 1 MW<sub>e</sub> to 20 MW<sub>e</sub>, typical conventional industrial CHP plants are gas-powered. We selected a gas turbine CHP plant (GT-CHP) as the reference conventional CHP technology. GT-CHP is commonly used to produce high quality heat; an unfired HRSG can generate steam at conditions approximately ranging between 10 bar and 80 bar (EPA, 2008). The reference GT-CHP was assumed to be equipped with today's state-of-the-art gas turbine. Currently the highest electrical efficiency (without steam injection) for 1-10 MW<sub>e</sub> scale is 38.5% achieved by Solar's 4.6 MW<sub>e</sub> Mercury50 recuperated gas turbine (GTW, 2007). This efficiency was assumed to become a standard for the gas turbines having a similar output for the timeframe considered in this study.



**Table 5-2:** Design parameters for the base case industrial SOFC systems

Parameter	Value	Parameter	Value
<b>SOFC</b>		<b>Oxygen supply</b>	
Fuel utilization rate ( $U_F$ ) <sup>1)</sup>	85%	Oxygen purity <sup>14)</sup>	95 vol%
Oxidant utilization rate ( $U_{Ox}$ ) <sup>2)</sup>	25%	Pressure and temperature <sup>15)</sup>	1.013 bar, 15 °C. Preheated to 500 °C.
Anode inlet temperature <sup>3)</sup>	800 °C	Specific electricity consumption <sup>16)</sup>	300 kWh tonne <sup>-1</sup>
Air inlet temperature to the SOFC module <sup>4)</sup>	600 °C	<b>Heat exchangers</b>	
Cathode inlet temperature to the SOFC stack <sup>5)</sup>	800 °C	Pressure drop	0 bar for internal heat exchange within the SOFC module. 0.05 bar for other heat exchangers.
Anode off-gas recycling rate <sup>6)</sup>	60%	Temperature Restrictions	Heat lower than 100 °C not utilized. <sup>17)</sup> Pinch temperature for HRSG: > 15 °C <sup>18)</sup>
Operation temperature and pressure <sup>7)</sup>	1000 °C, 1.35 bar	Approach temperature for HRSG:	10 °C <sup>19)</sup>
Internal reformer reaction temperature <sup>8)</sup>	750 °C	Minimum $\Delta T$ (gas-gas):	25 K <sup>20)</sup>
		Gas-gas:	30 W m <sup>-2</sup> K <sup>-1</sup>
		HRSG economizer:	42.6 W m <sup>-2</sup> K <sup>-1</sup>
		HRSG evaporator:	43.7 W m <sup>-2</sup> K <sup>-1</sup>
<b>Fuel cell off-gas separation for CO<sub>2</sub> capture</b>		<b>Overall heat transfer coefficients<sup>21)</sup></b>	
Fraction of barrier gas leakage to anode and cathode off-gas chambers <sup>9)</sup>	10% each (total 20% leakage)		
Amount of barrier gas injected into the pre-reformer <sup>10)</sup>	2% of anode off-gas	<b>Other balance-of-plant components</b>	
		Water pumps	$\eta_{Isentropic} = 0.7, \eta_{Mechanic} = 0.85$
		Air and gas blowers	$\eta_{Isentropic} = 0.85, \eta_{Mechanic} = 0.9$
		Deaerator temperature	105 °C
		Boiler blowdown rate <sup>22)</sup>	5%
		Condensate return (%-net steam output) <sup>23)</sup>	90%
		Condensate temperature <sup>24)</sup>	90 °C
		Makeup water temperature	15 °C
		CO <sub>2</sub> compression	5 stage compression to 110 bar at 30 °C. $\eta_{Isentropic} = 0.8, \eta_{Mechanic} = 0.9$
<b>Oxyfuel afterburner</b>			
Excess oxygen <sup>11)</sup>	1 % stoichiometric requirement		
Flue gas recycle ratio <sup>12)</sup>	70 %		
Recycled flue gas temperature <sup>13)</sup>	650 °C		
Afterburner pressure drop	0.05 bar		

<sup>1)</sup> Compared to fresh fuel input to the fuel cell system. The value agrees with the range reported in the literature for both experimental and modeling results between 80% (Riensch et al., 1998a; Riensch et al., 1998b) and 90% (TIAX LLC, 2003).

<sup>2)</sup> Based on Zhang et al. (2006). Literature values range between 17.9% (Campanari, 2001) and 71.6% (Maurstad et al., 2004). This value is chosen in order to control the temperature gradient in the SOFC stack within 150 °C.

- <sup>3)</sup> The values in the literature range between around 650 °C and 850 °C (Maurstad et al., 2004; Riensche et al., 2000; Riensche et al., 1998a; Riensche et al., 1998b; TIAX LLC, 2002).
- <sup>4)</sup> Literature values range between 600 °C (Lundberg et al., 2000) and 630 °C (Campanari, 2001; Riensche et al., 2000; Zhang, 2006).
- <sup>5)</sup> Most studies on tubular SOFC suggest a temperature around 820 °C (Campanari, 2001).
- <sup>6)</sup> The value is obtained from (Campanari, 2001; Riensche et al., 1998a) to supply sufficient amount of steam for methane reforming and to avoid carbon deposition.
- <sup>7)</sup> Literature values range between 800 °C and 1000 °C. SPC's demonstration plants are operated at 1000 °C (Veyo, 1997). The operating pressure is set sufficiently high to avoid extra air blowers/compressors in the SOFC system. SPC's demonstration plant in Italy operated at a higher pressure of 2 bar (Verda and Calì Quaglia, 2008).
- <sup>8)</sup> Winkler (2003).
- <sup>9)</sup> There are no data available from Siemens on the fraction of barrier steam that would leak into anode and cathode off-gas chambers. Because the barrier steam pressure will be set slightly higher than the off-gas chamber pressures (Draper and Veyo, 2002), we assume that the steam leakage will be minimized.
- <sup>10)</sup> There are no data available from SPC on the fraction of barrier steam that would be injected into the pre-reformer. Nevertheless, it was assumed that the amount would be minimal because there should be sufficient steam available in the recycled depleted fuel.
- <sup>11)</sup> In order to ensure full oxidation of the remaining fuel.
- <sup>12)</sup> See subsection "Afterburner design" in Section 5.2.1. A higher end value was used for this study.
- <sup>13)</sup> Maximum acceptable temperature in order to use low-temperature steel alloy heat exchangers downstream (Magistri et al., 2006).
- <sup>14)</sup> Highest achievable purity for PSA/VPSA technique (Goldstein et al., 2003; Universal Industrial Gases Inc., 2009), because the argon content cannot be removed with this technique.
- <sup>15)</sup> Ambient conditions.
- <sup>16)</sup> Wilkinson et al. (2003) reported 234 kWh tonne<sup>-1</sup> for 93% purity O<sub>2</sub> from a 131 t d<sup>-1</sup> VSA plant, and Speece (2003) reports 400 kWh tonne<sup>-1</sup> for 90-95% purity O<sub>2</sub> from a 1 t d<sup>-1</sup> PSA plant.
- <sup>17)</sup> Sjardin et al. (2006).
- <sup>18)</sup> Möllersten et al. (2006).
- <sup>19)</sup> A good approach temperature is between 8.3 °C and 16.7 °C (C-B Energy Recovery, 2009). The value used in the current study is on the lower side of the suggested range.
- <sup>20)</sup> The literature values range between 25-30 °C (Bolland, 1993; Chiesa et al., 2005; Hamelinck and Faaij, 2002).
- <sup>21)</sup> Coefficient for gas-gas heat exchanger is on the higher side of the range suggested in the literature (5-35 W m<sup>-2</sup> K<sup>-1</sup>) (Engineering Page, 2008; Perry and Green, 2007). Coefficients for HRSG obtained from a study on the optimization of HRSG (Casarosa et al., 2004).
- <sup>22)</sup> Blowdown rates for industrial boilers typically range from 4% to 8% of the feedwater flow rate (DOE, 2001).
- <sup>23)</sup> The condensate return rates are case-dependent. It was however assumed that the CHP plants considered in this study has a rather high condensate return rate, which saves makeup water consumption.
- <sup>24)</sup> Agrees with the temperature assumed (93 °C) for process model developed described in Bolland (1993).

To compare GT-CHP, ISOFC and ISOFC-CC on an equal footing, it is important that all systems compared are based on the same assumptions and parameter values. For conventional CHP plants, however, it is unnecessary to model the whole system because there are a number of publicly available softwares that calculates CHP efficiencies by feeding only a few operational parameter values. One of these softwares, Cogen Ready Reckoner software (Sinclair Knight Merz Pty Ltd., 2002), was used to calculate the overall energy performance of the reference GT-CHP. The technical assumptions used for the heat exchanger performance calculations are given in Table 5-2. The gas turbine specifications were based on those of the Mercury50 given in the GTW Handbook 2007-2008 (GTW, 2007).

### 5.2.2.3 Comparison of CHP options

The comparison of different CHP plants was based on identical heat output. The net electrical output for ISOFC was identical to that for GT-CHP unless the calculated cell voltage is below the defined minimum limit. Otherwise, the net electrical output of ISOFC is calculated for the minimum cell voltage and the excess electricity was assumed to be exported to the grid. Lastly, ISOFC-CC was assumed to have identical heat and electrical outputs as ISOFC.

### 5.2.2.4 Economic performance calculation

There are several indicators to estimate the cost performance of CO<sub>2</sub> capture. In case of power plants, commonly used economic indicators are cost of electricity (COE) and CO<sub>2</sub> mitigation cost compared to a reference system. For the case of CHP plants, however, COE largely depends on how costs are allocated to the produced electricity and heat. We therefore use CO<sub>2</sub> mitigation cost as the economic indicator for post-combustion CO<sub>2</sub> capture (Eq.(5.1)):

$$C_{CO_2} = \frac{\alpha * (I_{SOFC-CHP-CC} - I_{ref}) + (\Delta C_{fuel} + \Delta C_{O\&M})}{Em_{av}} \quad (5.1)$$

where  $C_{CO_2}$  is the CO<sub>2</sub> mitigation cost (\$ t<sup>-1</sup>),  $\alpha$  is the annuity factor (yr<sup>-1</sup>),  $I_{SOFC-CC}$  is the total capital requirement (TCR) for the SOFC with CO<sub>2</sub> capture (\$),  $I_{CHP}$  is TCR for the reference plant (ISOFC or GT-CHP) (\$),  $\Delta C_{fuel}$  is the annual cost or benefit for the change in fuel consumption (\$ yr<sup>-1</sup>),  $\Delta C_{O\&M}$  is the annual additional operation and maintenance (O&M) costs for ISOFC-CC compared to the reference plant (\$ yr<sup>-1</sup>) and  $Em_{av}$  is the annual CO<sub>2</sub> emission avoided compared to the reference plant (t yr<sup>-1</sup>). The cost data used in this study are presented in Table 5-3. All cost figures given in this paper are expressed in 2007 US\$. Other parameters and their values used in our calculations are presented in Table 5-4.

**Table 5-3:** Equipment cost data for the main components of industrial SOFC-CHP systems

Component	Cost (\$)	Cost definition	Base scale		Scaling factor
			Value	Unit	
Fuel processing <sup>1)</sup>	2.38 x 10 <sup>5</sup>	Installed cost	14	MW <sub>e</sub> net output	1
AC/DC inverter <sup>2)</sup>	2.85 x 10 <sup>5</sup>	Installed cost	2	MW <sub>e</sub> DC output	0.7
Heat exchanger (carbon steel) <sup>3)</sup>	1.19 x 10 <sup>5</sup>	Installed cost	100	m <sup>2</sup>	0.36
Heat exchanger (stainless steel) <sup>4)</sup>	3.57 x 10 <sup>5</sup>	Installed cost	100	m <sup>2</sup>	0.36
Evaporator <sup>5)</sup>	3.17 x 10 <sup>5</sup>	Installed cost	100	m <sup>2</sup>	0.49
Afterburner <sup>6)</sup>	3.57 x 10 <sup>5</sup>	Installed cost	1	MW heat duty	0.46
SOFC vessel, insulation and air plenum assembly <sup>7)</sup>	6.31 x 10 <sup>5</sup>	Installed cost	1.8	MW <sub>e</sub> output	1
Air blower <sup>8)</sup>	7.09 x 10 <sup>4</sup>	Capital cost	445	kW <sub>e</sub> consumption	0.67
CO <sub>2</sub> compressor <sup>9)</sup>	1.56 x 10 <sup>6</sup>	Installed cost	0.7	MW <sub>e</sub> consumption	0.67

**Table 5-3 continued**

Small scale stand-alone air separation unit <sup>10)</sup>	9.15 x 10 <sup>4</sup>	Capital cost	1	tO <sub>2</sub> d <sup>-1</sup>	0.83
Reference industrial GT-CHP plant <sup>11)</sup>	5.65 x 10 <sup>6</sup>	Installed cost	5	MWe net output	

<sup>1)</sup> From Lundberg et al. (2000), including desulfurizer, pre-reformer (including catalyst), fuel preheater and ejector, and a recirculation loop. Scaling factor is 1 because the fuel processing system is likely to be equipped for each cell substack of around 100 kW<sub>e</sub> capacity for multi-MW systems (Lundberg et al., 2000). The costs can be significantly more expensive for external reformers; Braun (2001) used a cost estimate of 20,000\$ (2000 value) for a 200 kW<sub>e</sub> plant.

<sup>2)</sup> Cost data is from Lundberg et al. (2000). It is assumed that every SOFC generator module (500kW<sub>e</sub> DC) has its own inverter. Scaling factor is taken from Calise et al. (2007).

<sup>3)</sup> For gas streams below 650 °C. Linear regression ( $R^2 = 0.93$ ) derived from the data in (DOE/NETL, 2002) for areas between 9 m<sup>2</sup> and 929 m<sup>2</sup>.

<sup>4)</sup> For gas streams above 650 °C. Stainless steel type 316L was assumed to be main material, which tolerates temperatures up to 930 °C, which is sufficiently high for the studied system. 316L stainless steel was assumed factor three more expensive than carbon steel (Braun, 2002). This value is in agreement with the cost ratios reported in (Perry and Green, 2007). The cost estimates from this assumption agrees well with those presented in Korobytsin et al. (2000).

<sup>5)</sup> Linear regression ( $R^2 = 0.99$ ) derived from the data in (DOE/NETL, 2002) for heat capacities between 9 m<sup>2</sup> and 929 m<sup>2</sup>.

<sup>6)</sup> Linear regression for three data points from (DOE/NETL, 2002) for firing duty between 1MW and 7MW.

<sup>7)</sup> George and Bessette (1997).

<sup>8)</sup> Calise et al. (2007).

<sup>9)</sup> Cost figure from Lokurlu et al. (2004). The scaling factor is obtained based on the cost figures for 0.7 MW<sub>e</sub> compressor (Lokurlu et al., 2004) and for 13 MW<sub>e</sub> compressor (Kreutz et al., 2005).

<sup>10)</sup> Cost figure is from Bowers (2003) for a plant generating 90-95% purity O<sub>2</sub> with power consumption of 400 kWh tonne<sup>-1</sup>. Together with the cost figure presented in Wilkinson et al. (2003), (5.29 M\$ for 131 tO<sub>2</sub> d<sup>-1</sup>, 234 kWh tonne<sup>-1</sup> at 93% purity), the presented scaling factor was derived. Another source suggested a much lower installed cost of 200,000 \$ (1996 value) for 26 tO<sub>2</sub> d<sup>-1</sup> plant (Worrell et al., 2008). For this study, a more conservative assumption is made.

<sup>11)</sup> From EPA (2008) for typical GT-CHP plants. This value agrees with the literature estimates, which range between 5.1 M\$ (Khrushch et al., 1999) and 5.7 M\$ (Axelsson et al., 2003). Generally the capital costs of state-of-the-art technologies are more expensive than the conventional ones. By applying the cost data of a typical (i.e. conventional) GT-CHP to the state-of-the-art GT-CHP, future potential cost reductions are implicitly taken into account.

**Table 5-4: Other parameters and their values used in this study**

<i>Parameter</i>	<i>Value</i>	<i>Parameter</i>	<i>Value</i>
Total plant cost (%-bare erected cost) <sup>1), 2)</sup>	130%	Natural gas price (incl. tax, \$ GJ <sup>-1</sup> ) <sup>6)</sup>	10
Total plant cost (%-bare erected cost): GT-CHP <sup>2), 3)</sup>	115%	Grid electricity CO <sub>2</sub> emission factor (t MWh <sup>-1</sup> ) <sup>1) 7)</sup>	0.5
Total capital requirement (%-total plant cost) <sup>1)</sup>	110%	Excess electricity price (\$ MWh <sup>-1</sup> ) <sup>8)</sup>	80
Plant economic lifetime: all CHP plants (years)	20	O&M costs: Conventional CHP (%-TCR) <sup>9)</sup>	5
Fuel cell stack lifetime (years) <sup>4)</sup>	5	O&M costs: stand-alone ASU (%-TCR) <sup>10)</sup>	2
Capacity factor (h yr <sup>-1</sup> )	8500	O&M costs: SOFC-CHPs (excl. SOFC replacement, %-total direct cost) <sup>11)</sup>	5
Real interest rate <sup>5)</sup>	12%	SOFC replacement cost (%-equipment cost)	5

- <sup>1)</sup> The presented value is obtained from van Harmelen et al. (2008) for power plants with CO<sub>2</sub> capture.
- <sup>2)</sup> Bare erected cost includes equipment costs and installation costs (material costs, direct and indirect labor costs) and sales tax (NETL, 2008). Total plant cost includes engineering fees and contingencies.
- <sup>3)</sup> 10% engineering fees and 5% contingencies is assumed.
- <sup>4)</sup> See Section 2.1. For this study a conservative value was used.
- <sup>5)</sup> The interest rates for the industry found in the literature were 10% for cement production (IEA GHG, 2008), 10% for methanol and hydrogen production (Hamelinck and Faaij, 2002), 10%-15% for paper mills (Möllersten et al., 2006; Möllersten et al., 2003), and 12% for iron and steel production (Gielen, 2003).
- <sup>6)</sup> The natural gas price forecasts for 2020-2025 widely differ from source to source. The IEA World Energy Outlook 2008 (IEA, 2008d) forecasts 12\$ GJ<sup>-1</sup> for Europe and 14.5 \$ GJ<sup>-1</sup> for the U.S. for 2025 (tax excluded), while U.S. Department of Energy's Annual Energy Outlook 2009 forecasts below 8 \$ GJ<sup>-1</sup> for the U.S. (EIA, 2009).
- <sup>7)</sup> The CO<sub>2</sub> emission intensity of electricity is heavily dependent on the energy and economic situation of a country. The value assumed here is very close to that for U.K. (IEA GHG, 2008).
- <sup>8)</sup> IEA World Energy Outlook forecasts that the electricity generation costs for the US would be at least 80 \$ MWh<sup>-1</sup>.
- <sup>9)</sup> Total plant cost comprises of bare erected cost, engineering fees and contingencies. Goldstein et al. (2003) estimates that the total O&M costs range between 5% and 6% of TPC.
- <sup>10)</sup> Because of extremely simple operation, PSA is normally operated unmanned (Bowers, 2003), thus a low O&M cost is expected.
- <sup>11)</sup> (Korobytsin et al., 2000).

## **5.3 Results**

### **5.3.1 Technical performance**

Table 5-5 and Table 5-6 present thermodynamic properties for the relevant points of the base case industrial SOFC-CHP systems without CO<sub>2</sub> capture (ISOFC) and with CO<sub>2</sub> capture (ISOFC-CC), presented in Figures 1 and 3, respectively. The results show that all the SOFC off-gases, i.e. afterburner flue gas in ISOFC (point 10 in Table 5-5), CO<sub>2</sub>-rich gas and cathode off-gas in ISOFC-CC (points 9 and 19 in Table 5-6), are cooled down to the assumed minimum temperature (100 °C) via heat exchangers, meaning that maximum heat recovery is achieved. The fuel cell module operating pressure has reduced from 1.35 bar for ISOFC to 1.3 bar for ISOFC-CC, because the pressure losses that occur in the heat exchangers are diverged into two fuel cell off-gas streams. The low CO<sub>2</sub> purity (85%, point 10 in Table 5-6) is due to the high nitrogen content in the natural gas. Table 5-7 shows the calculated heat exchanger areas for the base case industrial SOFC systems with and without CO<sub>2</sub> capture.

**Table 5-5:** Thermodynamic properties for the relevant points of the base case industrial SOFC-CHP plant without CO<sub>2</sub> capture (ISOFC) of Figure 5-1.

Point no.	Temperature °C	Pressure [Bar]	Mass flow [kg/s]	Molar composition											
				CH <sub>4</sub>	C <sub>2</sub> H <sub>6</sub>	C <sub>3</sub> H <sub>8</sub>	C <sub>4</sub> H <sub>10</sub>	C <sub>5</sub> H <sub>12</sub>	C <sub>6</sub> H <sub>14</sub>	N <sub>2</sub>	O <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	Ar	H <sub>2</sub>
1	15	1.01	0.311	0.8129	0.0287	0.0038	0.0015	0.0004	0.0005	0.1432	0.0001	0.0089	0	0	0
2	296	1.35	0.311	0.8129	0.0287	0.0038	0.0015	0.0004	0.0005	0.1432	0.0001	0.0089	0	0	0
3	800	1.35	1.98	0.1597	0.0056	0.0007	0.0003	0.0001	0.0001	0.0694	0	0.1986	0.4199	0	0.0822
4	968	1.35	2.78	0.0026	0	0	0	0	0	0.0517	0	0.2442	0.521	0	0.102
5	968	1.35	1.11	0.0026	0	0	0	0	0	0.0517	0	0.2442	0.521	0	0.102
6	1097	1.35	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
7	918	1.3	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
8	908	1.25	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
9	492	1.2	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
10	165	1.1	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
11	100	1.05	14.2	0	0	0	0	0	0	0.7513	0.1415	0.0304	0.068	0.0089	0
12	15	1.01	13.9	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0
13	126	1.45	13.9	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0
14	600	1.4	13.9	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0
15	800	1.35	13.9	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0
16	968	1.35	13.1	0	0	0	0	0	0	0.8152	0.1641	0.0003	0.0107	0.0097	0
17	180	10	1.98	0	0	0	0	0	0	0	0	0	1	0	0

**Table 5-6:** Thermodynamic properties for the relevant points of the base case industrial SOFC-CHP plant with CO<sub>2</sub> capture (ISOFC-CC) of Figure 5-3.

Point no.	Temperature °C	Pressure [bar]	Mass flow [kg/s]	Molar composition												
				CH <sub>4</sub>	C <sub>2</sub> H <sub>6</sub>	C <sub>3</sub> H <sub>8</sub>	C <sub>4</sub> H <sub>10</sub>	C <sub>5</sub> H <sub>12</sub>	C <sub>6</sub> H <sub>14</sub>	N <sub>2</sub>	O <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	Ar	H <sub>2</sub>	CO
1	15	1.013	0.327	0.8129	0.0287	0.0038	0.0015	0.0004	0.0005	0.1432	0.0001	0.0089	0	0	0	0
2	364	1.3	0.327	0.8129	0.0287	0.0038	0.0015	0.0004	0.0005	0.1432	0.0001	0.0089	0	0	0	0
3	800	1.3	2.20	0.151	0.0053	0.0007	0.0003	0.0001	0.0001	0.0685	0	0.2017	0.4261	0.0002	0.0826	0.0635
4	947	1.3	3.04	0.002	0	0	0	0	0	0.0516	0	0.2431	0.5195	0.0002	0.1038	0.0798
5	940	1.3	1.25	0.002	0	0	0	0	0	0.0517	0	0.2451	0.522	0.0003	0.1011	0.0778
6	1045	1.3	4.73	0	0	0	0	0	0	0.0519	0.0009	0.3217	0.621	0.0044	0	0
7	500	1.3	0.167	0	0	0	0	0	0	0.008	0.95	0	0	0.042	0	0
8	650	1.25	4.73	0	0	0	0	0	0	0.0519	0.0009	0.3217	0.621	0.0044	0	0
9	100	1.05	1.32	0	0	0	0	0	0	0.0519	0.0009	0.3217	0.621	0.0044	0	0
10	30	110	0.776	0	0	0	0	0	0	0.137	0.0024	0.8486	0.0004	0.0116	0	0
11	15	1.013	14.5	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0	0
12	168	1.4	14.5	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0	0
13	607	1.35	14.5	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0	0
14	800	1.3	14.5	0	0	0	0	0	0	0.7729	0.2075	0.0003	0.0101	0.0092	0	0
15	947	1.3	13.7	0	0	0	0	0	0	0.8152	0.1641	0.0003	0.0107	0.0097	0	0
16	947	1.3	13.7	0	0	0	0	0	0	0.8149	0.1641	0.0004	0.0109	0.0097	0	0
17	512	1.25	13.7	0	0	0	0	0	0	0.8149	0.1641	0.0004	0.0109	0.0097	0	0
18	205	1.2	13.7	0	0	0	0	0	0	0.8149	0.1641	0.0004	0.0109	0.0097	0	0
19	100	1.15	13.7	0	0	0	0	0	0	0.8149	0.1641	0.0004	0.0109	0.0097	0	0
20	180	10	1.98	0	0	0	0	0	0	0	0	0	1	0	0	0

**Table 5-7:** Calculated heat exchanger (HEX) areas for the base case industrial SOFC plants without CO<sub>2</sub> capture (ISOFC) and with CO<sub>2</sub> capture (ISOFC-CC) (unit: m<sup>2</sup>)

<i>Heat exchanger name</i>	<i>ISOFC</i>	<i>ISOFC-CC</i>
HEX1	802	1423
HEX2	697	667
HEX3	8	24
HEX4	345	922
HEX5	---	69
HEX6	---	8
Feedwater preheater	309	83
HRSG: evaporator	905	866

Table 5-8 also shows that  $U_F$  has a significant impact on the cell power density because of the change in current density. The increase of  $U_F$  from 0.85 to 0.9 for ISOFC increases the cell power density by 5%, while the total system efficiency shows a slight decrease (0.7%), which is mainly due to larger air blower power consumption. The calculated low cell voltage (0.592 V), however, may cause technical problems because lower voltage causes higher principal stresses due to higher local current density and larger temperature gradient (Chiang et al., 2008). Literatures suggest a lower voltage limit around 0.6V (Haltiner et al., 2004; Noetzel et al., 2006).

### 5.3.2 Economic performance

#### 5.3.2.1 Investment costs and O&M costs

Table 5-9 presents the calculated costs and Figure 5-4 presents the breakdown of specific capital cost for the base case industrial SOFC-CHP plants. CO<sub>2</sub> capture from industrial SOFC-CHP systems is found to increase the total capital requirement by around 50%, which is principally due to the oxygen production and the CO<sub>2</sub> compression equipment (Figure 5-5). The overall investment costs related to heat exchangers are higher for ISOFC-CC than for ISOFC. To assess the validity of the assumptions for oxygen production, the oxygen separation cost is calculated with electricity consumption given in Table 5-2 and an assumed grid electricity price of 50 \$ MWh<sup>-1</sup>, lifetime of 20 years and a real interest rate of 12%. The calculated cost was 45 \$ t<sup>-1</sup> O<sub>2</sub> (annualized capital: 26 \$, O&M cost: 4 \$, electricity cost: 15 \$). This value agrees with the value range indicated in the literature (Energetics, 2002). However, note that the oxygen separation cost depends largely on the assumed electricity price.



**Table 5-8:** Energy performance carbon dioxide emissions results for 5 MW<sub>e</sub> industrial SOFC-CHP plants with and without CO<sub>2</sub> capture, alongside with the results for the reference GT-CHP.

		Base case			Alternative $U_F$				Alternative $U_{Ox}$			
		GT-CHP	ISOFC	ISOFC-CC	$U_F = 0.8$		$U_F = 0.9$		$U_{Ox} = 0.2$		$U_{Ox} = 0.3$	
					ISOFC	ISOFC-CC <sup>2)</sup>	ISOFC	ISOFC-CC <sup>3)</sup>	ISOFC	ISOFC-CC <sup>2)</sup>	ISOFC	ISOFC-CC
<b>Energy performance</b>												
Fuel input	MW	13.7	11.8	12.4	11.7	12.4	11.9	12.5	12.2	12.9	11.6	12.2
Gross electrical output	MW	5.4	5.7	6.3	5.7	6.3	5.7	6.3	5.9	6.5	5.6	6.2
Auxiliary power consumption	MW	0.4	0.7	0.8	0.7	0.7	0.7	0.8	0.9	1.0	0.6	0.7
ASU power consumption	MW	0.0	0.0	0.2	0.0	0.2	0.0	0.1	0.0	0.2	0.0	0.2
CO <sub>2</sub> compression power	MW	0.0	0.0	0.3	0.0	0.4	0.0	0.3	0.0	0.4	0.0	0.3
Net electrical output	MW	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0
Process steam output	MW	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7
Net electrical efficiency		36.5%	42.3%	40.2%	42.7%	40.3%	42.0%	40.1%	41.0%	38.9%	43.2%	41.1%
Steam efficiency		34.6%	40.1%	38.1%	40.4%	38.2%	39.8%	38.0%	38.9%	36.9%	41.0%	38.9%
<b>Fuel cell performance<sup>1)</sup></b>												
Cell voltage	V	---	0.622	0.658	0.660	0.705	0.592	0.617	0.623	0.659	0.621	0.658
Current density	mA cm <sup>-2</sup>	---	656	541	583	447	711	615	655	541	656	540
Power density	kW m <sup>-2</sup>	---	4.08	3.56	3.85	3.15	4.21	3.79	4.08	3.56	4.07	3.55
<b>CO<sub>2</sub> emissions performance</b>												
Total onsite CO <sub>2</sub> generation rate	kg s <sup>-1</sup>	0.768	0.662	0.697	0.657	0.696	0.666	0.698	0.683	0.721	0.648	0.682
CO <sub>2</sub> capture rate	kg s <sup>-1</sup>	0	0	0.694	0	0.694	0	0.696	0	0.718	0	0.680
O <sub>2</sub> production rate	t d <sup>-1</sup>	0	0	14.4	0	19.2	0	9.6	0	14.9	0	14.1
Annual onsite CO <sub>2</sub> emissions	t yr <sup>-1</sup>	22.12	19.06	0.08	18.93	0.08	19.21	0.08	19.68	0.08	18.68	0.08
CO <sub>2</sub> emission reduction compared to GT-CHP		0	13.8%	99.6%	14.4%	99.7%	13.2%	99.7%	11.0%	99.7%	15.6%	99.7%
CO <sub>2</sub> purity		---	---	84.9%	---	84.4%	---	85.3%	---	84.9%	---	84.9%

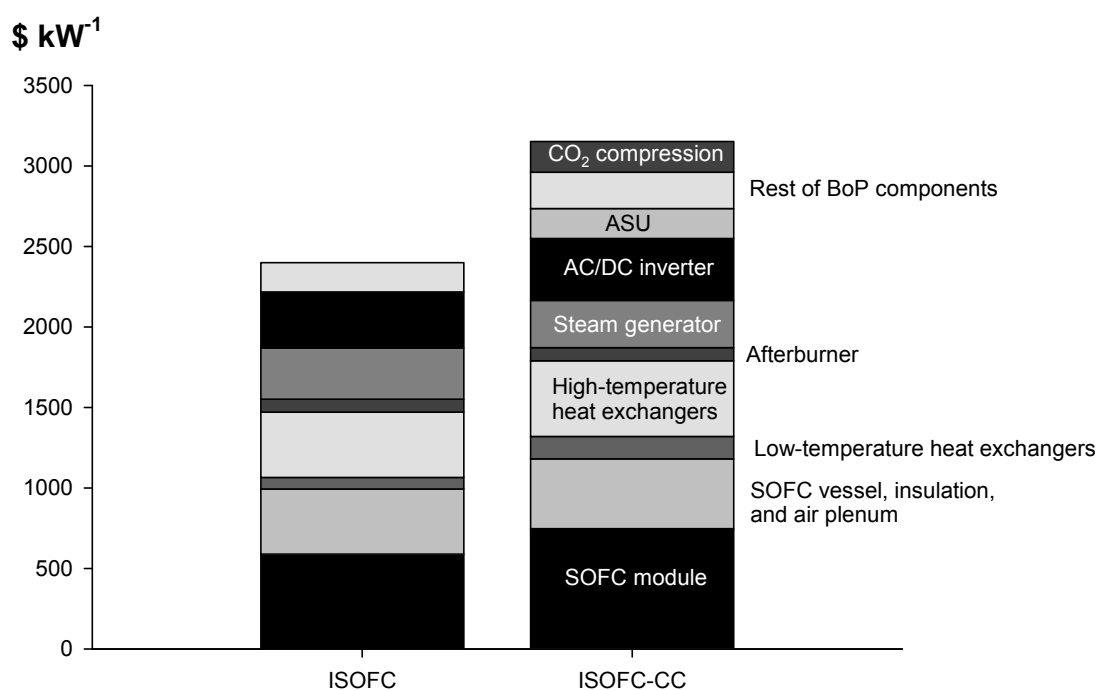
<sup>1)</sup> Based on the cell performance of 0.3 W cm<sup>-2</sup> at 0.7 V for the base case ISOFC, the overall cell resistance of 3.5556\*10<sup>-5</sup> Ω m<sup>2</sup> was derived. The derived overall cell resistance was used for the calculations of all other system performances.

<sup>2)</sup> Recycled afterburner flue gas temperature was changed from 650 °C to 750 °C.

<sup>3)</sup> Cathode inlet temperature was changed from 800 °C to 750 °C.

**Table 5-9:** Calculated costs for the base case industrial SOFC-CHP systems with and without CO<sub>2</sub> capture and for the GT-CHP plant (in 2007 US\$). Cell stack production cost (CSPC) is assumed to be 500 \$ kW<sup>-1</sup>.

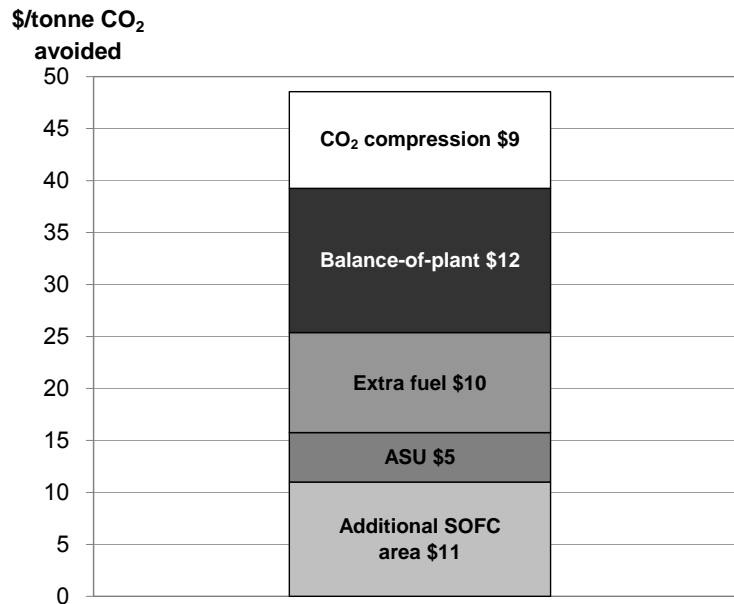
	<i>GT-CHP</i>	<i>ISOFC</i>	<i>ISOFC-CC</i>
<b>Plant component installed costs (10<sup>6</sup> \$)</b>			
Total direct cost	5.7	8.4	11.0
Total process investment	6.5	10.9	14.3
Total capital requirement	7.1	12.0	15.8
Specific capital cost	1430	2400	3150
<b>O&amp;M costs (10<sup>6</sup> \$ yr<sup>-1</sup>)</b>			
Fuel	3.9	3.4	3.6
SOFC stack replacement	0	0.3	0.39
ASU: O&M	0	0	0.0037
CO <sub>2</sub> compression: O&M	0	0	0.048
Other O&M	0.36	0.45	0.55
O&M total	4.3	4.2	4.6
<b>Total annual costs (10<sup>6</sup> \$ yr<sup>-1</sup>)</b>	<b>5.3</b>	<b>5.8</b>	<b>6.7</b>



**Figure 5-4:** Breakdown of specific capital costs for the base case industrial SOFC-CHP systems without CO<sub>2</sub> capture (ISOFC) and with CO<sub>2</sub> capture (ISOFC-CC). Cell stack production cost (CSPC) is assumed to be 500 \$ kW<sup>-1</sup>.

### 5.3.2.2 CO<sub>2</sub> mitigation cost

Figure 5-5 shows the breakdown of CO<sub>2</sub> mitigation cost for ISOFC-CC compared to ISOFC for the base case results. The CO<sub>2</sub> mitigation cost is calculated based on the nominal assumption values given in Table 5-4 and a CSPC of 500 \$ kW<sup>-1</sup>. Please note that the CO<sub>2</sub> mitigation cost largely depends on the CSPC because cell area increased for ISOFC-CC to meet the extra electricity demand for CO<sub>2</sub> capture and compression.

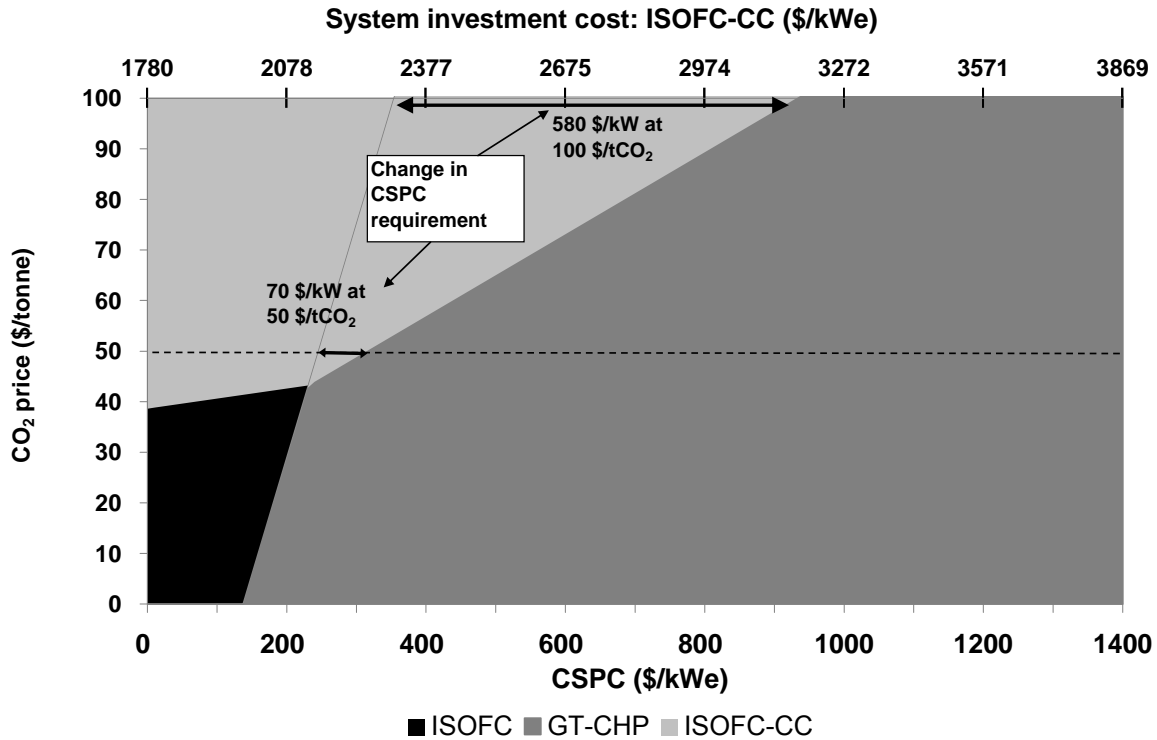


**Figure 5-5:** Breakdown of CO<sub>2</sub> mitigation cost for the base case industrial SOFC-CHP system with CO<sub>2</sub> capture (ISOFC-CC). All costs, except for extra fuel, include annualized capital costs and O&M costs. The costs are based on a cell stack production cost of 500 \$ kW<sup>-1</sup>.

### 5.3.2.3 Competition of three CHP options (GT-CHP, ISOFC and ISOFC-CC)

Figure 5-6 shows the competition of three CHP options (GT-CHP, ISOFC and ISOFC-CC) under varied CO<sub>2</sub> price and CSPC. The dotted line represents the break-even CSPC for ISOFC in comparison with GT-CHP for varied CO<sub>2</sub> prices, and the dashed line represents the CO<sub>2</sub> price of 50 \$ t<sup>-1</sup>. With a zero CO<sub>2</sub> price, CSPC has to be below 140 \$ kW<sup>-1</sup> for ISOFC to compete with GT-CHP. Considering the cost reduction potential by mass production and scale-up of cell stacks reported in Thijssen (2007), the calculated CSPC for tubular cells is unlikely to be achieved.

For CO<sub>2</sub> prices above 43 \$ t<sup>-1</sup>, ISOFC-CC is more economical than ISOFC and competes with GT-CHP for the most economical. This means that when the CO<sub>2</sub> price is above 43 \$ t<sup>-1</sup>, CO<sub>2</sub> capture strengthens the economic competitiveness of the industrial SOFC-CHP plants. ISOFC-CC economically outperformed GT-CHP at CSPCs beyond the dotted line and this effect becomes larger for higher CO<sub>2</sub> prices. At a CO<sub>2</sub> price of 50 \$ t<sup>-1</sup>, CSPC requirement for industrial SOFC to outperform GT-CHP increases by 70 \$ kW<sup>-1</sup>, from 245 \$ kW<sup>-1</sup> to 315 \$ kW<sup>-1</sup>, when equipped with CO<sub>2</sub> capture. This effect became significantly larger when the CO<sub>2</sub> price is higher at 100 \$ t<sup>-1</sup> (580 \$ kW<sup>-1</sup> increase, from 350 \$ kW<sup>-1</sup> to 930 \$ kW<sup>-1</sup>).



**Figure 5-6:** Competitiveness of the three CHP options investigated under varied  $CO_2$  price and cell stack production cost (CSPC).

### 5.3.2.4 Sensitivity analysis

The obtained results are based on many technical and economic assumptions. Therefore, the influence of a number of parameters on the competition of three CHP options was assessed. Figure 5-7 shows the influence of two technical parameters for the SOFC operation, i.e., fuel utilization rate ( $U_F$ ) and oxidant utilization rate ( $U_{Ox}$ ), on the economic competition of the three CHP options. Higher  $U_F$  values is found to be favorable for ISOFC-CC (Figure 5-7 above left), which is principally because  $CO_2$  capture cost becomes lower due to lower oxygen requirement. Regarding  $U_{Ox}$ , ISOFC is hardly competitive at  $U_{Ox} = 0.2$ , mainly due to larger electrical efficiency loss for air blowers. ISOFC-CC is found to be more competitive than ISOFC above the  $CO_2$  price of 40 \$  $t^{-1}$ .

Figure 5-8 shows the influence of ASU power requirement and cell stack lifetime. The cell stack lifetime has a very large impact on the change in CSPC requirement caused by  $CO_2$  capture. Literature suggests that a stack lifetime of 80,000 hours may already be achievable with existing tubular cell technology (Krewitt and Schmid, 2005). In such cases (9 years lifetime: Figure 5-8 bottom right), the CSPC requirement can be as high as 1300 \$  $kW^{-1}$  at 100 \$  $t^{-1}$   $CO_2$  price for ISOFC-CC to compete with GT-CHP. The ASU power requirement, on the contrary, has a limited effect on the competition of the three CHP technologies because total power consumption of ASU is relatively small when the plant scale is considered.

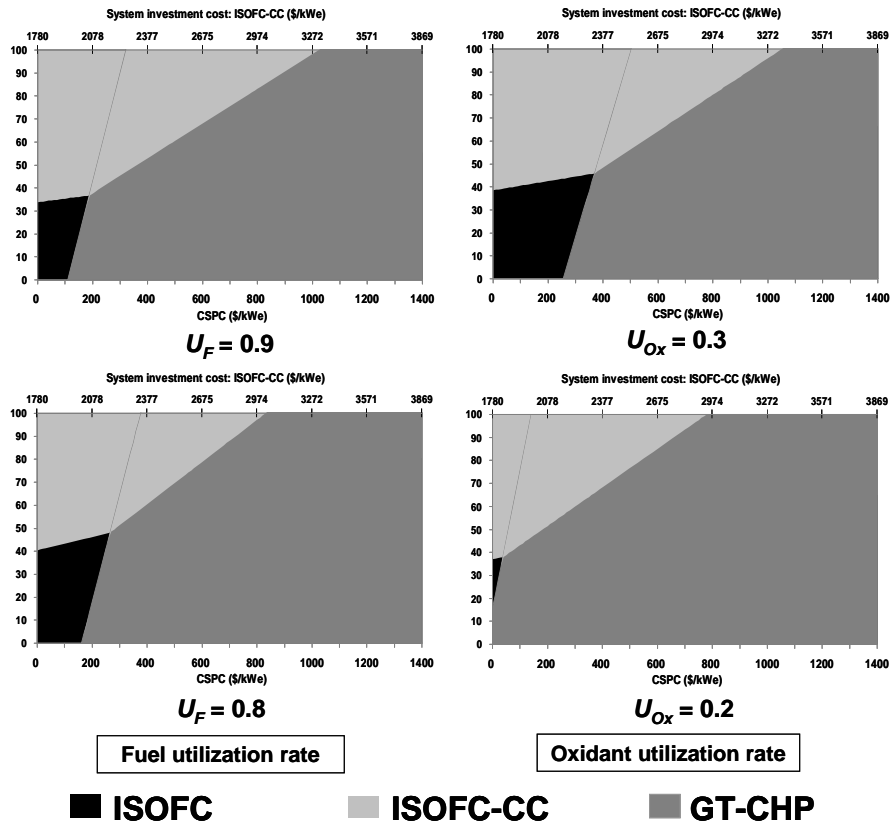


Figure 5-7: Influence of fuel utilization rate ( $U_F$ ) and oxidant utilization rate ( $U_{Ox}$ ) on the competition of the CHP options investigated.

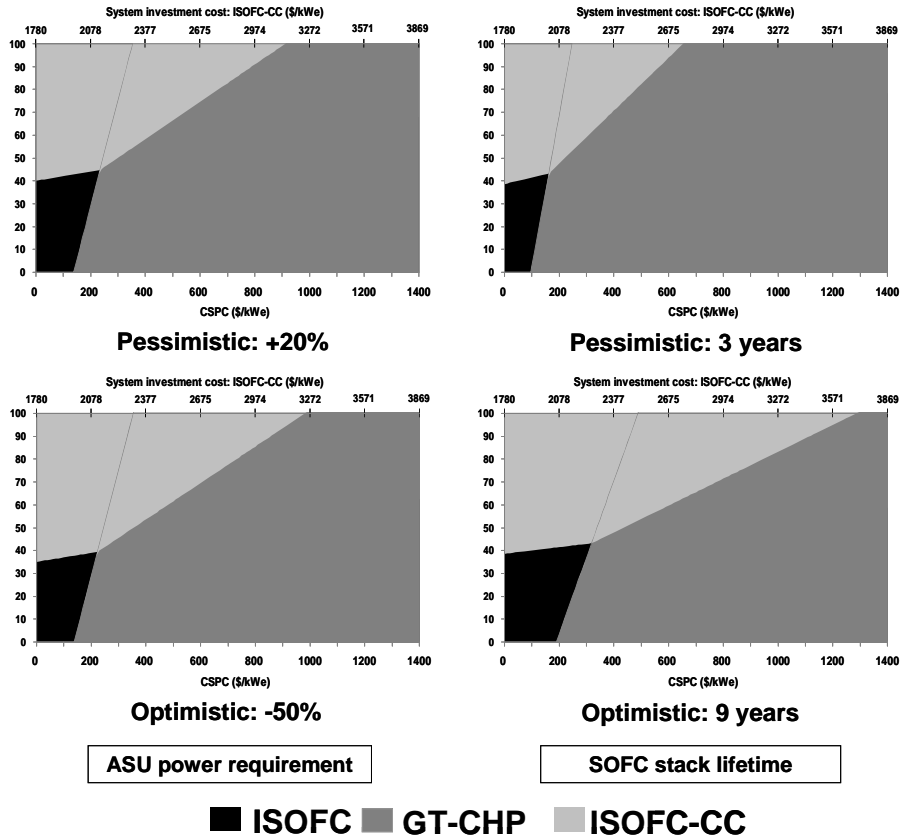
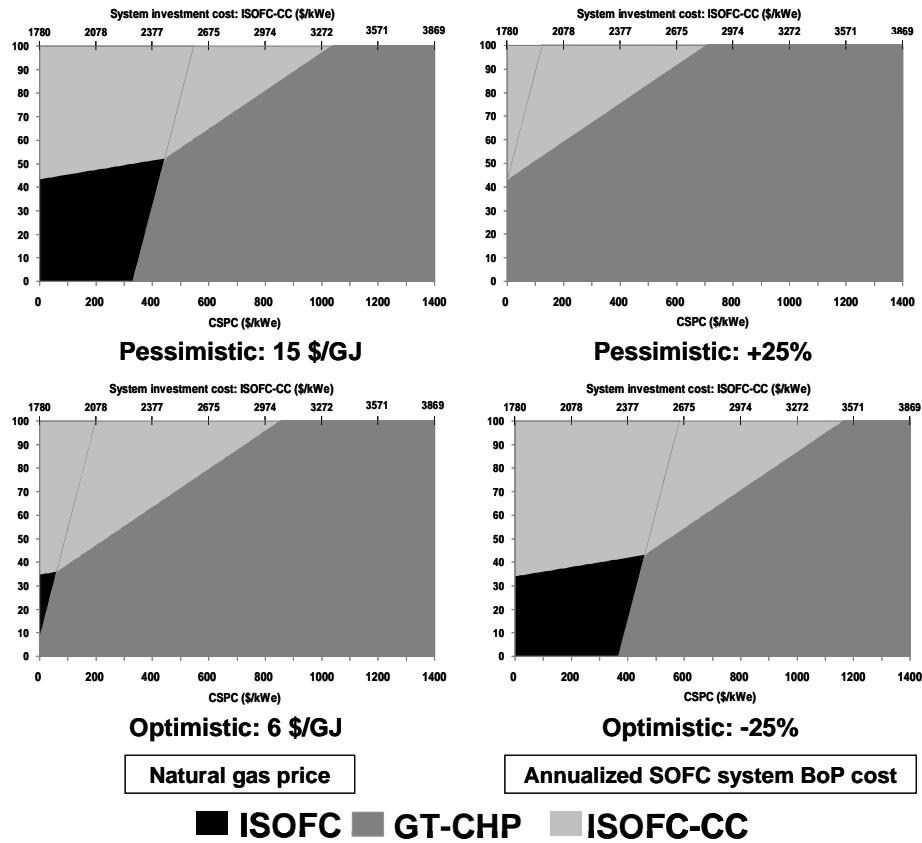


Figure 5-8: Influence of air separation unit power requirement and SOFC stack lifetime on the competition of the CHP options investigated.



**Figure 5-9:** Influence of natural gas price and annualized SOFC system balance-of-plant capital cost on the competition of three CHP technologies.

Figure 5-9 shows the influence of key economic parameters on the economic competition of the three CHP options. The figure shows that the impact of natural gas price on the competitiveness is stronger on ISOFC than on ISOFC-CC. When natural gas price is low, for example, the SOFC systems do not benefit economically from its higher energy efficiency compared to the GT-CHP. However, the benefit of CO<sub>2</sub> capture becomes larger because the economic losses for CO<sub>2</sub> capture due to extra fuel consumption decrease. CO<sub>2</sub> capture consequently counters the effect of natural gas price fluctuations.

The annualized balance-of-plant (BoP) costs can be influenced by factors such as the annuity factor, equipment material price increase and contingencies. Unsurprisingly, they affect the economic competitiveness of ISOFC and ISOFC-CC significantly.

## 5.4 Discussion

In this study we analyzed the effect of CO<sub>2</sub> capture on the CSPC requirement for industrial SOFC-CHP systems to compete in the early stage of SOFC's commercialization for industrial applications (up to 2025). The current study is one of the first to investigate whether CO<sub>2</sub> capture enables earlier market introduction and penetration of SOFC technologies into the energy market. In this section, several issues regarding the modeling assumptions and the results are discussed.

#### **5.4.1 Separation of anode and cathode off-gases**

It was assumed in our study that part of recycled afterburner flue gas is used as a barrier to separate anode and cathode off-gases. There are, however, a number of uncertainties for the separation technique. First, there are neither modeling nor experimental studies on off-gas separation available in the open literature. A/S Norske Shell, together with Siemens, planned to test a 250 kW CO<sub>2</sub>-separating SOFC plant (Haines et al., 2002), but the plan seems to have halted in 2003 due to economic reasons (CO<sub>2</sub> Norway, 2009). Although it is expected that the separation gas volume and its leakage into off-gas chambers are minimized, it remains an uncertainty.

#### **5.4.2 Future CO<sub>2</sub> price and SOFC development strategy**

The results obtained in this study show that with a CO<sub>2</sub> price above 40-50 \$ t<sup>-1</sup>, industrial SOFCs may economically outperform GT-CHP at a higher CSPC when equipped with CO<sub>2</sub> capture. Such a CO<sub>2</sub> price is very likely to be reached in the developed world within the timeframe considered in this study. IEA's World Energy Outlook 2008 predicts that with policy measures to limit the atmospheric greenhouse gas concentration below 550 ppmv CO<sub>2</sub>-equivalent, the CO<sub>2</sub> price of 40 \$ t<sup>-1</sup> for 2020 and 90 \$ t<sup>-1</sup> for 2030 are projected for OECD and non-OECD EU countries (IEA, 2008d). With a more stringent target of 450 ppmv CO<sub>2</sub>-equivalent, the CO<sub>2</sub> price may reach as high as 180 \$ t<sup>-1</sup> in 2030. Our results indicate that it may be worthwhile for SOFC manufacturers not only to focus on the SOFC itself but also on the development of SOFC systems with CO<sub>2</sub> capture capabilities. Nevertheless, the results also indicate that SOFC technology requires very significant cost reductions to compete in the energy market, with or without CO<sub>2</sub> capture.

#### **5.4.3 Costs of CO<sub>2</sub> purification and transport**

The CO<sub>2</sub>-rich stream from the industrial SOFC-CHP systems investigated in this study contained more than 15 vol% impurities, largely due to high nitrogen content in the fuel. For pipeline transport, literature suggests a CO<sub>2</sub> purity above 95 vol% (de Visser et al., 2008). A purification of CO<sub>2</sub>-rich stream may therefore be required. Purification of CO<sub>2</sub>-rich stream was not considered in this paper because it was considered uneconomical for such small volumes. For such small emission sources, CO<sub>2</sub> purification as well as transport should be done collectively with CO<sub>2</sub> streams from other capture sites to minimize costs. The CO<sub>2</sub> recovery rate after purification is reported to range around 87-96%, and the power consumption for CO<sub>2</sub> purification, compression and liquefaction at 110 atm around 450-600 kWh kg<sup>-1</sup>, depending on the amount of impurities (12-25 vol% dry) (Pipitone and Bolland, 2009).

#### **5.4.4 Optimization of heat recovery for the best economic performance**

In this study, ISOFC-CC was designed to maximize heat and electrical efficiencies. As a result, a significant increase in heat exchanger costs compared to ISOFC was observed and

this largely affected the CO<sub>2</sub> mitigation cost. A large cost reduction for heat exchanger in exchange for a limited reduction in heat efficiency and the consequent reduction of CO<sub>2</sub> mitigation cost may be possible.

## 5.5 Conclusion

The aim of this research was to investigate whether potentially low-cost CO<sub>2</sub> capture from SOFC systems would facilitate the penetration of SOFC in the energy market in a highly carbon-constrained society in the mid-term future (up to year 2025). An application of SOFC systems in industrial heat and power cogeneration was considered. Firstly, the assessment of the first generation 5 MW<sub>e</sub> scale SOFC systems was assessed. Secondly, the possible CO<sub>2</sub> capture options for SOFC systems using commercial or near-commercial technologies were investigated. Among many options proposed in the literature, the selection of the CO<sub>2</sub> capture option(s) was based on the technological maturity and the potential economic feasibility at small scales. As a result, the oxyfuel afterburner option with a stand-alone small-scale ASU (PSA/VSA technique) was selected. For the reference CHP technology, a gas turbine (GT-) CHP equipped with the state-of-the-art gas turbine was selected.

The technical performance results showed that the industrial SOFC plants are significantly more efficient (16% without CO<sub>2</sub> capture and 10% with CO<sub>2</sub> capture) than the GT-CHP equipped with the state-of-the-art gas turbine. Total energy efficiency loss of 4.3%-points was observed when the SOFC was equipped with CO<sub>2</sub> capture and compression. To meet the extra electrical demand for CO<sub>2</sub> capture and compression, the cell voltage is increased significantly and as a result, the current density decreases and consequently the cell area requirement increases by 30%.

The economic results showed that for CO<sub>2</sub> prices above 43 \$ t<sup>-1</sup>, industrial SOFCs were found to be more economical when equipped with CO<sub>2</sub> capture. At zero CO<sub>2</sub> price, the cell stack production cost requirement for SOFC-CHP without CO<sub>2</sub> capture to outperform GT-CHP was as low as 140 \$ kW<sup>-1</sup>. At a CO<sub>2</sub> price of 50 \$ t<sup>-1</sup>, the CSPC requirement increased to 245 \$ kW<sup>-1</sup>. When SOFC-CHP was equipped with CO<sub>2</sub> capture, CSPC requirement increased further to 315 \$ kW<sup>-1</sup>. The effect of CO<sub>2</sub> capture on CSPC became significantly larger at a CO<sub>2</sub> price of 100 \$ t<sup>-1</sup>. CSPC requirement was as high as 930 \$ kW<sup>-1</sup> with CO<sub>2</sub> capture, while it was 350 \$ kW<sup>-1</sup> without CO<sub>2</sub> capture. This study suggests a possibility for significant CO<sub>2</sub> emissions reduction from small-scale industrial power systems in a cost-effective manner in the mid-term future if the CO<sub>2</sub> transport and storage infrastructure is available. Moreover, the results of the current study suggest that to enhance the penetration of SOFC systems in the energy market, it is worthwhile for SOFC manufacturers not only to focus on the SOFC systems itself but also on the development of SOFC systems with integrated CO<sub>2</sub> capture.

This study, however, has its limitations. Issues such as additional costs for CO<sub>2</sub> purification, actual performance of SOFC off-gas separation, optimization of heat



recovery efficiency and cost, need to be further researched to obtain a better understanding on the technical and economic performance of the SOFC system investigated in this paper.

It is worthwhile to investigate the long-term prospects for SOFC systems with CO<sub>2</sub> capture as well. As described earlier, the planar cell technology is expected to deliver significantly better techno-economic performance than the tubular cell technology in the longer term. Moreover, SOFC can maximize its potential for very low-cost CO<sub>2</sub> capture when it is highly integrated with advanced membrane technologies, e.g., OCM afterburner. To date, techno-economic assessments on SOFC with advanced CO<sub>2</sub> capture technologies has been performed only for power-only generation plants. In the long-term, SOFC systems may remain an option for large-scale CHP plants in industrial and residential heating applications. Therefore, it may be interesting to analyze the economic competitiveness of large-scale (planar) SOFC systems with advanced CO<sub>2</sub> capture in distributed power generation markets in the longer term.

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## 6

# Techno-economic prospects for CO<sub>2</sub> capture from distributed energy systems

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Submitted for publication.

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### Abstract

CO<sub>2</sub> emissions from distributed energy systems are expected to become increasingly significant, accounting for about 20% for current global energy-related CO<sub>2</sub> emissions in 2030. This article reviews, assesses and compares the techno-economic performance of CO<sub>2</sub> capture from distributed energy systems taking into account differences in timeframe, fuel type and energy plant type. The analysis includes the energy plant, CO<sub>2</sub> capture and compression, and distributed transport between the capture site and a trunk pipeline. Key parameters, e.g., capacity factor, energy prices and interest rate, were normalized for the performance comparison.

The findings of this study indicate that in the short-mid term (around 2020-2025), the energy penalty for CO<sub>2</sub> capture ranges between 23-30 % for coal-fired plants and 10-28 % for natural gas-fired plants. CO<sub>2</sub> avoidance costs are between 30-140 €/tonne for plant scales larger than 100 MW<sub>LHV</sub> (fuel input) and 50-150 € t<sup>-1</sup> for 10-100 MW<sub>LHV</sub>. In the long-term (2030 and beyond), the energy penalty for CO<sub>2</sub> capture might reduce to between 4% and 9% and the CO<sub>2</sub> avoidance costs to around 10-90 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 25-100 €/tonne for 10-100 MW<sub>LHV</sub> and 35-150 €/tonne for 10 MW<sub>LHV</sub> or smaller.

CO<sub>2</sub> compression and distributed transport costs are significant. For a distance of 30 km, 10 €/tonne was calculated for scales below 500 tCO<sub>2</sub>/day and more than 50 €/tonne for scales below 5 tCO<sub>2</sub>/day (equivalent to 1 MW<sub>LHV</sub> natural gas). CO<sub>2</sub> compression is responsible for the largest share of these costs.

CO<sub>2</sub> capture from distributed energy systems is not prohibitively expensive and has a significant cost reduction potential in the long term. Distributed CO<sub>2</sub> emission sources should also be considered for CCS, adding to the economies of scale of CO<sub>2</sub> transport and storage, and optimizing the deployment of CCS.

## 6.1 Introduction

CO<sub>2</sub> capture and storage (CCS) is considered to become an important technology to mitigate anthropogenic global warming. CCS is currently expected to be deployed in the power sector and the industry, both in which many large scale point sources are found (IEA, 2008a). Smaller scale energy systems, however, may account for a considerable fraction of total energy-related CO<sub>2</sub> emissions in the future. Distributed generation (DG) is expected to become increasingly important in the future energy supply infrastructure particularly in future electric utilities in the economies where deregulation takes place (Ogden, 2002). DG stations are generally smaller than 100-150 MW<sub>e</sub> (Ackermann et al., 2001) and Combined Heat and Power generation (CHP) is one of the major applications of DG due to its high overall energy efficiency. A study made by the International Energy Agency (IEA) estimates that in a scenario to half the global GHG emissions in 2050, compared to 2007 level (BLUE Map scenario), 30% of global electricity generation from fossil fuels would come from gas-fired CHP plants (IEA, 2010a). In this scenario, it is also estimated that more than a quarter of gas-fired CHP electricity generation will be from gas-fired CHP plants.

Distributed energy systems are generally not considered for CCS due to the relatively high costs at such scales (Damen et al., 2007; IPCC, 2005). However, there are a number of differences between DG plants and centralized power plants that may challenge this claim. Firstly, CHP plants can have heat integration with CO<sub>2</sub> capture process that centralized power plant cannot. Secondly, the operating conditions of DG plants are different from those for centralized power plants. Thirdly, literature indicates that CO<sub>2</sub> capture from distributed energy systems would become economical in the longer term. The IEA Greenhouse Gas R&D Programme (IEA GHG) performed a screening of technologies for “medium scale” (1-100 MW<sub>th</sub>) distributed energy systems and an order-of-magnitude analysis (OMA) for the economic performance of five selected technologies (IEA GHG, 2007a). The OMA results identified solid oxide fuel cells (SOFC) and oxyfuel coal boilers using oxygen conducting membranes (OCM) to be potentially promising. Damen et al. (2006) shows that CO<sub>2</sub> capture from a 20 MW<sub>e</sub> SOFC system may enable CO<sub>2</sub> avoidance cost as low as those for large-scale systems (600 MW<sub>e</sub>) in the long term. Therefore, it is important to quantify the performance development potential for CO<sub>2</sub> capture in distributed energy systems.

Till now, no study assessed and compared the performance of CO<sub>2</sub> capture in distributed energy systems taking into account the diversity in type of generator technologies used, applications, plant scales, operational patterns and applicable CO<sub>2</sub> capture technologies. There are a considerable number of CO<sub>2</sub> capture technologies proposed in the literature. A broad assessment is therefore desired to obtain better insights into the techno-economic possibilities for distributed CO<sub>2</sub> capture.

In addition, it is necessary to obtain insights into the minimum scale of CO<sub>2</sub> emission sources when CO<sub>2</sub> capture can be applied economically. Several studies on CCS use a 100 ktCO<sub>2</sub>/yr as the scale limit (e.g. Damen et al., 2009; IEA GHG, 2006; IPCC, 2005; van den Broek et al., 2010). The IPCC Special Report on CCS excludes emissions from sources smaller than 100 ktCO<sub>2</sub>/yr with the argument that their emissions represent only a small fraction of the total CO<sub>2</sub> emissions (IPCC, 2005). The 100 ktCO<sub>2</sub>/yr limit, therefore, seems to be a practical limitation and not a technical or economic limitation.

In this context, the objectives of this study are two-fold: (1) to provide an overview of techno-economic performance of CO<sub>2</sub> capture from distributed energy systems in the short-mid term (ST/MT) and the long term future (LT), which explicitly accounts for differences in timeframe, fuel type and energy plant type, and (2) to assess the relative cost of CO<sub>2</sub> capture from distributed generation compared to large-scale centralized power plants, taking into account the differences in plant scale, operational conditions, and the combination of CO<sub>2</sub> compression and distributed transport.

This paper is structured as follows. Section 2 provides an overview of distributed CO<sub>2</sub> emission sources and the type of technologies used for decentralized energy conversion systems. This is followed by a description of the key methodological aspects used in this study (Section 6.3). Sections 6.4 and 6.5 assess the techno-economic performance CO<sub>2</sub> capture technologies for decentralized emission sources for the ST/MT and the LT, respectively. For the ST/MT, the number of commercially feasible CO<sub>2</sub> capture technologies is limited. We investigated how the energy penalty for a particular ST/MT CO<sub>2</sub> capture technology would differ by the type and application of energy systems. In the LT, in contrast, there is a variety of advanced technologies and concepts proposed in the literature. We therefore put emphasis on the assessment of the technology itself. Section 6 combines the results from Sections 6.4 and 6.5 to assess the technological and economic performance improvement potential from ST/MT to LT. Section 6.7 assesses the effect of CO<sub>2</sub> capture scale on compression and distributed transport costs. Section 6.8 discusses the limitations of the study while conclusions are drawn in Section 6.9.

## **6.2 CO<sub>2</sub> emissions from distributed energy systems**

In this research, three main types of distributed energy systems are considered: CHP plants, boilers and distributed hydrogen plants. This section briefly describes the technology, applications, and global CO<sub>2</sub> emissions of each type. Note that while this study will include the three technologies as far as the availability of data allows it, the focus will be on CHP plants.

**Table 6-1:** Overview of combined heat and power (CHP) plant technologies. Source: EPA (2008), IEA (2008b), IEA GHG (2007a) and IPCC (2005). Energy efficiency values are in LHV terms.

CHP system	Typical capacity [MW <sub>e</sub> ]	Power efficiency <sup>1)</sup>	Overall efficiency <sup>1)</sup>	Typical HPR	Installed cost [€/kW <sub>e</sub> ]	O&M costs [€/kWh <sub>e</sub> ]	Fuel types	Uses for thermal output	Exhaust gas CO <sub>2</sub> concentration (v/v, dry)
Gas turbine	0.5-40 (NGCC: ~250)	24-40%	77-83%	0.5-2	710-950 (5-40 MW <sub>e</sub> )	0.003-0.01	NG, biogas, propane, oil	Heat, hot water, LP-HP steam	3-4% (NG)
Reciprocating engine	0.01-5	24-44%	77-88%	1-2	800-1600	0.007-0.02	NG, biogas, propane, landfill gas	Hot water, LP steam	3-4% (NG)
Steam turbine	0.5-250	16-42%	84-88%	3-10	310-800	<0.004	All types	LP-HP steam	12-14% (coal)
Fuel Cells	0.005-2	33-66%	72-88%	0.5-1	370-4800	0.02-0.03	H <sub>2</sub> , NG, propane, methanol	Hot water, LP-HP steam	System dependent <sup>3)</sup>

<sup>1)</sup> Efficiency values reported in high heating value (HHV) are converted to low heating value (LHV) terms using a multiplication factor of 1.05 for coal and 1.1 for other fuels.

<sup>2)</sup> IEA GHG (2007a) presents a higher range (9-14%), but we consider the range an overestimation. Gas engine exhaust gas typically contains 11-16% O<sub>2</sub> (IPPC, 2006), leading to 3-5 vol% (dry) CO<sub>2</sub>.

<sup>3)</sup> Anode off-gas of high temperature fuel cells (e.g. solid oxide fuel cells (SOFC) and molten carbonate fuel cells (MCFC)) is CO<sub>2</sub>-rich when fuels rich in hydrocarbons or carbon monoxide are used. However, the anode off-gas contains some unoxidized fuel that it cannot be compressed and transported for underground storage. In conventional SOFC and MCFC systems, anode off-gases are fully oxidized with air. Therefore, CO<sub>2</sub> concentrations of plant exhaust gases are low; only 2-3 vol% (dry) for SOFC (Campanari, 2001; Kuramochi et al., 2011 (Chapter 5)) and around 7 vol% (dry) for MCFC (Campanari, 2002). IEA GHG (2007a) presents somewhat higher values (8-10%), but the report does not provide references.

### **6.2.1 CHP plants**

The main advantage of CHP is the high overall energy efficiency due to the simultaneous production of useful thermal and electrical energy (EPA, 2008). The type of installed energy systems depends on the heat demand and the quality of heat product required. Table 6-1 presents an overview of heat demands by sector and the applicability of various combined heat and power (CHP) generator types.

While CHP enables large energy savings and CO<sub>2</sub> emissions reduction, its CO<sub>2</sub> emissions are still significant. CHP accounts for more than 6 EJ<sub>e</sub>/yr, or more than 10% of global electricity generation today (IEA, 2007a), and is expected to increase considerably in the coming decades. An IEA study suggests that in G8+5 countries<sup>20</sup>, which account for more than two-thirds of global primary energy consumption, the share of CHP in electricity generation may increase from 11% in 2005 to 24% in 2030 in a scenario with a pro-CHP policy regime (IEA, 2008b). Assuming the same share worldwide, CHP would account for about 23EJ<sub>e</sub>/yr<sup>21</sup> and over 3 GtCO<sub>2</sub> emissions per year using conventional technology and conservative assumptions<sup>22</sup>.

### **6.2.2 Boilers**

The IEA GHG (2007a) estimates that boilers with a fuel input of 1-100 MW<sub>th</sub><sup>23</sup>, which corresponds to the typical scale of district heating (DH), industrial and large commercial installations, account for about 9% of global energy-related emissions (more than 2 GtCO<sub>2</sub>/yr). The majority of emissions is attributable to coal boilers in China (IEA GHG, 2007a). Assuming that global CO<sub>2</sub> emissions from boilers remain constant, distributed energy systems (CHP plants and boilers combined) in 2030 could account for about 20% of current global energy-related CO<sub>2</sub> emissions<sup>24</sup>.

### **6.2.3 Distributed hydrogen plants**

Hydrogen may gradually take over the function of natural gas in the residential sector in the longer term (Damen et al., 2007). Barreto et al. (2003) present an future energy scenario in which CHP in industrial and residential stationary fuel cell applications and electricity generation from mobile hydrogen-based fuel cells in the transportation sector together account for nearly 40% of the global electricity generation in 2100. Some fuel

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<sup>20</sup> G8+5 countries include G8 nations (Canada, France, Germany, Italy, Japan, Russia, the United Kingdom and the United States) plus Brazil, China, India, Mexico and South Africa.

<sup>21</sup> The CHP potentials reported in IEA (2008b) are based on the global electricity generation figures of the Alternative Policy Scenario (APS) in the World Energy Outlook 2007 (IEA, 2007b). APS assumes 12% reduction in global electricity generation in 2030 compared to the reference scenario, which estimates 29737 TWh/yr. Therefore, global electricity generation in 2030 under APS is 94EJ<sub>e</sub>/yr. Hence, 94\*24%=22.6 EJ<sub>e</sub>/yr.

<sup>22</sup> All CHP plants around the world were assumed to be natural gas-fired with an electrical efficiency of 40%.

<sup>23</sup> A 100 MW<sub>th</sub> coal fired plant operating for 7,500 hours/year would produce about 250,000 tonnes/year of CO<sub>2</sub> and a natural gas fired plant would produce about 150,000 t/yr, assuming an emission factor of 95 kg/GJ<sub>LHV</sub> and 56 kg/GJ<sub>LHV</sub>, respectively (IEA GHG, 2005).

<sup>24</sup> Note that the increased penetration of CHP would affect global boiler capacity. For example, CO<sub>2</sub> emissions from Chinese coal boilers may reduce drastically in the future because the Chinese government is closing down coal-fired power plants with capacities smaller than 100 MW<sub>e</sub> to reduce coal consumption and CO<sub>2</sub> emissions (Yu et al., In Press)

cell technologies may become economical for CHP on a house/building (block) level for newly build projects (Damen et al., 2007).

In addition, a significant development of hydrogen markets might be seen if society takes a path for hydrogen-powered vehicles, e.g. fuel cell vehicles. In this case, large-scale hydrogen plants using fossil feedstock are considered the most economical, but the cost of building a distribution infrastructure and the need for sufficient demand could be major barriers. Therefore, distributed hydrogen generation at refueling stations using the existing natural gas infrastructure may be considered (Sjardin et al., 2006).

### 6.3 Research methods

In this study, in principle only newly built plants were considered. Nevertheless, performance data for retrofit cases reported in the literature were also used when they prove to be useful for this section.

In this study, we reviewed literature that investigated the techno-economic performance of CO<sub>2</sub> capture from energy systems of scales up to about 150 MW<sub>th</sub> fuel input. This scale limit covers “medium scale” DG systems, which is defined to be of 5-50 MW<sub>e</sub> scale by Ackermann et al. (2001).

Regarding the economic performance data available in the literature, we only considered those from studies published after 2000 to take into account the latest developments in CO<sub>2</sub> capture R&D and to avoid economic data which are outdated.

The system boundaries defined in this study included the following components: energy conversion plant, CO<sub>2</sub> capture and compression, and distributed CO<sub>2</sub> transport between the CO<sub>2</sub> capture site and trunk CO<sub>2</sub> pipeline. All plants were assumed to be newly-built and trunk CO<sub>2</sub> pipelines were assumed to be already available. The CO<sub>2</sub> capture technologies assessed in this study were categorized into ST/MT technologies (10-15 years) and LT (20 years or more) technologies. ST/MT technologies are defined as those that are either in pilot, demonstration or commercialization phase today (Peeters et al., 2007). The technologies are categorized as ST/MT also when all required components are commercially available today, even if the process as a whole has not been tested or demonstrated. All other technologies, either in modeling or laboratory phase today, are considered to be LT options.

#### 6.3.1 Technical performance indicator

We used the energy penalty of an energy plant type  $i$  (condensing power plant, CHP, H<sub>2</sub>, boiler, furnace) due to CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ : dimensionless) as an indicator of technical performance of CO<sub>2</sub> capture, which is expressed as:



$$\Delta\eta_{pen} = 1 - \frac{\eta_{CC}}{\eta_{Ref}} \quad (6.1)$$

where  $\eta$  is the conversion efficiency of the plant. Subscripts *CC* stands for the plant with CO<sub>2</sub> capture and *Ref* for the reference plant, respectively. For CHP plants,  $\eta$  is calculated on exergy terms (adapted from Philipsen et al. (1998)):

$$\eta = \frac{E_{el,out} + E_{th,out} * f}{E_{PE}} \quad (6.2)$$

where  $E_{el,out}$  is the net electrical output (MW<sub>e</sub>),  $E_{th,out}$  is the net heat product output (MW<sub>th</sub>),  $f$  is the exergy factor for heat, and  $E_{PE}$  is the primary energy input (MW<sub>LHV</sub>).

The exergy content method provides a thermodynamically acceptable allocation base between steam and electricity, acknowledging the higher usefulness of electricity (Svending, 2003). The exergy-based allocation is also suggested to be the most meaningful and accurate among various methods for allocating CO<sub>2</sub> emissions for cogeneration processes (Rosen, 2008).

In this study, the energy performance of DH CHP plants was calculated using the data for condensing plants, taking into account various opportunities for heat recovery. Figure 6-1 shows a simplified diagram of a district heating CHP plant with CO<sub>2</sub> capture and compression, and possible heat recovery opportunities. The energy performance of DH CHP plants with and without CO<sub>2</sub> capture were compared on leveled heat and power outputs.  $E_{el,out}$  and  $E_{th,out}$  were calculated as follows:

$$E_{el,out} = E_{el,cond} - E_{el,comp} - E_{el,CC} - E_{th,CC} * f_{CC} - E_{th,DH} * f_{DH} \quad (6.3)$$

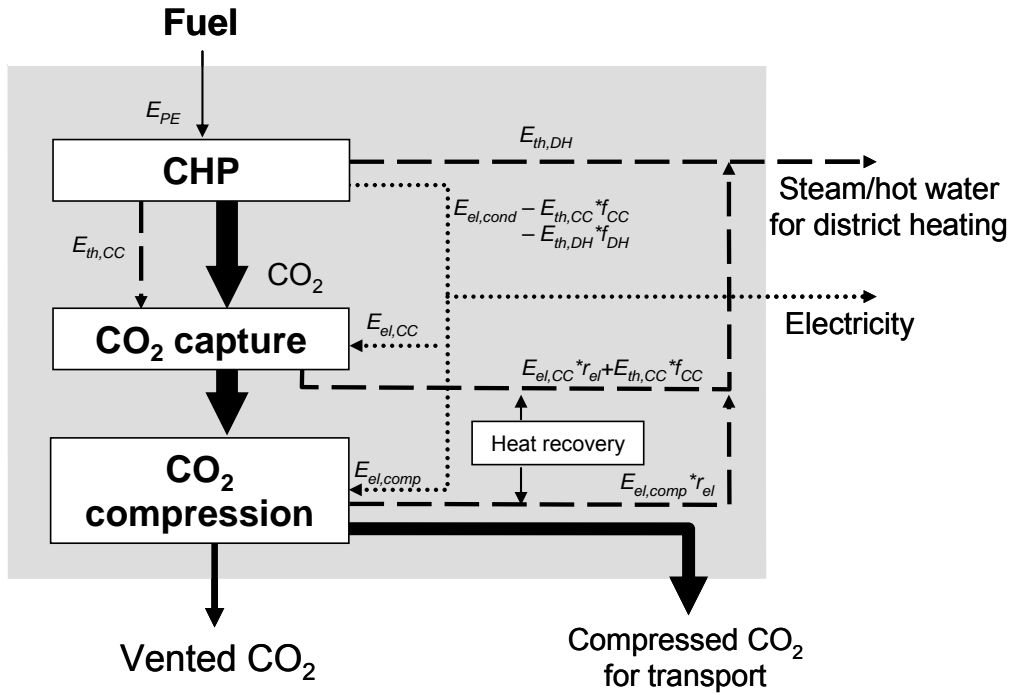
$$E_{th,out} = E_{th,DH} + E_{th,CC} * r_{th} + (E_{el,CC} + E_{el,comp}) * r_{el} \quad (6.4)$$

where  $E_{el,cond}$  is the net power output at condensing mode (power-only mode) (MW<sub>e</sub>),  $E_{el,comp}$  is the power consumption for CO<sub>2</sub> compression (MW<sub>e</sub>),  $E_{el,CC}$  is the power consumption for CO<sub>2</sub> capture (MW<sub>e</sub>),  $E_{th,CC}$  is the heat extracted for CO<sub>2</sub> capture (MW<sub>th</sub>),  $f_{CC}$  is the exergy factor for CO<sub>2</sub> capture heat (dimensionless),  $E_{th,DH}$  is the heat extracted for DH (MW<sub>th</sub>),  $r_{th}$  is the fraction of CO<sub>2</sub> capture heat recovered for DH, and  $r_{el}$  is the fraction of CO<sub>2</sub> capture and compression power consumption recovered as heat for DH.

For boilers,  $\eta$  is calculated as:

$$\eta = \frac{E_{th,out}}{E_{PE} + \frac{E_{el,in}}{\eta_{grid}}} \quad (6.5)$$

where  $E_{el,in}$  is the net electricity consumption ( $MW_e$ ), and  $\eta_{grid}$  is the centralized electricity generation efficiency.



**Figure 6-1:** District heating CHP plant with CO<sub>2</sub> capture and compression. The figure also shows possible heat recovery opportunities.

For hydrogen plants,  $\eta$  is calculated based on primary energy terms as proposed in Damen et al. (2006):

$$\eta = \frac{E_{H_2} + \frac{E_{el,out}}{\eta_{grid}} + \frac{E_{th,out}}{\eta_{boiler}}}{E_{PE} + \frac{E_{el,in}}{\eta_{grid}} + \frac{E_{th,in}}{\eta_{boiler}}} \quad (6.6)$$

where  $E_{th,in}$  is the net steam consumption ( $MW_{th}$ ),  $\eta_{boiler}$  is the boiler efficiency (90%).

Table 6-2 presents exergy factors for various heat qualities, including centralized electricity generation efficiency ( $\eta_{grid}$ ). The exergy factors for heat product,  $f$ , are differentiated by the temperature of the heat product.

**Table 6-2:** Exergy factors for various heat products and centralized electricity generation efficiency

<i>Parameter</i>	<i>Symbol</i>	<i>Value</i>
Exergy factor for heat (f: J <sub>ex</sub> /J <sub>th</sub> ) <sup>1)</sup>		
District heating (steam or hot water)	f <sub>DH</sub>	0.17
CO <sub>2</sub> capture solvent regeneration heat (low temperature steam)	f <sub>CC</sub>	0.22
Industrial process heat (medium-high temperature steam)	f <sub>Ind</sub>	0.35
Centralized electricity generation efficiency	η <sub>grid</sub>	0.5

<sup>1)</sup> DH, CO<sub>2</sub> capture solvent regeneration heat and industrial steam were assumed to be extracted at 110 °C, 140 °C, and 240 °C, respectively. The exergy factors were taken from a figure in Bolland and Undrum (2003).

### 6.3.2 Economic performance indicator

In this study, we used CO<sub>2</sub> avoidance cost ( $C_{CO_2}$ : €/tonne) as the main economic performance indicator. CO<sub>2</sub> avoidance cost was calculated based on cost of an energy carrier ( $CE$ : €/GJ):

$$C_{CO_2} = \frac{CE_{CC} - CE_{Ref}}{Em_{Ref} - Em_{CC}} \quad (6.7)$$

where  $Em$  is the specific CO<sub>2</sub> emission intensity of an energy carrier (tCO<sub>2</sub>/GJ). For CHP plants, the  $CE$  depends largely on how costs are allocated to the electricity and heat that are produced. In this study, we used the exergy factor of final energy carriers to allocate costs. Regarding  $Em$ , not only onsite CO<sub>2</sub> emissions related to fossil fuel use but also CO<sub>2</sub> emissions related to electricity consumption were also taken into account.  $CE$  and  $Em$  are calculated using the standardized parameters presented in the following section (6.3.3.1).

### 6.3.3 Normalization of parameters

All cost figures were converted to €<sub>2007</sub>. Inflation and material price increases for the energy systems and CO<sub>2</sub> capture process plants were accounted for by applying the CERA European Power Capital Cost Index (EPCCI) (IHS CERA, 2011). Costs that are reported in US\$ were first converted to Euro of the original year using the exchange rate data (year average) from OANDA (2008), then updated to €<sub>2007</sub>. The Chemical Engineering Plant Cost Index (CEPCI; Chemical Engineering, 2010) was used for distributed CO<sub>2</sub> transport components. Other cost figures were corrected for inflation by using GDP deflators (IMF, 2009). Note that for advanced CO<sub>2</sub> capture technologies, technical performance figures are forecasts and cost projections are uncertain.

With regard to capital investment, we considered total capital requirement (TCR), which includes the following components:

- Process plant cost (costs for the equipment pieces and their installation) plus engineering fees and contingencies;

- Owner costs (royalties, preproduction costs, inventory capital, land costs and site preparation) and interests during construction

### 6.3.3.1 Key parameters

Table 2-2 presents the parameters that are normalized for technical and economic performance calculations throughout the study. High and low parameter values were used to assess the sensitivity of the results (see Section 6.3.4).

**Table 6-3:** Parameters normalized for technical and economic performance calculations in this study.

<i>Parameters and variables</i>	<i>Unit</i>	<i>Nominal</i>	<i>Low</i>	<i>High</i>
Real interest rate	%	10	7.5	12.5
Economic lifetime of plants <sup>1)</sup>				
Coal-fired installations	yr	35	30	40
Gas-fired installations (conventional technologies)	yr	25	20	30
Membrane-based technologies	yr	20	15	25
Annual operation time				
Industrial installations	h/yr	7500	6000	8500
Centralized power plants	h/yr	7000	6000	8000
Other installations <sup>2)</sup>	h/yr	5000	3500	6500
Total capital requirement <sup>3)</sup>	% - total plant cost	110	105	130
Total plant cost <sup>3)</sup>	% - process plant cost	130	120	150
Energy prices				
Coal <sup>4)</sup>	€/GJ <sub>LHV</sub>	2.6	2	3.2
Natural gas <sup>4)</sup>	€/GJ <sub>LHV</sub>	8	5	11
Electricity <sup>5)</sup>	€/MWh	55	40	70
CO <sub>2</sub> emission factors				
Natural gas <sup>6)</sup>	g/MJ <sub>LHV</sub>	56	-	-
Coal <sup>6)</sup>	g/MJ <sub>LHV</sub>	95	-	-
Grid electricity (Em <sub>Sp,Elec</sub> ) <sup>7)</sup>	g/MJ <sub>e</sub>	111	89	133

<sup>1)</sup> Lifetime for membrane-based technologies, e.g. fuel cells and membrane reactors, are based on (Fontell et al., 2004; Ma et al., 2009).

<sup>2)</sup> "Other installations" include district heating, institutional and commercial installations.

<sup>3)</sup> Process plant cost (PPC) comprises equipment cost and installation costs. Total plant cost (TPC) comprises PPC and engineering fees and contingencies. Total capital requirement (TCR) comprises TPC, owner costs and interests during construction. The values are within the ranges observed for power plant construction (van Horssen et al., 2009).

<sup>4)</sup> Nominal value is from IEA (2010b). The high and low values assumed here agree with those forecasted by the IEA for the EU, the US and Japan for years between 2020 and 2030 (IEA, 2010b).

<sup>5)</sup> Electricity price for large industries differs significantly by country, from 0.028 €/kWh in Russia (in 2006) to 0.177 €/kWh in Italy (in 2007). The price used in this study is similar to that in the USA (0.048 €/kWh in 2007), South Korea (0.052 €/kWh in 2007) and Poland (0.061 €/kWh in 2007) (US DOE, 2010b).

<sup>6)</sup> From (IPCC, 2006).

<sup>7)</sup> The changes in electrical consumption due to CO<sub>2</sub> capture are likely to affect base load power generation by base-load fossil fuel-fired power plants. We assumed that in the industrialized world where CCS is also deployed for industrial processes, around 40% of base-load fossil fuel-fired power plants are equipped with CO<sub>2</sub> capture. Assuming an average 75% CO<sub>2</sub> avoidance rate by CO<sub>2</sub> capture, we estimated that the CO<sub>2</sub> emissions from base-load fossil fuel-fired power plants would be reduced by 30% due to CO<sub>2</sub>

capture. The nominal value assumes that natural gas- and coal-fired power plants share the electricity generation 50% each. The low end value corresponds to a ratio of 80:20 between natural gas power plants and coal power plants, and the high end value corresponds to a ratio of 20:80.

### 6.3.3.2 Normalization of CO<sub>2</sub> compression, distributed transport and CO<sub>2</sub> purity

#### CO<sub>2</sub> compression

CO<sub>2</sub> is compressed to 110 bar for pipeline transport in a supercritical phase. CO<sub>2</sub> compression is performed in two steps (2006). Firstly, CO<sub>2</sub> is compressed using a multistage compressor up to the critical pressure of CO<sub>2</sub> (7.38 MPa), followed by a further compression using a pump for liquid/supercritical phase. When electricity consumption for CO<sub>2</sub> compression is not reported in the original literature, specific power consumption is estimated using the following equation adapted from Koornneef et al. (2008):

$$E_{s,comp} = \frac{ZRT_1}{M\eta_{is}\eta_m} \frac{N\gamma}{\gamma-1} \left\{ \left( \frac{p_c}{p_l} \right)^{\gamma-1/N\gamma} - 1 \right\} + \frac{p_{out} - p_c}{\rho\eta_p} \quad (6.8)$$

where  $E_{s,comp}$  is the specific electricity requirement (kJ/kgCO<sub>2</sub>),  $Z$  is the average CO<sub>2</sub> compressibility factor<sup>25</sup> (0.89),  $R$  is the universal gas constant (8.3145 J/(mole K)),  $T_1$  is the suction temperature (303.15 K),  $\gamma$  is the specific heat ratio ( $c_p/c_v$ ) (1.294),  $M$  is the molar mass (44.01 g/mole),  $\eta_{is}$  is the isentropic efficiency (80%),  $\eta_m$  is the mechanical efficiency (99%),  $p_l$  is the suction pressure (101 kPa),  $p_c$  is the critical pressure (7,380 kPa),  $p_l$  is the outlet pressure (110 bar = 11,000 kPa),  $N$  is the number of compressor stages (= 4),  $\rho$  is the density of CO<sub>2</sub> during pumping (630 kg/m<sup>3</sup>), and  $\eta_p$  is the pump efficiency (75%). When the compression pressure reported in the literature differs from the value used in this study (110 bar), we also adjusted the specific electricity consumption using Eq.(6.8). CO<sub>2</sub> compressor cost was calculated using the parameters presented in Table 6-4. The CO<sub>2</sub> compression system capital cost (installation cost) was calculated using a regression curve fitted to data from the following sources (Kreutz et al., 2005; Lokurlu et al., 2004; van Bergen et al., 2003; Wong et al., 2001):

$$C_{comp} = 2.7 * E_{el,comp}^{0.53} \quad (n=4, R^2 = 0.98, 0.11 < E_{el,comp} < 13) \quad (6.9)$$

where:

$C_{comp}$ : Capital cost for the CO<sub>2</sub> compressor (M €),

$E_{el,comp}$ : Power consumption for CO<sub>2</sub> compression (MW<sub>e</sub>).

n: number of CO<sub>2</sub> compression system capital cost data

R<sup>2</sup>: coefficient of determination.

<sup>25</sup> The derived value is the average of the compressibility factor values calculated for the average pressure of each compressor stage (four stages in total). Compressibility factors for each compression stage were calculated using the Peng-Robinson equation of state programmed by ChemSOF (2011).

CO<sub>2</sub> compression may have a significant impact on the economic performance of smaller scale CO<sub>2</sub> capture (IEA GHG, 2007a). We therefore looked carefully into the CO<sub>2</sub> compressor costs reported in the literature reviewed in this study. If those assumed in the literature were considered too low, we recalculated the compressor capital costs using Eq.(6.9).

**Table 6-4:** Parameters for CO<sub>2</sub> compression and distributed CO<sub>2</sub> transport cost calculations.

<i>Parameter</i>	<i>Unit</i>	<i>Value</i>
Truck transport cost <sup>1)</sup>	€ <sub>2007</sub> /(t km)	0.22
CO <sub>2</sub> recompression system capital cost <sup>2)</sup>	M€ <sub>2007</sub>	1.2* compression power (MW <sub>c</sub> ) + 0.07
CO <sub>2</sub> storage tank cost (200 tCO <sub>2</sub> capacity) <sup>3)</sup>	k€ <sub>2007</sub>	150
O&M cost: pipeline and CO <sub>2</sub> storage tank <sup>2)</sup>	%-capital cost	2

<sup>1)</sup> From van Bergen et al. (2003). The values reported in the literature range between 0.04 €/t km (Midwest Geological Sequestration Consortium, 2003) and 0.32 €/t km (Klein et al., 2003). Although it is not explicit in the literature, the low estimate seems to take into account only the truck operation cost (driver and fuel), and the high estimate seems to include CO<sub>2</sub> liquefaction. Values that lie within the range are reported in Wong (2005), Herzog and Golomb (2004) and MGSC (Midwest Geological Sequestration Consortium, 2003).

Because of the assumed transport distance, CO<sub>2</sub> boil-off during transport and CO<sub>2</sub> emissions from transport trucks were assumed to be negligible. The CO<sub>2</sub> boil-off rate can be as high as 10% depending on the length of time the CO<sub>2</sub> is kept in the truck (Wong, 2005). For ocean tankers, a boil-off rate of 1% per day is suggested (Heddle et al., 2003).

<sup>2)</sup> From McCollum and Ogden (2006). The minimum storage capacity was assumed to be 20 tonne, which is equivalent to a typical capacity of CO<sub>2</sub> transport trucks.

<sup>3)</sup> From Wong (2005). The scaling factor was assumed to be 0.6. The minimum tank capacity was assumed to be 20 tCO<sub>2</sub>, equivalent to the truck transport capacity.

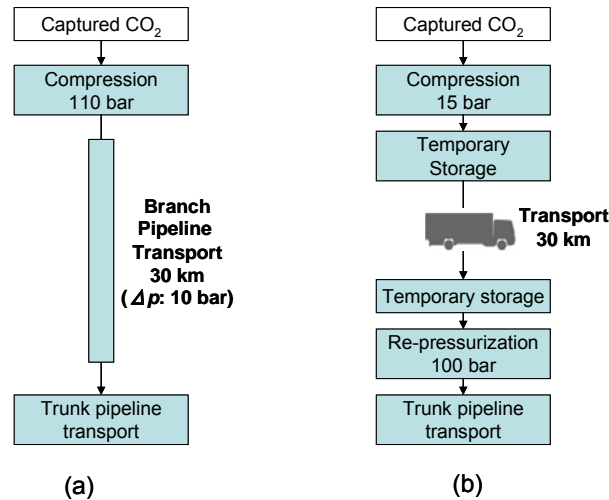
An additional CO<sub>2</sub> purification process was assumed in case CO<sub>2</sub> purity reported in the literature is lower than 95 vol%<sup>26</sup>. In such cases, CO<sub>2</sub> capture rate was adjusted by a multiplication factor  $\eta_{Rec}$  as some CO<sub>2</sub> would be vented together with the removed impurities in the purification process.  $\eta_{Rec}$  was assumed 90%, 92% and 94% for CO<sub>2</sub> purities below 75 vol%, between 75-80 vol%, and above 80 vol%, respectively (Ritter et al., 2009).

#### *Distributed CO<sub>2</sub> transport between the CO<sub>2</sub> capture site and the trunk pipeline*

Literature indicates that the use of trucks maybe more suitable than pipelines for small-scale and short-distance CO<sub>2</sub> transport (Herzog and Golomb, 2004; Wong, 2005). We therefore considered both branch pipeline and truck as options for distributed CO<sub>2</sub> transport between the CO<sub>2</sub> capture site and a trunk pipeline (see Figure 6-2). In each case, the cheapest option was selected. The parameters used for the calculations are presented in Table 6-4. A distance of 30 km was assumed and we did not consider the networking of clustered distributed energy systems for collective CO<sub>2</sub> capture. Note that these

<sup>26</sup> 95 vol% is a typical concentration at which existing CO<sub>2</sub> pipelines operate (de Visser et al., 2008).

assumptions are conservative for industrial areas or residential areas in which CO<sub>2</sub> capture-distributed energy systems are expected to be installed<sup>27</sup>.



**Figure 6-2:** Distributed CO<sub>2</sub> transport options considered in this study

For branch pipeline transport cost calculations, the model presented in IEA GHG (2002) was used (see also Chapter 4<sup>28</sup>). CO<sub>2</sub> is transported in trucks in liquid phase, typically at 17 bar, -30 °C (Wong, 2005)<sup>29</sup>. The CO<sub>2</sub> delivered by truck is re-pressurized for trunk pipeline transport. In this study, we assumed that total CO<sub>2</sub> compression cost (onsite compression and re-compression at the trunk pipeline inlet) for truck transport is equivalent to that for onsite CO<sub>2</sub> recompression to 100 bar. This assumption can be justified because CO<sub>2</sub> compression is performed in multiple stages (often more than four) and there would be limited effect of scale on capital costs and electricity consumption.

### 6.3.4 Sensitivity analysis

A sensitivity analysis was conducted for the following parameters: energy prices, plant lifetime, interest rate, capital cost, and grid electricity CO<sub>2</sub> emission factor (see Table 2-2). To assess the combined effect of the parameters on the results, we applied a general equation for uncertainty propagation as described in, e.g., (NIST/SEMATECH, 2011; Taylor, 1997). In this approach, for a given indicator  $C$ , which is a function of variables  $X_1, X_2, \dots, X_n$ , the standard deviation of  $C$  ( $\sigma_C$ ) is calculated as follows:

<sup>27</sup> Large industrial or urban areas in Europe where CCS is considered, e.g., Randstad (includes Amsterdam and Rotterdam) in the Netherlands and the Merseyside and Deeside Basin (includes Liverpool) in the UK, fit within a radius of 30-40 km (Damen et al., 2007; IEA GHG, 2007b).

<sup>28</sup> Except for a terrain factor which was reduced from 2 to 1.4 based on the argument in van den Broek et al.(2010).

<sup>29</sup> This means that captured CO<sub>2</sub> is not compressed to 110 bar as assumed in the previous section. Nevertheless, studies indicate that specific power consumption and capital cost are comparable for liquefaction at low pressures (15-20 bar) and at critical pressure. Studies on ship transport at 50 °C and -7 to -6.5 bar report specific electricity consumption of 110-123 kWh/tCO<sub>2</sub> (Aspelund et al., 2004; IEA GHG, 2004b). The reported capital cost is 80M\$<sub>2004</sub> for 102 MW<sub>e</sub> power, which is in agreement with the capital cost equation for the systems for 110 bar (Eq.6.8). Therefore, we assumed that the onsite electricity consumption calculated for branch pipeline transport is also valid for truck transport.

$$\sigma_C = \sqrt{\sum_{i=1}^n \left( \frac{\partial C}{\partial X_i} \right)^2 \sigma_{X_i}^2 + 2 \sum_{i=1}^{n-1} \sum_{j=i+1}^n \frac{\partial C}{\partial X_i} \frac{\partial C}{\partial X_j} \sigma_{X_i} \sigma_{X_j} r_{ij}} \quad (6.10)$$

where  $\sigma_{X_i}$  is the standard deviation of variable  $X_i$ ,  $r_{ij}$  is the correlation coefficient between two variables  $X_i$  and  $X_j$ . Because there is no detailed information on the distribution of the parameter values, we assumed a uniform distribution. In this case,  $\sigma_{X_i}$  is calculated as (NIST/SEMATECH, 2011):

$$\sigma_{X_i} = \frac{X_{i,High} - X_{i,Low}}{\sqrt{12}} \quad (6.11)$$

The correlation coefficients among variables were not considered in this study.

## 6.4 Assessment of short-mid term CO<sub>2</sub> capture technologies

As noted earlier, we investigated how the energy penalty for a particular ST/MT CO<sub>2</sub> capture technology would differ by the type and application of energy systems. Then we looked at a number of promising CO<sub>2</sub> capture options for distributed energy systems that have not yet been covered in earlier studies.

### 6.4.1 Literature review and selection of short term technologies

Table 6-5 shows the key performance parameters for ST/MT CO<sub>2</sub> capture from small-medium scale emission sources reported in the literature. Most of the data are for industrial applications.

#### 6.4.1.1 Post-combustion capture

Post-combustion capture in the ST/MT would most likely apply a chemical absorption process. Chemical absorption is more favorable for power plants with steam turbines than for boilers and simple gas turbine power plants because steam turbines can minimize the exergy losses due to the extraction of steam for CO<sub>2</sub> capture. In case of capture from natural gas combustion flue gas, MEA-based process consumes around 20-30% of total thermal energy input for solvent regeneration. For boilers, this directly leads to  $\Delta\eta_{pen}$  of 20-30% (Table 6-5). For NGCC plants, however,  $\Delta\eta_{pen}$  is much smaller (e.g., IEA GHG (2004a) reports 7.5%) because the MEA process requires low-temperature steam, which has low exergy content.

#### *District heating CHP plants*

Regarding DH plants, the study by Desideri and Corbelli (1998) on a 5 MW<sub>e</sub> steam-injection gas turbine plant is, to the authors' knowledge, the only one available in the public literature that assesses the techno-economic feasibility of retrofitting CO<sub>2</sub> capture to smaller scale DH CHP plants. The results show that the energy penalty differs significantly by the operational mode of the CHP plant (Desideri and Corbelli, 1998). There are a number of studies that investigate the integration between large scale DH CHP



plants and CO<sub>2</sub> capture (e.g., IEA GHG, 1999a; Knuutila et al., 2009). These studies suggest that chemical absorption CO<sub>2</sub> capture may be energetically more economical for DG-CHP plants than for condensing power plants because of better heat integration possibilities.

In the chemical absorption CO<sub>2</sub> capture process, low-grade heat can be recovered from flue gas cooling, CO<sub>2</sub> regenerator condenser, and lean solvent cooling. Regarding flue gas cooling, flue gas has to be cooled down from about 110 °C to about 50 °C before it enters the CO<sub>2</sub> scrubber. Heat is not recovered in condensing power plants because the temperature is too low for the low-pressure evaporator (Lucquiaud et al., 2009), but it can still be used to generate hot water.

More heat can be recovered from the CO<sub>2</sub> regenerator condenser. CO<sub>2</sub> is released at the top of the stripper with considerable amount of steam. In condensing power plants, the steam is condensed and separated from CO<sub>2</sub> in a water cooler. The IEA GHG (1999a) has shown that in case of NGCC plants, about 30% of the thermal energy extracted from the steam turbines for CO<sub>2</sub> capture can be recovered at the CO<sub>2</sub> regenerator condenser in the form of hot water<sup>30</sup> when the DH water returns at 40 °C. Knuutila et al. (2009) and Gode and Hagberg (2008) report a similar recovery rate<sup>31</sup>. It is suggested that this heat recovery may be done with negative incremental capital cost because the direct contact cooling tower is replaced by additional heat recovery steam generator surface, the cost of which is marginal (IEA GHG, 1999a).

A considerable amount of heat is released in the lean amine cooling process. The solvent regenerated in the CO<sub>2</sub> stripper has to be cooled down from 100-140 °C to 40-50 °C to be fed into the CO<sub>2</sub> absorber via the lean-rich heat exchanger and an additional cooler. Heat recovery from the additional cooling, however, is unrealistic because the temperature difference between the cooler inlet and outlet is small (IEA GHG (1999a) assumes 27 °C).

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<sup>30</sup> 55MW<sub>th</sub> can be recovered for a plant with 854 MW<sub>th</sub> fuel input and 90% CO<sub>2</sub> capture. Assuming a CO<sub>2</sub> emission factor of 56 g/MJ, specific heat recovery is calculated to be 1.28 GJ/tCO<sub>2</sub> captured. The IEA GHG study assumes specific heat consumption of 4.22 GJ/tCO<sub>2</sub> captured, thus 1.28/4.22 = 0.303.

<sup>31</sup> The figure from the Knuutila study also includes heat recovery from CO<sub>2</sub> compression, but we considered this fraction in total heat recovered to be small. 107 MW of hot water can be generated at 108 kgCO<sub>2</sub>/s capture rate and a specific heat consumption of 3.22 GJ/tCO<sub>2</sub> captured. The heat recovery rate is calculated to be 30.7%. Bode and Hagberg (2008) report that the electrical efficiency reduces by 11.4%-points while DH efficiency increases by 8.9%-points. Using the exergy factor ( $f_{cc}=0.22$ ) presented in Table 6-2 and assuming that CO<sub>2</sub> compression accounts for 3.6%-point electrical efficiency reduction (based on specific electricity consumption of 0.4GJ<sub>e</sub>/tCO<sub>2</sub>), the heat recovery rate is 8.9/((11.4-3.6)/0.22) = 25.1%.

**Table 6-5:** Energy, economic and CO<sub>2</sub> performance data for small-medium scale installations with short-mid term CO<sub>2</sub> capture technologies reported in the literature

DG category	Fuel type	Generator type	Plant scale [MW]	CO <sub>2</sub> capture technology	CO <sub>2</sub> capture efficiency	System energy efficiency ( $\eta$ )	Energy penalty ( $\Delta\eta$ )	Capital cost [€ <sub>2007</sub> /kW]	O&M cost [%-Cap. cost]	Reference
Condensing/ CHP plants	NG	GT-CHP	5	Chemical absorption	68-90%	26-32%	19-25%	3800 <sup>1)</sup>	7.3	Desideri and Corbelli (1998)
			5	Pre-combustion	85%	41%	19%	2000	2.6	IEA GHG (2007a)
			50	Chemical absorption	90%	46-53%	12-18%	1500-1600	4	Kuramochi et al. (2010: Chapter 4)
Boilers	NG	Boiler	140	Chemical absorption	86%	70%	28%	400	2.9	Switzer et al. (2005)
	Coal	Boiler	50	Chemical absorption	85%	58%	33%	990	3.6	IEA GHG (2007a)
Furnaces	Refinery gas	Various refinery furnaces	17-100	Oxyfuel (stand-alone ASU)	90%	68-77%	6-23%	620-1150	1.8-3.1	Wilkinson et al. (2003)
			78-110	Oxyfuel (stand-alone ASU)	90%	22-25%	30-40%	1200-1580	N.D.	Zanganeh et al. (2004)

<sup>1)</sup> The values are presented for cases 1A and 2A. Both cases are operated at 60% of the maximum fuel input, and case 1A is operated at 25% heat efficiency and 2A at 40% efficiency. The high energy penalty ( $\Delta\eta$ ) value was calculated for case 2A and the low  $\Delta\eta$  value was calculated for case 1A.

Capital costs are expressed in €<sub>2007</sub> per kW rated electrical capacity in maximum power generation (condensing operation) mode.

<sup>2)</sup> 60% fuel input rate with 25% heat efficiency for a reference plant.

### *Industrial CHP plants*

To date, few feasibility studies have been published on CO<sub>2</sub> capture from industrial CHP plants. For NGCC-CHP plants, our previous study (Kuramochi et al., 2010 (Chapter 4))<sup>32</sup> shows that the energy penalty for plants in partial load operation (about 60%) may become 13-15% lower than that for condensing plants when the CHP plant is operating at a low heat efficiency of 25%. This is mainly due to a better use of the capacity of the NGCC-CHP plants. By increasing the fuel input rate to meet energy requirements for CO<sub>2</sub> capture, the plant efficiency increases and partly offsets the CO<sub>2</sub> capture energy requirements.

In such cases, in the ST/MT CO<sub>2</sub> avoidance costs for NGCC-CHP plants at scales as small as 100 MW<sub>e</sub> (condensing plant scale) may become lower than those for 400 MW<sub>e</sub> condensing plants. However, energy penalty becomes 40% higher than that for power-only NGCC plants when the CHP plant is operating at a higher heat efficiency of 40%, because supplementary firing will be required to meet heat demand for CO<sub>2</sub> capture.

Chemical absorption CO<sub>2</sub> capture from simple gas turbine (GT) CHP plants is considered to be uneconomical. Most GT-CHP plants are smaller than 40 MW<sub>e</sub> (Table 6-5), which is indicated to be expensive for NGCC-CHP plants even when making the best of unused capacity (Kuramochi et al., 2010 (Chapter 4)). Moreover, heat extraction from GT-CHP plants results in larger exergy losses compared to that from NGCC-CHP plants. Therefore, this option is not investigated further in this study.

#### **6.4.1.2 Oxyfuel combustion CO<sub>2</sub> capture**

Studies on the application of oxyfuel combustion CO<sub>2</sub> capture to smaller scale emission sources are scarce. There are some pilot oxyfuel coal boiler plants up to 30 MW<sub>th</sub> scale (e.g., Schwarze Pumpe in Germany) being tested to date (IEA, 2008a). Reduction of the CO<sub>2</sub> capture energy penalty can be achieved through heat integration in DH CHP plants. Low-grade heat can be recovered from the ASU and CO<sub>2</sub> compression unit. A case study on a German coal fluidized bed power plant (Simonsson et al., 2009) shows that more than 20% of power consumption for CO<sub>2</sub> capture can be recovered in the form of DH and capital cost can be reduced by 5% compared to condensing plants.

At natural gas-fired power plants, oxyfuel combustion capture is reported to be considerably more expensive than other CO<sub>2</sub> capture technologies (e.g., IEA GHG (2005), Kanniche et al. (Kanniche et al., 2010)) amongst others because special gas turbines have to be used for gas-fired oxyfuel power plants (IEA GHG, 2000b). Therefore, ST/MT oxyfuel combustion CO<sub>2</sub> capture for natural gas-fired installations is not investigated further in this study.

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<sup>32</sup> This study also made estimates for mid-term future (2020-2025) based on the future technological development estimated by Peeters et al. (2007). However, we in recent years consider the estimates to be on the optimistic side taking into account the developments in CCS R&D in recent years. Therefore, the mid-term future results are not considered here.

Relatively cheap CO<sub>2</sub> capture, however, may become possible for carbon-intensive gases such as refinery gases. Moreover, CO<sub>2</sub> emission sources are clustered in a refinery such that it would be more economical to capture CO<sub>2</sub> collectively (CCP, 2005; Farla et al., 1995). This is economically attractive particularly for oxyfuel combustion capture because the oxygen separation cost can be reduced significantly through upscaling of the ASU.

#### **6.4.1.3 Pre-combustion capture**

For coal-fired plants, integrated gasification combined cycle (IGCC) is a state-of-the-art power generation technology that could potentially have low-cost CO<sub>2</sub> capture in the longer term (Klara and Plunkett, 2010). However, pre-combustion CO<sub>2</sub> capture would not be the most economical option in the ST/MT, especially for distributed energy systems. Moreover, pre-combustion capture does not use any component that is of modular nature, so it is unlikely that it would have scale advantages over other CO<sub>2</sub> capture technologies at smaller plant scales. Also, IGCC technology is still in the early phase of market introduction and it is unlikely that small-medium scale IGCC power plants would be built for the purpose of district heating in the ST/MT.

For natural gas-fired plants, pre-combustion capture is considered to be significantly more expensive than other CO<sub>2</sub> capture processes for 500 MW<sub>e</sub> scale condensing plants (Kanniche et al., 2010). However, for smaller scale systems such as simple GT-CHP plants, pre-combustion may enable lower energy penalty compared to chemical absorption capture. The IEA GHG (IEA GHG, 2007a) proposes a 5 MW<sub>e</sub> system using autothermal reforming (ATR), water-gas shift reactor and pressure swing adsorption (PSA) for CO<sub>2</sub> capture.

Table 6-6 presents the potential energy and capital cost reduction as a result of heat integration for ST/MT CO<sub>2</sub> capture technologies in district heating and industrial applications. Our literature review indicated that the following smaller scale CO<sub>2</sub> capture options may be economical when fitted to DH CHP plants: (1) chemical absorption for NGCC-CHP, and (2) chemical absorption and (3) oxyfuel combustion for PC-CHP plants.

#### **6.4.2 Performance of ST/MT CO<sub>2</sub> capture technologies for DH CHP plants**

Chemical absorption and oxyfuel combustion were considered for coal-fired CHP plants and chemical absorption was considered for NGCC-CHP plants. As noted in Section 6.3.1, full-scale coal-fired power plants with and without CO<sub>2</sub> capture were scaled down to one-tenth calculate the performance for smaller scale DH CHP plants.

**Table 6-6:** Potential energy and capital cost reduction as a result of heat integration for ST/MT CO<sub>2</sub> capture technologies in district heating and industrial applications.

<i>Fuel</i>	<i>Application</i>	<i>Post-combustion (MEA)</i>	<i>Oxyfuel</i>
NG	DH	~30% solvent regeneration heat recovery	No information available. Possibly heat recovery from ASU and CO <sub>2</sub> compression, and capital cost reduction due to system integration as for coal DH CHP plants.
	Industrial	NGCC at partial load operation: ~15% energy penalty reduction	
Coal	DH	~30% solvent regeneration heat recovery	~25% of ASU and CO <sub>2</sub> compression power recovered as DH ~5% capital cost reduction for CHP
	Industrial	No particular benefits	No particular benefits

For PC-CHP plants, heat to power ratio (HPR) of 1 and 1.5 were considered. For NGCC-CHP plants, we assumed a low HPR of 0.5. It is unlikely that chemical absorption CO<sub>2</sub> capture will be fitted to NGCC-CHP plants operated at high HPR because of the competition for heat and the limit to the HPR that NGCC-CHP plants can reach. There are two options to meet both electricity and heat demands while capturing CO<sub>2</sub> at high capture efficiency (85-90%): (1) use auxiliary boiler for additional heat, or (2) build a larger CHP plant with increased electricity generation capacity. However, both of these options are problematic. The first option is not recommended because of increase in CO<sub>2</sub> capture energy penalty since only low-grade heat is generated from a high quality fuel. The second option requires larger capital investments and the increased electricity generation can become problematic if the electricity market is unfavorable for exporting excess electricity to the grid. In addition, for DG systems there is a technical limit of the grid to accept the electrical capacity (Ackermann et al., 2001). Therefore, this option is also unrealistic.

#### 6.4.2.1 Technical and economic parameters

The technical and economic performance for DH-CHP plants with ST/MT CO<sub>2</sub> capture were calculated based on the performance data for full-scale condensing plants. To ensure that their performance data are comparable, the base performance data for coal-fired (NETL, 2008) and natural gas-fired plants (NETL, 2007) were taken from a set of reports prepared by the NETL published between 2007 and 2008. Coal-fired plants were assumed to be bituminous coal-fed supercritical plants<sup>33</sup>.

Regarding the calculation of technical and economic performance for smaller scale plants, Table 6-7 shows the effect of scale on the technical performance of power plant and CO<sub>2</sub>

<sup>33</sup> Although supercritical plants are not common for small scales today, we expect it will become by the time CCS becomes commercial.

capture components. The maximum total gross power and heat efficiency of CHP plants was limited at 95%. In case the CHP plant cannot meet CO<sub>2</sub> capture energy demand for the nominal CO<sub>2</sub> capture efficiency due to the limit in maximum gross power and heat efficiency or the minimum gross power efficiency, CO<sub>2</sub> capture rate was adjusted to the heat that can be generated.

**Table 6-7:** Effect of scale on the technical performance of power plant and CO<sub>2</sub> capture components

<i>Parameter</i>	<i>Assumption and references</i>
Coal fired power plant efficiency	Unaffected by plant scale. This assumption agrees with the plant efficiencies calculated by the IECM model (CEEM CMU, 2010).
NGCC efficiency	Varies directly with the 0.0679 <sup>th</sup> power of gross power capacity as reported in Kuramochi et al. (2010), which was derived from data provided in the Gas Turbine World Handbook 2007-2008 (GTW, 2007)
Chemical absorption capture specific energy requirement	Unaffected by plant scale. This assumption is consistent with the assumptions made in previous studies for NGCC plants of 6-62 MW <sub>e</sub> scale (Gelowitz et al., 1995) and coal-fired steam turbine plants of 100-2500 MW <sub>e</sub> scale (CEEM CMU, 2010).
ASU specific power consumption (GJ <sub>c</sub> /tO <sub>2</sub> )	Varies directly with the -0.0474 <sup>th</sup> power of plant scale (tO <sub>2</sub> /d). The value 0.0474 is estimated from a figure that depicts the correlation between specific power consumption and plant scale presented in Castle (2002).

Capital costs for the full-scale power plants, with and without CO<sub>2</sub> capture, were broken down into key components. A scaling factor for each component was applied to calculate total capital costs at smaller plant scales (Eq.(6.12)):

$$C_X = \sum_i \left\{ C_{Ref,i} * \left( \frac{S_{X,i}}{S_{Ref,i}} \right)^{SF_i} \right\} \quad (6.12)$$

where:

$C_X$ : Capital cost for plant scale  $X$  MW<sub>e</sub>,

$C_{Ref,i}$ : Capital cost for the component  $i$  for the reference (full-scale) plant,

$S_{X,i}$ : Capacity of component  $i$  for the plant scale  $X$ ,

$S_{Ref,i}$ : Capacity of component  $i$  for the reference plant scale  $X$ ,

$SF_i$ : scaling factor for component  $i$ .

Table 6-8 shows the capital cost scaling factors for various components of power plants and CO<sub>2</sub> capture processes investigated in this study.

**Table 6-8:** Capital cost scaling factors

<i>Equipment</i>	<i>Scaling based on:</i>	<i>Scaling factor</i>	<i>Reference and notes</i>
Flue gas desulphurization	Gross power capacity under condensing operation	0.72	Sargent & Lundy (2010)
Coal-fired power plant	Gross power capacity under condensing operation	0.77 <sup>1)</sup>	CEEM CMU (2010)
NGCC	Gross power capacity in condensing operation	Varies directly with $(732+1.1X)/(1+0.0052X)$ ( $X$ : gross power capacity)	Kuramochi et al. (2010: Chapter 4)
De-NO <sub>x</sub> (selective catalytic reduction)	Flue gas volume	0.7	IPPC (2006)
Chemical absorption CO <sub>2</sub> capture unit	CO <sub>2</sub> capture capacity	0.6	Rao et al. (2004)
Air separation unit	Oxygen separation capacity	0.85 <sup>2)</sup>	Hamelinck and Faaij (2002)
CO <sub>2</sub> compression and purification	Compression power	0.53	See Eq.(6.9)

<sup>1)</sup> For 100-2000 tO<sub>2</sub>/d.

<sup>2)</sup> The capital costs were calculated by the IECM model to be 1852 \$/kW<sub>e</sub> net for 96.4 MW<sub>e</sub> net plant and 1275 \$/kW<sub>e</sub> net for 482 MW<sub>e</sub> net plant.

#### 6.4.2.2 Results

Table 6-9 shows the performance results for coal-fired CHP plants with and without CO<sub>2</sub> capture for different plant scales (full-scale and one-tenth scale). The results indicate slightly better net CHP efficiency for chemical absorption capture than for oxyfuel combustion capture. The effect of increased specific power consumption for ASU on the net CHP efficiency is found to be limited for the plant scales investigated in this study. In comparison with the plants in condensing operation, the efficiency penalty for chemical absorption CO<sub>2</sub> capture decreases by 18% from 28.2% to 23.2%, while for oxyfuel combustion CO<sub>2</sub> capture the decrease is marginal (4%).

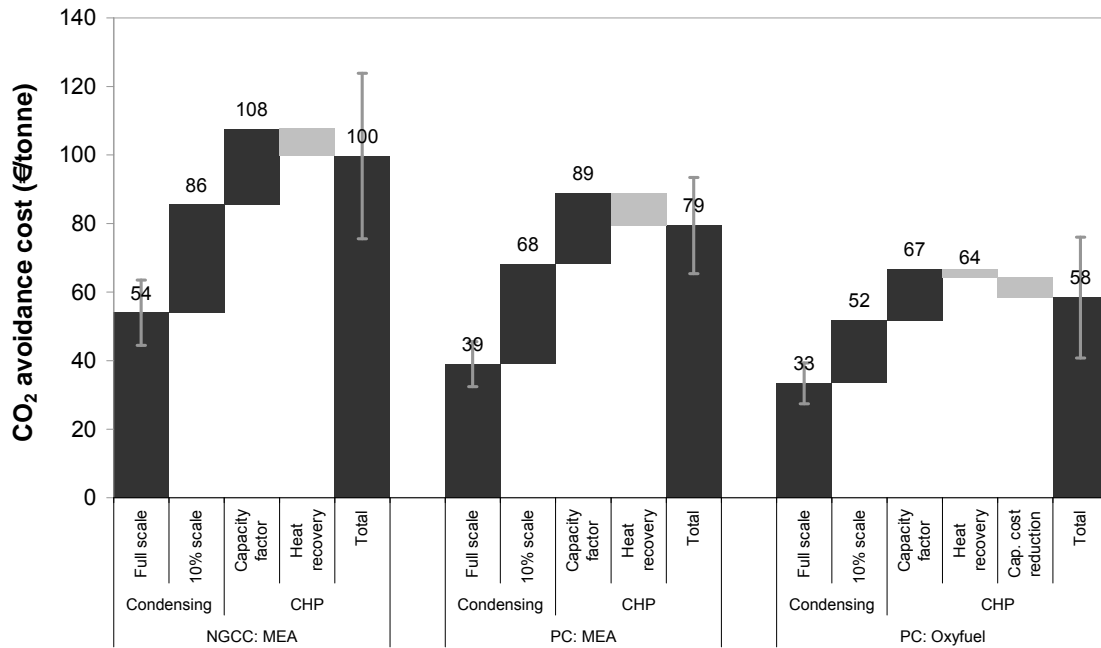
**Table 6-9:** Performance results for district heating (DH) coal-fired CHP plants with and without CO<sub>2</sub> capture in the ST. The performance data are also shown for full-scale condensing plants from NETL (NETL, 2007, 2008), on which the calculations were based. Cond.: condensing plant.

CO <sub>2</sub> capture technology Plant scale	No capture		Chemical absorption capture			Oxyfuel combustion capture			
	Full scale		Full scale		One-tenth scale	Full-scale		One-tenth scale	
	Cond.	1.5	Cond.	1.5	1.5	Cond.	1.5	Cond.	1.5
HPR									
Net electrical output (MW <sub>e</sub> )	550	470	549	470	47	549	470	53	47
Gross electrical efficiency	43.1%	37.2%	35.3%	32.5%	32.5%	43.4%	39.4%	43.4%	39.6%
System power consumption									
Power plant	-2.2%	-2.2%	-2.5%	-2.2%	-2.2%	-2.1%	-2.1%	-2.1%	-2.1%
CO <sub>2</sub> capture plant	-	-	-1.1%	-1.1%	-1.1%	-	-	-	-
Air separation unit	-	-	-	-	-	-7.0%	-7.0%	-7.8%	-7.8%
CO <sub>2</sub> compression and purification	-	-	-2.4%	-2.4%	-2.4%	-4.0%	-4.0%	-4.0%	-4.0%
Net electrical efficiency	40.8%	34.9%	29.3%	26.8%	26.8%	30.4%	26.4%	29.6%	25.7%
DH heat efficiency	-	34.9%	-	26.8%	26.8%	-	26.4%	-	25.7%
CO <sub>2</sub> capture heat recovered as DH heat	-	-	-	10.7%	10.7%	-	2.8%	-	3.0%
Gross energy conversion efficiency	43.1%	72.1%	64.6%	84.2%	84.2%	43.4%	65.8%	43.4%	65.3%
Net total CHP efficiency	-	69.8%	-	53.6%	53.6%	-	52.7%	-	51.4%
Energy penalty ( $\Delta\eta_{pen}$ )	-	-	28.2%	34.4%	23.2%	25.6%	24.5%	27.6%	26.4%
Capital cost (€ <sub>2007</sub> /kW <sub>e</sub> net)	1270	1490	2300	2560	5020	2160	2370	3810	4210
O&M cost (%-capital cost)	3.1%	2.6%	2.9%	2.3%	1.6%	2.6%	2.2%	1.9%	1.6%



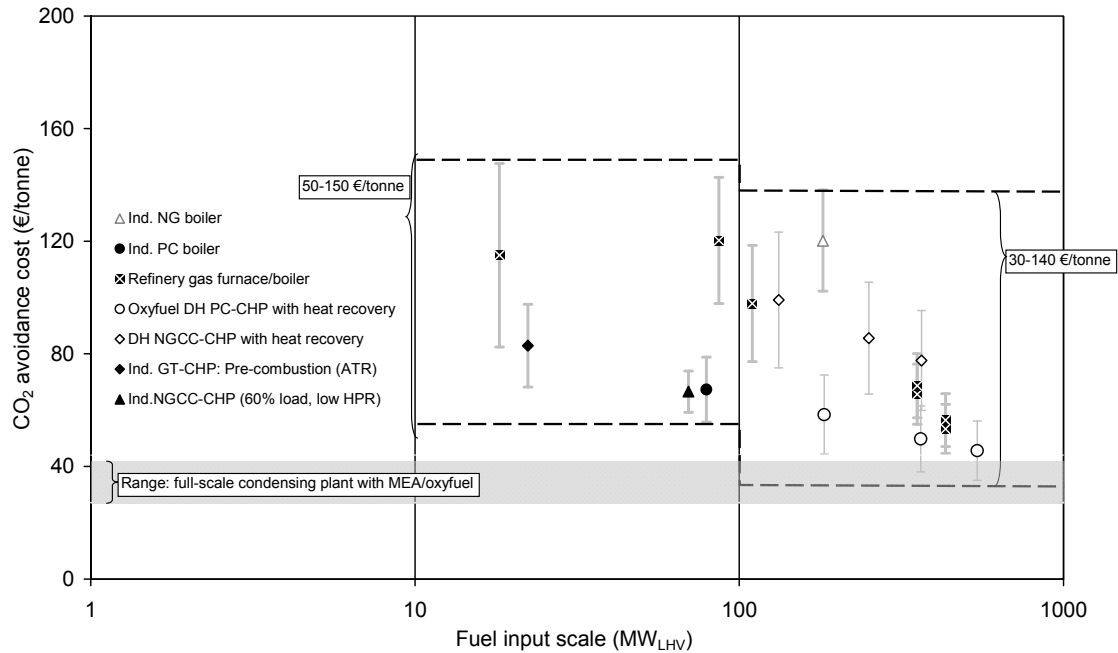
**Table 6-10:** Technical performance results for district heating (DH) NGCC-CHP plants with and without CO<sub>2</sub> capture for different plant scales.  
Cond.: condensing plant.

CO <sub>2</sub> capture technology	No capture			Chemical absorption capture		
	Full scale		One-tenth scale	Full scale		One-tenth scale
	Cond.	0.5	0.5	Cond.	0.5	0.5
Plant scale						
HPR						
Net electrical output (MW <sub>e</sub> )	560	516	52	482	516	52
Gross electrical efficiency	57.4%	53.0%	45.3%	52.4%	50.1%	42.4%
System power consumption: Total	-1.0%	-1.0%	-0.8%	-3.9%	-3.5%	-3.4%
Power plant	-1.0%	-1.0%	-0.8%	-1.4%	-1.0%	-0.9%
CO <sub>2</sub> capture plant	-	-	-	-1.0%	-1.0%	-1.0%
Air separation unit	-	-	-	-	-	-
CO <sub>2</sub> compression and purification	-	-	-	-1.5%	-1.5%	-1.5%
Net electrical efficiency	56.4%	52.0%	44.5%	48.5%	46.6%	39.0%
DH heat efficiency	-	26.0%	22.2%	-	23.3%	19.5%
Electrical efficiency reduction due to CO <sub>2</sub> capture energy extraction	-	-	-	-5.0%	-5.0%	-5.0%
CO <sub>2</sub> capture heat recovered as DH heat	-	-	-	-	22.9%	22.9%
Gross total heat and power generation efficiency	57.4%	79.0%	67.6%	-	73.3%	61.9%
Net total CHP efficiency	-	78.0%	66.7%	-	69.9%	58.6%
Energy penalty ( $\Delta\eta_{pen}$ )	-	-	-	14.0%	10.5%	12.2%
Capital cost (€ <sub>2007</sub> /kW <sub>e</sub> net)	410	440	790	860	980	1250
O&M cost (%- capital cost)	3.0%	2.0%	1.6%	2.6%	2.0%	1.6%



**Figure 6-3:** Difference in CO<sub>2</sub> avoidance costs between full scale condensing plants (PC: 550 MW<sub>e</sub>, NGCC: 560 MW<sub>e</sub>) and 10% scale CHP plants caused by plant scale, capacity factor, heat recovery, and capital cost reduction. The error bars represent sensitivity of the results due to varied energy prices, interest rate, economic plant lifetime and capital costs.

Figure 6-3 shows CO<sub>2</sub> avoidance costs for DH-CHP plants with chemical absorption and oxyfuel CO<sub>2</sub> capture for varied plant scales. Costs are broken down into components to show the effect of factors contributing to the difference in CO<sub>2</sub> avoidance costs between full scale condensing plants and one-tenth scale CHP plants. For NGCC, the performance results without heat recovery are also presented as a reference. For coal-fired CHP plants, CO<sub>2</sub> avoidance costs are similar for the two capture technologies at full-scale but oxyfuel capture becomes significantly cheaper than chemical absorption capture at smaller scales because of lower capital costs. For NGCC-CHP plants operating at low HPR (=0.5), the figure shows that the cost reduction due to heat recovery is less than 10 €/tCO<sub>2</sub> avoided. The figure also shows that the lower capacity factor of CHP plants compared to centralized power plants increases the cost by around 20 €/tCO<sub>2</sub> avoided.



**Figure 6-4:** CO<sub>2</sub> avoidance costs reported in the literature (after normalization) or calculated in this study as a function of fuel input scale (logarithmic scale) for short-mid term technologies.

### 6.4.3 Section summary

#### 6.4.3.1 Energy penalty due to CO<sub>2</sub> capture in the ST

Based on the literature data presented in Table 6-5, the energy penalty due to CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ ) was found to vary significantly by fuel type, energy system type and its application. For natural gas-fired systems,  $\Delta\eta_{pen}$  was found to range between 28% for boilers down to 11% for NGCC-CHP plants operated in low HPR by having a good heat integration of the system.  $\Delta\eta_{pen}$  was found to be larger for smaller plants because the efficiency reduction is similar for large and small scale plants in percentage point terms while the reference system efficiency of small scale plants is lower.

For coal-fired plants,  $\Delta\eta_{pen}$  of 33% was observed for coal boilers without steam turbine and around 23-28% for PC power plants. The effect of scale on  $\Delta\eta_{pen}$  is smaller than with natural gas-fired plants because the power plant electrical efficiency does not reduce as much as in gas-fired plants.

#### 6.4.3.2 CO<sub>2</sub> avoidance cost

Figure 6-4 shows CO<sub>2</sub> avoidance costs reported in the literature presented in Table 6-5 (after normalization) and calculated in this study as a function of fuel input scale<sup>34</sup>. The figure also shows the cost range for full-scale condensing power plants (29-45 €/tonne). It can be concluded that in the ST/MT, CO<sub>2</sub> avoidance costs range about 30-140 €/tonne for

<sup>34</sup> The cost calculation for Kuramochi et al. (2010, Chapter 2 of the thesis) were performed for 500 h/yr capacity factor because the technical performance data is only valid for partial load operations.

plant scales larger than 100 MW<sub>LHV</sub> (fuel input) and 50-150 €/tonne for 10-100 MW<sub>LHV</sub>. The CO<sub>2</sub> avoidance costs for DH-CHP plants calculated in this study are in line with the range of costs calculated for technologies presented in Table 6-5. From the results it is concluded that the more promising options are: (1) industrial NGCC-CHP plants with chemical absorption capture in low HPR and low load operations, (2) oxyfuel combustion capture from PC-CHP plants, and (3) some refinery gas-fired plants with oxyfuel combustion capture.

Figure 6-4 indicates a clear effect of scale on CO<sub>2</sub> avoidance costs, with the exception of oxyfuel combustion CO<sub>2</sub> capture for refinery gases. The differences observed are caused by two key reasons. Firstly, the degree of furnace/boiler efficiency improvement by converting to oxyfuel combustion differs by furnace/boiler type. Secondly, the amount of oxygen required depends on the composition of the refinery gas. The uncertainty on costs becomes significantly larger for smaller scale CO<sub>2</sub> capture than for large-scale capture mainly because of the higher specific capital costs. However, small scale CO<sub>2</sub> capture may still become competitive to large-scale capture under specific circumstances, e.g., low interest rate combined with low energy prices.

## 6.5 Assessment of long-term CO<sub>2</sub> capture technologies

As noted in the introduction, this section emphasizes the assessment of the technology itself rather than the combination of CO<sub>2</sub> capture technologies and energy system types. Table 6-11 shows the technical and economic performance data for small-medium scale installations with CO<sub>2</sub> capture technologies that may become available in the LT. In many of the proposed systems, CO<sub>2</sub> capture process is entirely or partially comprised of membranes. As it was done in the previous section, we also assessed the applicability of large scale technologies for smaller scale emission sources.

### 6.5.1 Post-combustion capture

Most post-combustion CO<sub>2</sub> capture using membranes focus on coal combustion because of the high CO<sub>2</sub> concentration in the flue gas. Membrane separation processes are driven by the CO<sub>2</sub> partial pressure difference. A brief description of different types of membranes under investigation follows.

#### 6.5.1.1 Ceramic membranes

The molten carbonate fuel cell (MCFC) may become not only a future power generation technology but also a cost-effective CO<sub>2</sub> concentrator. In an MCFC stack, CO<sub>2</sub> is used as cathode fuel to generate electricity and it transports to anode side of the cell in the form of carbonate ions (CO<sub>3</sub><sup>2-</sup>). As a result, anode off-gas has a high concentration of CO<sub>2</sub>. A large fraction of this gas is recycled back to the cathode inlet to supply CO<sub>2</sub> in conventional CFC systems<sup>35</sup>. However, the CO<sub>2</sub>-rich anode off-gas can also be sent to a purification and

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<sup>35</sup> The MCFC plant CO<sub>2</sub> production is ultimately contained in the cathode outlet flow where CO<sub>2</sub>-rich gas is diluted by air (Campanari, 2002).

compression process if there is an external source of CO<sub>2</sub> that can be fed to the cathode. Several studies modeled and experimented systems to capture CO<sub>2</sub> from conventional combustion plant flue gas using MCFC (Amorelli et al., 2004; Campanari, 2002; Lusardi et al., 2004; Sugiura et al., 2003; Toyota et al., 2008). A pilot plant demonstrated a stable operation of CO<sub>2</sub> separation from the flue gas of a coal fired power plant with a SO<sub>2</sub> concentration of 20-70 ppm<sup>36</sup> for 2000 hours with little fuel cell voltage degradation, indicating a possible continuous operation of 40,000-50,000 hours (Toyota et al., 2008). Although a number of concerns<sup>37</sup> over the practicality of CO<sub>2</sub> capture from flue gas using MCFC has been expressed (Pennline et al., 2009), this technology could become commercial in the long term.

#### **6.5.1.2 Polymer-based membranes**

CO<sub>2</sub> capture using commercially available polymeric membranes have resulted in higher energy penalties compared to a chemical absorption process (IPCC, 2005), but significant performance improvement is expected in the future. Polymeric membrane CO<sub>2</sub> separation is driven by the partial CO<sub>2</sub> pressure difference between the feed and permeate gases by either using a compressor on the feed side or a vacuum-pump on the permeate side.

Polymeric membrane separation may become attractive for smaller scale emission sources although there is no feasibility study for smaller scale applications available. Literature suggests that the use of vacuum pumps reduces energy consumption considerably<sup>38</sup> (Favre, 2007; Ho et al., 2008; Zhao et al., 2008) and possibly lower total capital costs (Ho et al., 2008). The fraction of membranes in the total capital costs (39-47%) is considerably larger (5%) for vacuum pump-driven systems than for compressor-driven systems (see Figure 6-5). This is favorable for smaller scale CO<sub>2</sub> capture because the capital cost increase as a result of scaling down may be limited.

#### **6.5.1.3 Facilitated transport membranes**

Matsumiya et al. (2004) report that facilitated transport membranes show better energy and economic performance than polymer membranes. The essential element of facilitated transport membranes is the specific chemical interaction of a gas component with a compound present in the membrane, the so-called carrier (IPCC, 2005). Carbonates, amines and molten salt hydrates have been suggested as carriers for CO<sub>2</sub> separation. The separation process is driven by the difference in partial pressure of the component to be transported (IPCC, 2005).

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<sup>36</sup> These values are much higher than the limit for chemical absorption capture using MEA, which is reported to be 10 ppm (Chapel et al., 1999).

<sup>37</sup> In the literature, the following concerns have been named: (1) molten carbonate is a corrosive material operating at high temperatures, making it difficult to handle; (2) high temperatures necessary for cell operation are extremely difficult for integration into a post-combustion application; (3) only small voltage is allowable to avoid decomposition of the molten carbonate, leading to a huge stack requirement, (4) sensitivity to the contaminants such as SO<sub>2</sub>.

<sup>38</sup> The required membrane area becomes much larger for vacuum pump-driven systems due to the lower pressure ratio, but the consequent cost increase could be offset by the omission of expensive feed gas compressor.

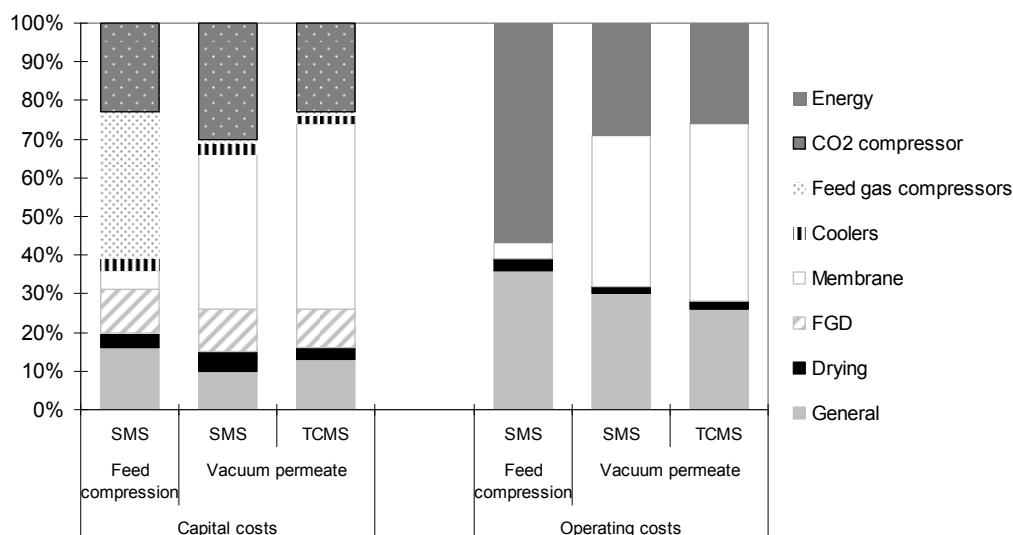
**Table 6-11:** Energy, economic and CO<sub>2</sub> performance data for small-medium scale installations with long term CO<sub>2</sub> capture technologies reported in the literature

Energy system type	Scale	Fuel type	Energy plant type	Plant scale [MW]	CO <sub>2</sub> capture technology	CO <sub>2</sub> capture efficiency	Conversion efficiency ( $\eta$ )	Energy penalty ( $\Delta\eta_{pen}$ )	Capital cost [€/kW]	O&M cost [%-capital cost]	Reference
Condensing/CHP plants	Medium (5-50 MW <sub>e</sub> )	Natural gas	SOFC-CHP (industrial)	5	Oxyfuel (stand-alone ASU)	94% <sup>1)</sup>	53%	5%	2100 <sup>2)</sup>	6	Kuramochi et al. (2011: Chapter 5)
			SOFC-GT-ST	20	Oxyfuel (integrated OCM)	100%	59%	- <sup>3)</sup>	2400	3	Lokurlu et al. (2004)
			AZEP (condensing)	50	Oxyfuel (membrane-based)	85-100%	49-50%	5-9% <sup>3)</sup>	1900-2100	4	Möller et al. (2006)
	Small (0.005- <5 MW <sub>e</sub> )	Natural gas	SOFC-CHP (DH)	0.5	Oxyfuel (integrated OCM)	100%	52%	5% <sup>3)</sup>	2500	4	IEA GHG (2007a)
			SOFC-GT		Oxyfuel (stand-alone ASU)	100%	58%	4% <sup>4)</sup>	1600	4	Franzoni et al. (2008)
Boilers	Medium (5-50 MW <sub>e</sub> )	Natural gas	Boiler (industrial and DH)	170	Oxyfuel (OCM)	92%	93%	4%	130-360	5	Switzer et al. (2005) IEA GHG (2007a)
		Coal	Boiler (industrial)	50	Oxyfuel (OCM)	86-98%	81%	6-8 %	560-680	4	IEA GHG (2007a)
H <sub>2</sub> plants	Small (0.005-5 MW <sub>H2</sub> )	Natural gas	WGSMR (refueling, DH)	2	Pre-combustion	57%	69%	4%	770 <sup>2)</sup>	15	Sjardin et al. (2006)
				1	Pre-combustion	50%	67%	9%	2800	N.D.	Kurokawa et al. (2008)
Others	Various	Various	MCFC using external CO <sub>2</sub> as cathode fuel	4	MCFC (electro-chemical)	100-200%	26-31%	21-36%	2700-2900	16	Own estimate based on Frangini (2010) <sup>5)</sup>

<sup>1)</sup> The reported CO<sub>2</sub> purity is 84%. Therefore, the CO<sub>2</sub> capture efficiency was recalculated as described in section 6.3.3.

<sup>2)</sup> Compressor costs are recalculated using Eq.(6.9). The fuel cell stack cost was assumed to be 500 \$<sub>2008</sub>/kW for flat tubular cell geometry. Note that this is an ambitious cost target for tubular SOFC (e.g., see Kuramochi et al. (2011: Chapter 5)).

- <sup>3)</sup> The reference literature compares the technical performance with a conventional NGCC of identical scale with an electrical conversion efficiency of 40% (Lokurlu et al., 2004). We considered the comparison to be unfair because the state-of-the-art NGCC of similar scale (L20A-1, 25 MW<sub>e</sub> net) already shows 48.6% efficiency (LHV) (GTW, 2007).
- <sup>4)</sup> The referenced study does not discuss the lifetime of fuel cell stacks and the costs for cell stack replacement. Therefore, it is not clear how these costs are taken into account. The referenced study also does not provide separate cost figures for variable O&M costs (chemicals, water, etc.). These costs, however, are reported to be of limited influence on total costs (Rienschel et al., 1998b).
- <sup>5)</sup> Frangini (2010) estimates that the electrical efficiency of MCFC decreases by 10% when using coal combustion flue gas and by 30% when using natural gas combustion flue gas compared to the conventional operation of recycling the anode off-gas as cathode fuel. While the Frangini study uses the current cost for the MCFC system, we assumed that the capital cost decreases to one-tenth of the current level in the LT. This is equivalent to around 1200 €/kW<sub>e</sub> without CO<sub>2</sub> capture, which is roughly in line with the predictions presented in Araki and Keppo (2005). The electrical efficiency is assumed unchanged in the LT.



**Figure 6-5:** Breakdown of capital costs and operating costs for CO<sub>2</sub> separating polymeric membranes operating under vacuum permeate and feed compression conditions. SMS: single stage membrane system, TCMS: two stage cascade membrane system. Source: Adapted from Ho et al. (2008).

#### 6.5.1.4 Membrane-gas hybrid processes

Membrane gas absorption (MGA) combines the advantages on absorption (high selectivity) and membranes (modularity and small size) (Feron and Jansen, 2002). A porous, water repellent, polymeric membrane separates the gas phase (feed gas) from the liquid phase (absorbent). CO<sub>2</sub> in the flue gas diffuses through the membrane and gets absorbed into the liquid absorbent. The CO<sub>2</sub>-rich absorbent is regenerated by heating it in a stripper, just as in a conventional chemical absorption process. It is suggested that the MGA process may enable lower CO<sub>2</sub> capture cost compared to conventional chemical absorption process at smaller plant scales due to its modular nature (IEA GHG, 2007a). Other literature, however, suggests otherwise. Capital costs are dominated by components other than the membrane module such as the CO<sub>2</sub> stripper tower (Grønvold et al., 2005; Yan et al., 2008)<sup>39</sup>.

The membrane flash process investigated by RITE (e.g. Matsumiya et al., 2004; Okabe et al., 2010) may become suitable for small-scale applications. In this process, the feed gas and carrier solution are both supplied to the high-pressure feed side of the hollow fiber membrane module. Driven by the vacuum pump at the permeate side, the CO<sub>2</sub>-rich solution permeates the membrane and liberates the dissolved CO<sub>2</sub> at the permeate side of the membrane. The system proposed by Matsumiya et al. (2004) uses neither absorber nor desorber towers. Therefore, the membrane module is reported to account for about 60% of

<sup>39</sup> Grønvold et al. (2005) show that the fraction of membrane contactor in total capital cost (excluding FGD and including CO<sub>2</sub> compression unit) is only 14%. The cost estimates from Yan et al. (2008) show that also for a smaller plant scale of 2.9 kgCO<sub>2</sub>/s captured, the fraction of membrane contactor in total capital cost (excluding both FGD and CO<sub>2</sub> compression units) is less than 30%.



total capital costs (excluding FGD, including CO<sub>2</sub> purification and liquefaction) (Matsumiya et al., 2004).

### **6.5.2 Oxyfuel combustion capture**

Oxygen conducting membranes (OCM) is a promising advanced air separation technology. OCM is a ceramic membrane made of perovskites, which have both electronic and oxygen ionic conductivity when operated at high temperatures around 700 °C to 900 °C (Anheden et al., 2005b). Oxygen ions are transferred under a gradient of oxygen partial pressure on the opposite side of the membrane. OCM can produce oxygen of very high purity, above 99% (Smith and Klosek, 2001). A high level of system integration can be achieved for gas-fired boilers. In a gas-fired boiler system proposed by Switzer et al. (2005), the gradient of oxygen partial pressure is generated by placing a fuel combustor at the permeate side of the membrane.

Another oxyfuel combustion-based technology that may become economically competitive for smaller scale gas-fired power generation is the advanced zero emission power (AZEP) concept. A general description of the AZEP technology can be found in, e.g., (Damen et al., 2006; Möller et al., 2006; Sundqvist et al., 2001). The key component is a mixed conducting membrane (MCM) reactor. Compressed air flows through the MCM reactor, where oxygen is separated and directed to the combustion chamber. The combustion heat is transferred to O<sub>2</sub>-depleted compressed air via the MCM reactor. Electricity is generated from an expansion of the hot O<sub>2</sub>-depleted compressed air in a gas turbine and an expansion of steam generated from the gas turbine exhaust gas, just like in a conventional NGCC plant. A modeling study indicates that the AZEP plant may perform better at smaller scales (about 50 MW<sub>e</sub>) than at large scales (around 400 MW<sub>e</sub>) because of the smaller difference in combustor outlet temperature between the reference NGCC and AZEP plant. Although there are still technical challenges in the MCM development (Damen et al., 2006), it may become an attractive option for medium-size DG.

### **6.5.3 Pre-combustion CO<sub>2</sub> capture: Hydrogen plants**

Small-scale H<sub>2</sub> production plants may serve as refueling stations for future hydrogen vehicles as well as for district and residential fuel cell systems. Water gas shift membrane reactor (WGSMR) technology suits well for these applications because of its modular nature. In addition, WGSMR technology is expected to enable cheap CO<sub>2</sub> capture (Sjardin et al., 2006). Tokyo Gas has been running 120 kW scale demonstration units since 2004 (Kurokawa et al., 2008). CO<sub>2</sub> avoidance cost estimates for 2020 are reported to be well above 100 €/tonne (Kurokawa et al., 2008) and long-term durability of the membranes still remains a challenge (Shirasaki et al., 2009). Nevertheless, the cost estimates reported in Sjardin et al. (2006) shown in Table 6-11 should be achievable in the longer term.

### **6.5.4 Fuel cell systems: Solid oxide fuel cells (SOFC)**

Solid oxide fuel cell (SOFC) consists of electrolytes made of a solid, non-porous metal oxide (Y<sub>2</sub>O<sub>3</sub>-stabilized ZrO<sub>2</sub>), and it operates at 600-1000 °C where ionic conduction by

oxygen ions take place (EG&G Technical Services Inc., 2004). SOFC is seen as a promising future energy conversion technology, particularly as DG systems, for mainly two reasons (Wee, 2010). Firstly, SOFC enables high electrical conversion efficiency for both large and small scale systems. In particular, SOFC hybrid systems, which integrate SOFC with GT or steam turbine, may achieve electrical conversion efficiency as high as 70% (an overview of literature can be found in, e.g. (Wee, 2010)).

Secondly, SOFC has a potential to enable cheap CO<sub>2</sub> capture. In a SOFC system, oxygen is transported through the electrolyte from the cathode side to the anode side, where the fuel is oxidized. Therefore, the generated CO<sub>2</sub> contained in the anode off-gas is not diluted by nitrogen and is potentially easy to capture. An overview of applicable capture technologies can be found in, e.g., Kuramochi et al. (2011: Chapter 5)<sup>40</sup> and Damen et al. (2006). A review study by Wee (2010) suggests that the SOFC hybrid systems of 1-100 MW<sub>e</sub> scale may achieve electrical conversion efficiency higher than 60% even when equipped with CO<sub>2</sub> capture. Although SOFC is currently considered to be too expensive for commercialization, the potential for cheap CO<sub>2</sub> capture from advanced fuel cell technologies such as SOFC may facilitate the market introduction of these fuel cells in a carbon-constrained society (Kuramochi et al., 2010(Chapter 4)). To date, there are only a few studies that explored the techno-economic performance of CO<sub>2</sub> capture from SOFC systems (Franzoni et al., 2008; IEA GHG, 2007a; Kuramochi et al., 2011; Lokurlu et al., 2004).

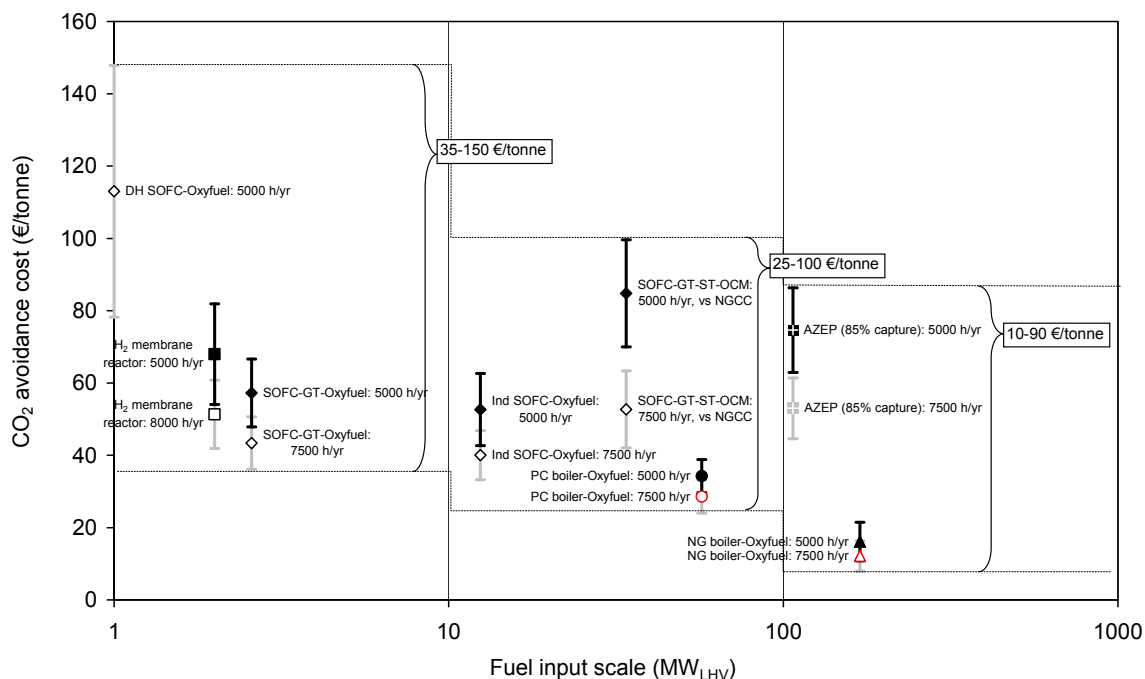
### 6.5.5 Results: LT

As seen in Table 6-11, for advanced energy conversion technologies, the energy penalty due to CO<sub>2</sub> capture compared with the identical plant without CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ ) reported in the literature ranged around 4% and 9% for CO<sub>2</sub> capture efficiency of 85-100%. These values are significantly smaller than those reported for ST/MT technologies because the advanced systems are designed to capture CO<sub>2</sub>.

Figure 6-6 shows CO<sub>2</sub> avoidance costs for distributed energy systems based on advanced technologies as a function of fuel input scale. Note that the performance of SOFC-GT-ST hybrid system (Lokurlu et al., 2004) was compared with that of the reference NGCC plant (50 MW<sub>e</sub>) used for the AZEP power technology because the performance of an identical SOFC-GT-ST plant without CO<sub>2</sub> capture is not reported in the referenced study (Lokurlu et al., 2004). The calculations were performed for different capacity factors: 7500 h/yr and 5000 h/yr. Note that these cost estimates assume that the technologies are mature. CO<sub>2</sub> avoidance costs range around 10-90 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 25-100 €/tonne for 10-100 MW<sub>LHV</sub> and 35-150 €/tonne for 10 MW<sub>LHV</sub> or smaller.

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<sup>40</sup> The referenced study defines the timeframe “up to 2025”. However, taking into account the recent developments for commercialization of SOFC, it is more likely that CO<sub>2</sub>-capture SOFC systems will be introduced to the market after 2030.

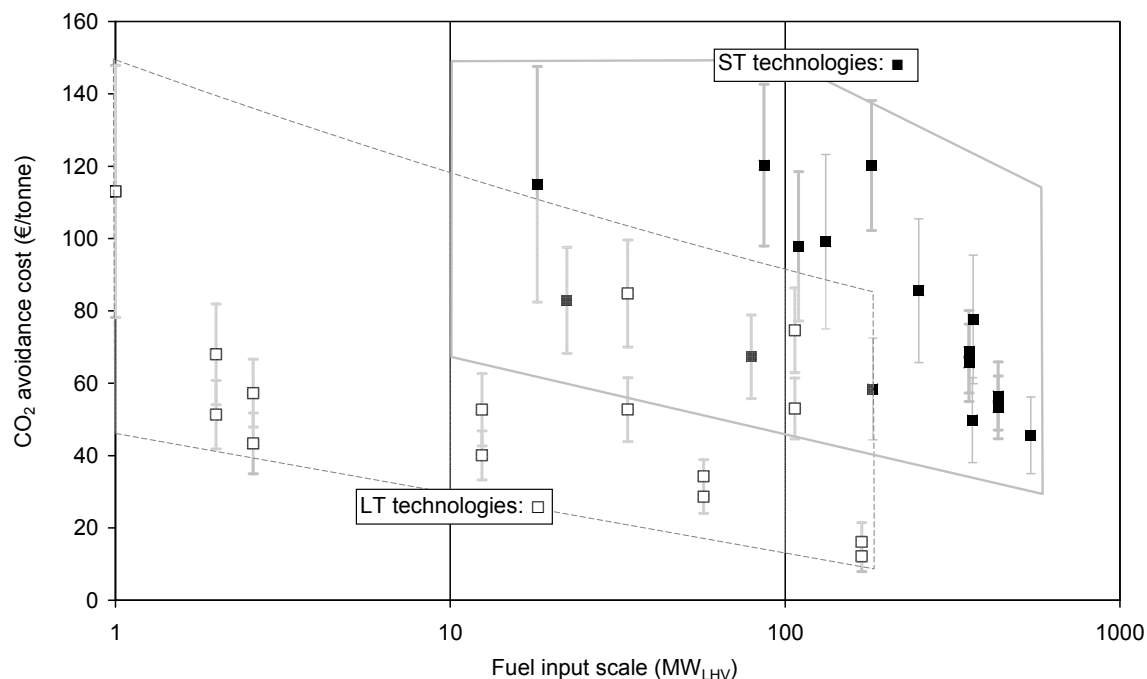


**Figure 6-6:** CO<sub>2</sub> avoidance costs reported in the literature as a function of fuel input scale (logarithmic scale) for long term technologies. It is shown that the reduced load leads to considerably higher costs.

The effect of capacity factor is significant (10-30 €/tonne increase in CO<sub>2</sub> avoidance cost) for some capital-intensive technologies that competes economically with conventional technologies by demonstrating a considerably higher energy efficiency, e.g., SOFC systems and H<sub>2</sub> membrane reactors. For boilers, the effect is limited because of limited additional capital costs. The effect of scale is not clear; there are several reasons to explain this. Firstly, some technologies reported for smaller scale systems may be more advanced than those reported for larger scale systems. Secondly, there is more uncertainty in the performance data of LT technologies because they are studied in less detail compared to ST/MT CO<sub>2</sub> capture technologies, for which case-specific feasibility studies are available.

## 6.6 Combined results

Figure 6-7 combines the short term and long term CO<sub>2</sub> avoidance cost results presented in Figure 6-4 and Figure 6-6. The two colored bands roughly show the range of CO<sub>2</sub> avoidance costs for short term technologies and long-term technologies, respectively. In the ST/MT, CO<sub>2</sub> avoidance costs would range around 40-160 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 60-200 €/tonne for 10-100 MW<sub>LHV</sub> and 80-200 €/tonne or more for 10 MW<sub>LHV</sub> or smaller. For the LT, it is estimated that the CO<sub>2</sub> avoidance costs may range around 10-40 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 10-80 €/tonne for 10-100 MW<sub>LHV</sub> and 10-120 €/tonne for 10 MW<sub>LHV</sub> or smaller.



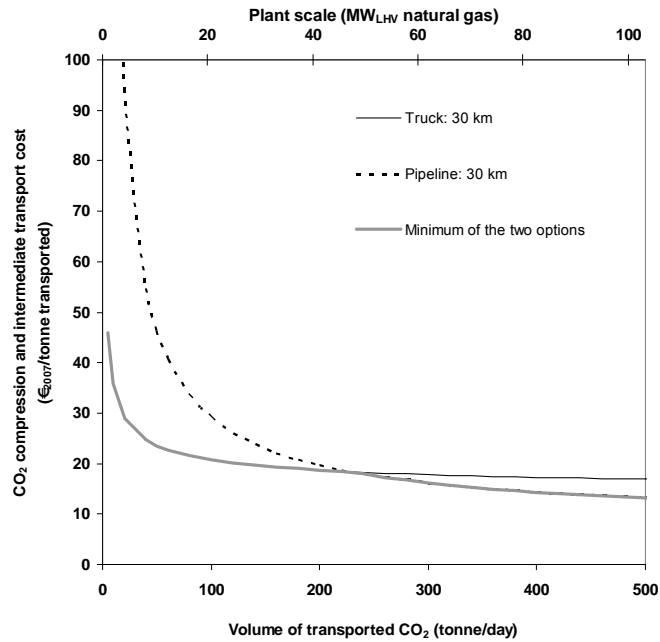
**Figure 6-7:** CO<sub>2</sub> avoidance costs reported as a function of fuel input scale (logarithmic scale) for short-term and long-term technologies. The clusters and its borders are only for illustrative purposes.

## 6.7 CO<sub>2</sub> compression and distributed transport costs

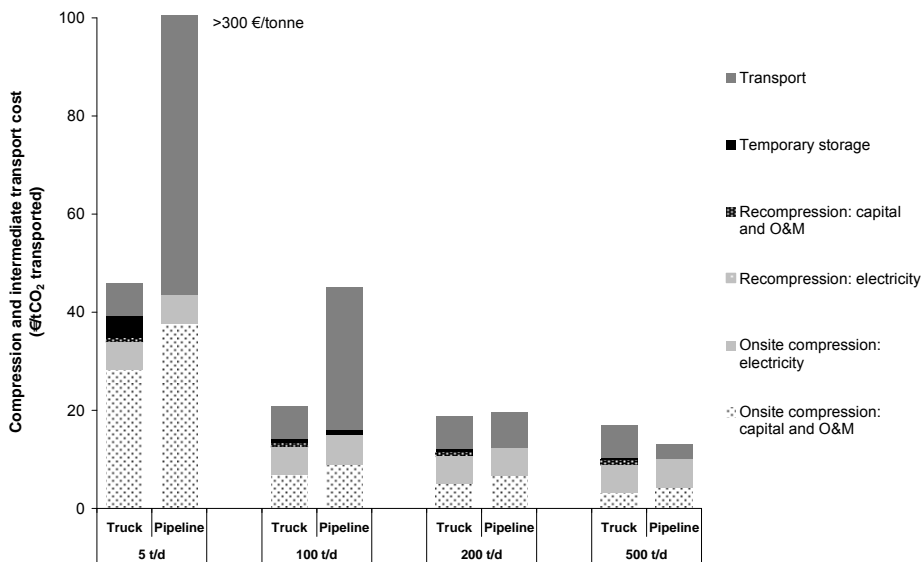
In this section, the costs of CO<sub>2</sub> compression and distributed transport between the CO<sub>2</sub> capture site and trunk pipeline using branch pipeline and truck are evaluated. The schemes of two transport options compared were presented in Figure 6-2 and the parameters used for the calculations were presented in Table 6-4. Note that the assumptions on the transport distance (30 km) and the lack of network for distributed CO<sub>2</sub> collection is conservative. Many distributed emission sources would be located within a distance far less than 30 km from the nearest trunk pipeline once a large-scale CO<sub>2</sub> transport infrastructure is developed.

Figure 6-8 shows the CO<sub>2</sub> compression and distributed transport costs calculated for the two options while Figure 6-9 shows the breakdown of the costs by component for different transport scales. Truck transport becomes cheaper than pipeline transport for emission sources smaller than around 200 tCO<sub>2</sub>/d (equivalent to combustion plants of 20-30 MW<sub>LHV</sub> natural gas input) because of lower investment costs. As shown in Figure 6-9, the costs are dominated by CO<sub>2</sub> compression (65% at the lowest) and the cost for transport itself is small (35% at the highest) for all scales. For larger scales when branch pipeline transport is the cheapest, the branch pipeline cost is about 5 €/tonne (see 500 t/d in Figure 6-9). For smaller scales, when truck transport is the cheapest, the cost is about 6 €/tonne and is unaffected by the scale (see 5 and 100 t/d in Figure 6-9). Figure 6-9 also shows that the CO<sub>2</sub> recompression cost for the truck transport option is negligible.

Our results are in line with those from IEA GHG study (IEA GHG, 2007b), which reports a marginal cost of 34 \$/tCO<sub>2</sub> transported (about 25 €<sub>2007</sub>/tCO<sub>2</sub>) for collecting 0.7 MtCO<sub>2</sub>/yr from 49 sources (average 14 ktCO<sub>2</sub>/yr per source).



**Figure 6-8:** Small scale CO<sub>2</sub> transport costs as a function of transported CO<sub>2</sub> volume per day.



**Figure 6-9:** Breakdown of CO<sub>2</sub> compression and distributed transport costs.

## 6.8 Discussion

The results obtained in this study and their main uncertainties and sensitivities have been discussed in their respective sections. In this section, we mainly discuss the limitations of this study and the implications of the obtained results in relation to the expected future CO<sub>2</sub> price.

A number of limitations of this study are identified. Firstly, the literature survey showed that there are only a few studies available on the techno-economic feasibility of CO<sub>2</sub> capture from distributed energy systems. For many energy conversion technologies with CO<sub>2</sub> capture, there was only one techno-economic study available. Therefore it was difficult to assess whether the reported values are reliable. Moreover, some of the reported economic performance data may be outdated. Furthermore, many studies do not have clear definition of “capital cost” because these studies are of an order-of-magnitude analysis nature.

Secondly, the results obtained in this study may be sensitive to the reference technologies selected for comparison. Since DG systems will always be compared with centralized generation systems, the economic competitiveness of CO<sub>2</sub> capture-fitted DG systems will likely depend largely on the local energy market circumstances of the CO<sub>2</sub> capture site. Moreover, advanced energy conversion systems investigated in this study are designed to capture CO<sub>2</sub> and it may not make economical sense to apply the technology without CO<sub>2</sub> capture. In such a case, the competitor of an advanced energy system with CO<sub>2</sub> capture would be conventional energy conversion systems and therefore, the CO<sub>2</sub> mitigation performance results may differ significantly from those reported in this study.

The third limitation relates to the nature of distributed energy systems. In this study, the uncertainty of CO<sub>2</sub> avoidance costs was found to be considerably larger than that for large-scale centralized power plants. This is logical because decentralized energy conversion systems show a much larger diversity in their operational conditions (e.g. load factor and heat-to-power ratio) compared to centralized power plants.

## 6.9 Conclusions

CO<sub>2</sub> emissions from distributed energy systems may become significant; accounting in 2030 for about 20% of the current global energy-related CO<sub>2</sub> emissions. This study assessed and compared the technical and economic performance of CO<sub>2</sub> capture technologies for distributed energy systems. We took into account the diversity in type of generator technologies used, applications, plant scales, operational patterns, and timeframe in which the technologies may become commercial. We also investigated the compression and distributed transport of captured CO<sub>2</sub> between the capture site and trunk pipeline. The study covered CHP plants, boilers and distributed hydrogen plants. The following conclusions are drawn:

In the short-mid term, the energy penalty due to CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ ) varies significantly by fuel type, energy system type and its application. For natural gas-fired installations,  $\Delta\eta_{pen}$  was found to range between 28% for boilers down to 11% for NGCC-CHP plants. For coal-fired installations,  $\Delta\eta_{pen}$  of 33% was found for coal boilers and around 23-28% for PC power plants. It can be concluded that in the short-term (around 2020-2025), CO<sub>2</sub> avoidance costs range between 30-140 €/tonne for plant scales larger than 100 MW<sub>LHV</sub> (fuel input) and 50-150 €/tonne for 10-100 MW<sub>LHV</sub>. The most promising smaller-scale options identified are: (1) industrial NGCC-CHP plants with chemical absorption capture in low HPR and low load operations, (2) oxyfuel combustion capture from PC-CHP plants, and (3) oxyfuel combustion capture from some refinery gas-fired plants.

We also investigated the heat integration potentials between a power plant and a CO<sub>2</sub> capture process. For chemical absorption capture, it was suggested that up to 30% of the solvent regeneration heat can be recovered in the form of hot water for DH without additional capital cost. For oxyfuel combustion capture from coal-fired plants, it was suggested that 30% of the energy consumption for ASU and CO<sub>2</sub> compression can be recovered and about 5% reduction of capital cost can be achieved by heat integration.

In the long-term future, a number of advanced CO<sub>2</sub> capture technologies using oxygen separation membranes have been found for both coal and natural gas-fired systems. Membrane-based oxyfuel combustion processes increases the furnace/boiler efficiency without significant electricity consumption. Other promising technologies include solid oxide fuel cells (SOFC) for power and CHP applications, and water-gas shift membrane reactor (WGSMR) for hydrogen production. For coal-fired power plants, there are also a number of advanced post-combustion capture technologies that may enable low-cost CO<sub>2</sub> capture at small scales, including modified molten carbonate fuel cells (MCFC) and polymer-based membranes.

*The energy penalty due to CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ ) was found to range around 4-9%. Regarding the economic performance, it was concluded that in the LT, CO<sub>2</sub> avoidance costs would be around 10-90 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 25-100 €/tonne for 10-100 MW<sub>LHV</sub> and 35-150 €/tonne for 10 MW<sub>LHV</sub> or smaller. The influence of the capacity factor is significant for some technologies that demonstrate considerably higher energy efficiency than the conventional technologies but are also more capital cost-intensive. The effect of scale on CO<sub>2</sub> avoidance costs was not clear due to lack of data points and the possibility that some smaller scale technologies are more advanced than the larger scale technologies reported.*

Our results also show that CO<sub>2</sub> compression and distributed transport costs could be significant. For the distance of 30 km, 10 €/tonne was calculated for scales below 500 tCO<sub>2</sub>/day and more than 50 €/tonne for scales below 5 tCO<sub>2</sub>/day (equivalent to 1 MW<sub>LHV</sub> natural gas). CO<sub>2</sub> compression accounted for the vast majority of these costs. Transport distance was found to have limited influence. The results indicate that the reduction of

CO<sub>2</sub> compression cost, which is closely linked to the CO<sub>2</sub> transport mode, may become a key to economical CO<sub>2</sub> capture for smaller scale energy systems.

The results of this article highlight the possibilities for cost-effective CO<sub>2</sub> capture in distributed energy systems in case CO<sub>2</sub> transport and storage infrastructure has been developed. The costs of CO<sub>2</sub> capture are not prohibitively expensive even in the ST/MT and have a significant cost reduction potential in the LT. CCS from distributed emission sources would also contribute to the economies of scale of CO<sub>2</sub> transport and storage. However, further research is required to have better understanding on the potential of CO<sub>2</sub> capture in distributed energy systems. Examples are: (1) analysis on the minimization of CO<sub>2</sub> compression and distributed transport costs by comparing various system configurations, and (2) exploring the possibilities to extend annual operation time for DH CHP plants to enable more cost-effective CO<sub>2</sub> capture.

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## Summary and conclusions

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This chapter summarizes the background, scope and objectives of this thesis and the main results of the research presented in previous chapters.

### 7.1 Background

Global warming is unequivocal from the observations of changes in atmosphere, cryosphere and oceans. It is reported that most of the global average warming observed since the mid-20<sup>th</sup> century is very likely due to anthropogenic greenhouse gases (GHG) increases. Among anthropogenic greenhouse gases, carbon dioxide (CO<sub>2</sub>) is the most important. The primary source of the increased atmospheric concentration of CO<sub>2</sub> since the pre-industrial period is fossil fuel use. For global average temperature increases exceeding 1.5 to 2.5 °C and concomitant atmospheric CO<sub>2</sub> concentration increases, major changes in nature systems with predominantly negative consequences for biodiversity and ecosystem goods and services such as water and food supply are projected.

To reduce climate change risks, impacts and damages, to date over hundred countries have adopted a global warming target limiting the average temperature increase to 2 °C or below compared to pre-industrial levels as a guiding principle for mitigation efforts. In order to achieve such a target, a portfolio of mitigation measures will be needed. CO<sub>2</sub> capture and storage (CCS) is considered an important option to reduce GHG emissions given the expected large-scale use of fossil fuels over the coming decades.

CCS comprises the capture of CO<sub>2</sub> from industrial and energy-related sources, transport of CO<sub>2</sub> to an underground storage location and long-term isolation from the atmosphere. Cost of CO<sub>2</sub> capture is the largest component of overall CCS costs for most large emission sources and thus, significant amount of research has been performed to reduce them.

Large scale centralized power plants have been the main focus of research, development, demonstration and deployment (RDD&D) for CO<sub>2</sub> capture. In addition, the potential for CCS in the industry and petroleum refineries is expected to be very significant. CCS has a large potential in these sectors mainly for two reasons. Firstly, these sectors today account for nearly 40% of total global CO<sub>2</sub> emissions directly and indirectly. Secondly, there are many industrial processes that generate gas streams rich in CO<sub>2</sub>, or in some cases pure CO<sub>2</sub>, which could increase the cost-effectiveness of CCS by reducing the amount of effort needed to separate CO<sub>2</sub> from flue gases.

In the future, a large transition from centralized power generation to distributed generation (DG) may be seen, particularly in the economies where deregulation takes place. Such a transition would lead to an increase in CO<sub>2</sub> emissions from smaller point sources. It is estimated that CO<sub>2</sub> emissions from distributed energy systems may become significant,

accounting for about 20% for current global energy-related CO<sub>2</sub> emissions in 2030 (Chapter 6). Large-scale deployment of combined heat and power plants (CHP), a main DG option leads to considerable reductions in primary energy consumption and CO<sub>2</sub> emissions. The increase of DG capacity, however, also means an increase in emissions from smaller scale CO<sub>2</sub> emission sources. If stringent CO<sub>2</sub> emissions reduction is needed, DG systems will have to consider options that enable large CO<sub>2</sub> emissions reduction, such as CCS.

In this context, a number of gaps in knowledge were identified in this thesis:

**(1) Techno-economic performance of CO<sub>2</sub> capture from industrial processes**

There are an increasing number of feasibility studies on CO<sub>2</sub> capture from industrial processes. However, a comparative assessment of proposed technologies has not yet been performed using consistent system boundaries and standardized parameters.

**(2) The effect of CO<sub>2</sub> capture on air pollutant emissions from industrial processes**

The evaluation of CCS to reduce CO<sub>2</sub> emissions should take into account the potential effect for other emissions. Such an assessment has been performed for the power sector, but not yet for the industry.

**(3) Techno-economic performance of CO<sub>2</sub> capture from distributed energy systems**

Limited attention has been paid to the performance of CO<sub>2</sub> capture technologies for distributed energy systems, explicitly taking into account differences in timeframe, fuel type, energy system type, plant scale and operational conditions.

## 7.2 Scope and objectives of this thesis

In this context, the objective of this thesis is to assess technical and economic performance for CO<sub>2</sub> capture from industrial plants and distributed energy systems. The research questions (RQs) are as follows:

*RQ.1: What are technical and economic possibilities for CO<sub>2</sub> capture in key industrial sectors?*

*RQ.2: What are the possible impacts of CO<sub>2</sub> capture technologies on the emission of air pollutants (NO<sub>x</sub>, SO<sub>2</sub>, particulate matter (PM) and NH<sub>3</sub>) from key industrial processes?*

*RQ.3: What are technical and economic possibilities for CO<sub>2</sub> capture in distributed energy systems?*

RQs 1 and 2 are addressed in Chapters 2 and 3, respectively. RQ.3 is principally addressed in Chapter 6, with Chapters 4 and 5 comprising part of it by addressing specific issues

regarding CO<sub>2</sub> capture from distributed energy systems. In the following section, the RQs are answered by presenting a summary of the key findings from previous chapters.

In this thesis, natural gas price was assumed 7.3-8 €/GJ<sub>LHV</sub> for the short-mid term (ST/MT: up to 2025) and 8 €/GJ<sub>LHV</sub> for the long-term future (LT: beyond 2030). Coal price is assumed 2.6 €/GJ<sub>LHV</sub> for all timeframes. The interest rate is assumed 10-12% for the industry and 10% for other sectors. The economic lifetime of plants are assumed 35 years for coal-fired energy systems and 20-25 years for other installations.

### 7.3 Summary and conclusions

#### ***Research question 1: What are the technical and economic possibilities for CO<sub>2</sub> capture in key industrial processes?***

This research question is addressed in Chapter 2, which presents a techno-economic assessment and comparison of CO<sub>2</sub> capture technologies for iron and steel, cement, and refineries and petrochemical sectors. The assessment is based on an extensive literature review, covering studies from both industries and academia. Key parameters, e.g. capacity factor, energy prices, interest rate, economic plant lifetime, CO<sub>2</sub> compression pressure, and grid electricity CO<sub>2</sub> intensity, were standardized to enable a fairer comparison of technologies. The analysis is also focusing on the changes in energy, CO<sub>2</sub> emissions and material flows, due to the deployment of CO<sub>2</sub> capture technologies.

The findings indicate that in the *iron and steel sector* in the short-mid term future (ST/MT), a CO<sub>2</sub> avoidance cost of 40-65 €/tCO<sub>2</sub> at a CO<sub>2</sub> avoidance rate of about 0.7-0.8 tCO<sub>2</sub>/t hot rolled coil, or 40-45% of the total carbon input to the plant, are possible with top gas recycling blast furnace (TGRBF) technique. This CO<sub>2</sub> capture technique can be retrofitted to existing blast furnaces, but the costs are likely to be somewhat higher. In addition, it is important to note that the uncertainty of the cost performance is high ± 25 €/tCO<sub>2</sub> avoided for CO<sub>2</sub> capture with VPSA), because the switch from air-blown BF to TGRBF changes the process energy and material flows considerably. Add-on CO<sub>2</sub> capture for air-blown blast furnace may also enable CO<sub>2</sub> capture at similar costs (40-65 €/tCO<sub>2</sub> avoided), but the CO<sub>2</sub> avoidance rate will be only about 0.3-0.4 tCO<sub>2</sub>/t hot rolled coil. CO<sub>2</sub> capture from the COREX process may be done at costs less than 30 €/tCO<sub>2</sub> avoided. However, in order to achieve significant reduction in CO<sub>2</sub> emissions, the COREX process gas needs to be shifted. Advanced add-on CO<sub>2</sub> capture technologies that may emerge in the long term future (LT), e.g., membrane separation and hydrate crystallization, may not offer significant economic advantages over conventional technologies (30-55 €/tCO<sub>2</sub> avoided). Selective carbon membranes will enable CO<sub>2</sub> capture from air-blown BF at around 30 €/tCO<sub>2</sub> avoided and the hydrate crystallization technology, which consumes significant amount of electricity, may become an attractive option when low-cost, low-carbon electricity is available.

When building new plants, smelting reduction technologies such as COREX process may become a strong competitor to conventional blast furnace based steel making process in a carbon-constrained society because of its low-cost CO<sub>2</sub> capture. The results also showed that the COREX process with CO<sub>2</sub> capture offers lower steel production cost and lower specific CO<sub>2</sub> emissions compared to the conventional blast furnace based process. Moreover in the LT, advanced smelting reduction technologies will incorporate CO<sub>2</sub> capture as part of production efficiency improvement while achieving significantly lower steel production costs compared to existing technologies. Although conventional iron and steel making using BF is expected to dominate the market in the LT, strong need for drastic CO<sub>2</sub> emissions reduction may drive the sector towards large scale implementation of advanced smelting reduction technologies.

It should also be noted that the DRI process may become important option for low-carbon steel production because it already removes CO<sub>2</sub>. A combination of DRI and smelting reduction with CO<sub>2</sub> removal may offer opportunities for low-cost CCS already in the ST/MT when the CO<sub>2</sub> transport and storage infrastructure is built.

In the *cement sector*, post-combustion CO<sub>2</sub> capture may likely be the only commercially available technology for the cement sector. With regard to the use of MEA as CO<sub>2</sub> capture solvent, the most economic option seems to be the import of steam from neighboring power plants, which may enable about 65-70 €/tCO<sub>2</sub> avoided at a CO<sub>2</sub> avoidance rate of 0.6 tCO<sub>2</sub>/t clinker, which is about 60% of onsite CO<sub>2</sub> emissions. When the imported steam has a high economic value, the CO<sub>2</sub> avoidance cost may increase up to 90-135 €/tonne. Our results also indicate that it may be more economical not to capture CO<sub>2</sub> from the steam plant (boiler or CHP). The results also indicate that when the use of state-of-the-art CO<sub>2</sub> capture solvent (e.g. KS-1) is combined with the import of power plant steam, the CO<sub>2</sub> avoidance cost may reach below 50 €/tonne.

In the LT, a number of technologies may enable 25-55 €/tCO<sub>2</sub> avoided. Oxyfuel combustion of clinker kiln and precalciner will likely be commercialized, enabling CO<sub>2</sub> capture at around 40 €/tCO<sub>2</sub> avoided. CaO-looping CO<sub>2</sub> capture only from the precalciner may achieve even lower costs (25-30 €/tCO<sub>2</sub> avoided). Post-combustion capture using advanced solvents may also develop significantly with LT cost of 35-55 €/tCO<sub>2</sub> avoided. The cost performance uncertainty is smaller compared to that for the iron and steel sector, because CO<sub>2</sub> capture does not affect either raw material inputs or valuable product outputs.

For *petroleum refineries and petrochemical sectors*, our findings suggests that oxyfuel CO<sub>2</sub> capture (50-60 €/tCO<sub>2</sub> avoided) is more economical than post-combustion capture (>70 €/tCO<sub>2</sub> avoided), though largely dependent on the economic evaluation of the consumed steam, for both combined heaters and furnaces of 2 MtCO<sub>2</sub>/yr scale and catalytic crackers of 1 MtCO<sub>2</sub>/yr scale. However, the retrofit construction for oxyfuel CO<sub>2</sub> capture from heaters and furnaces are expected to be more complicated than that for boilers. The state-of-the-art solvents and other advanced solvents currently under development may bring down the post-combustion capture costs considerably. In the

longer term, oxyfuel CO<sub>2</sub> capture with an integrated ASU/CHP plant may become most economical. An onsite CHP plant supplies power for ASU and CO<sub>2</sub> compression, and steam which will replace some of the existing boilers. The CO<sub>2</sub> avoidance cost may become 30 €/tonne or lower, but due to the large excess electricity production, the overall economical performance will strongly be dependent on the conditions of the power market. The CO<sub>2</sub> avoidance rate, calculated around 70-80 % under nominal parameter values, will also depend largely on the credits that can be obtained from the exporting electricity to the grid.

Crude estimates of technical potentials for worldwide CO<sub>2</sub> emissions reduction for 2030 were made for the industrial processes investigated with the short-mid term technologies. The technical potential for CO<sub>2</sub> reduction from the industrial processes investigated altogether sums up to about 4 Gt/yr: 1 Gt/yr for the iron and steel sector, about 2 Gt/yr for the cement sector, and up to 1 Gt/yr for petroleum refineries.

Note that the actual deployment level in 2030 will likely be much smaller. The IEA's CCS roadmap (IEA, 2008a) estimates that actual deployment for the industry and petroleum refineries in the BLUE Map scenario would be around 0.8 Gt/yr altogether: 0.2 Gt/yr for the iron and steel sector, 0.15 Gt/yr for the cement sector, 0.1 Gt/yr from the petrochemical sector (from ammonia production) and 0.3 Gt/yr for the fuel transformation sector.

Finally, three findings should be pointed out. First, for all industrial sectors, there is no single CO<sub>2</sub> capture technology that is likely to dominate future capture of CO<sub>2</sub>. Because different capture technologies affect the industrial process in different ways, their respective economic performance will depend largely on operational conditions of individual plants. Second, CO<sub>2</sub> capture from industrial processes can significantly increase the electricity exports to the grid (chemical absorption capture and advanced oxyfuel combustion capture using OCM). Because of this, the overall CO<sub>2</sub> capture performance may become largely dependent on the conditions of the local power market. Thus a good integration of industrial plants and power plants is essential for cost-effective CO<sub>2</sub> capture. Finally, potential synergies between CO<sub>2</sub> capture technologies and industrial processes would lead to more efficient industrial production. Exploiting such synergies may become key to a successful CCS deployment in the industrial sector.

***Research question 2: What are the possible impacts of CO<sub>2</sub> capture technologies on the emissions of air pollutants (NO<sub>x</sub>, SO<sub>2</sub>, PM and NH<sub>3</sub>) from key industrial processes?***

This research question is addressed in Chapter 3, where the potential impact of CO<sub>2</sub> capture on the emission of NECD substances (PM, SO<sub>2</sub>, NO<sub>x</sub> and NH<sub>3</sub>) in Europe (EU-27) from key industrial sectors in the short term (timeframe: 2020) was quantitatively assessed. The following industrial sectors were investigated: cement, petroleum refining (furnaces and boilers), and iron and steel (integrated steel plants). The impact on onsite

emissions as well as emissions associated with the increase or decrease of centralized electricity production due to CO<sub>2</sub> capture was assessed. For the cement sector and for furnaces and heaters used in petroleum refineries, post-combustion CO<sub>2</sub> capture using monoethanolamine (MEA) was considered. For the iron and steel sector, TGRBF with CO<sub>2</sub> capture using vacuum pressure swing adsorption (VPSA) was selected in our analysis. For post-combustion CO<sub>2</sub> capture, different solvent regeneration heat supply options were investigated. The main differences are: (1) fuel type (natural gas or coal) and (2) the treatment of CO<sub>2</sub> from the heat supply plant (vented or captured together with CO<sub>2</sub> from the industrial process).

Our results show that short-term CO<sub>2</sub> capture may affect SO<sub>2</sub> and NH<sub>3</sub> emissions considerably. When all three industrial processes in the EU-27 are fully equipped with CO<sub>2</sub> capture, total industrial SO<sub>2</sub> emissions in EU-27 may decrease by 90%. About half of the decrease in SO<sub>2</sub> emissions is attributable to onsite emissions reduction and the rest to avoided emissions from centralized power plants by exporting electricity to the grid. Total industrial NH<sub>3</sub> emissions may increase by 900%, or about 12% of total European NH<sub>3</sub> emissions. It is, however, important to note that a water-wash of CO<sub>2</sub> scrubber outlet gas was not considered; it can remove 95% of the NH<sub>3</sub>. Industrial NO<sub>x</sub> emissions may also increase by up to 20%, which is attributable to the cement sector when coal-fired CHP plants are used to supply CO<sub>2</sub> capture energy. The total changes in NECD substance emissions due to CO<sub>2</sub> capture in the short-term were found to be attributable to the cement and the petroleum refinery sectors, not to the iron and steel sector.

Regarding individual industrial processes, for the cement sector the changes in NO<sub>x</sub> emissions ranges between a 5% decrease and a 12% increase compared to total EU-27 industrial emissions, largely depending on the CO<sub>2</sub> capture heat supply plant option. SO<sub>2</sub> emissions were found to decrease by 20-50% compared to total EU-27 industrial emissions, when fitted with CO<sub>2</sub> capture. For the petroleum refining sector, the reduction in SO<sub>2</sub> emissions amounted up to about 20% of the total EU-27 industrial emissions. The changes in PM emissions were found to be marginal compared to total EU-27 industrial emissions for both sectors. The study also found that in the case of post-combustion capture for the cement and the petroleum refining sectors, avoided centralized power plant emissions, in particular SO<sub>2</sub> and PM, due to electricity export from the CHP plant are as important as changes in onsite emissions.

For the iron and steel sector, the effect of CO<sub>2</sub> capture was found to be limited. SO<sub>2</sub> emissions may increase by about 5% of total EU-27 industrial emissions, but other NECD substance emission levels were found to remain unaffected. The sector-wide reduction of SO<sub>2</sub> emissions is also negligible compared to total EU-27 industrial emissions. CO<sub>2</sub> capture from blast furnace does not largely affect the most emitting processes within the integrated steel plant such as pelletizing, sintering and coking processes.

The results presented in chapter 3 are for specific CO<sub>2</sub> capture technologies in the cement, petroleum refinery, and iron and steel sector, and therefore do not necessarily apply to other CO<sub>2</sub> capture technologies or sectors.

***Research question 3: What are the technical and economic possibilities for CO<sub>2</sub> capture in distributed energy systems?***

This research question is addressed in Chapters 4, 5 and 6. In Chapter 6 the technical and economic performance of CO<sub>2</sub> capture technologies is assessed taking into account the diversity in type of generator technologies used, applications, plant scales, operational patterns, and timeframe in which the technologies may become commercial. The results from Chapter 4 and 5 address specific aspects of CO<sub>2</sub> capture from distributed energy systems. Results are compared to those for large-scale centralized power plants to assess the effect of the economies of scale. It was also investigated how the CO<sub>2</sub> compression and distributed transport of captured CO<sub>2</sub> between the capture site and trunk pipeline contribute to total costs. The study covers CHP plants, boilers and distributed hydrogen plants.

**Short-mid term performance (2020-2025)**

Regarding the technical performance, energy efficiency reduction due to CO<sub>2</sub> capture (energy penalty:  $\Delta\eta$ ) for ST CO<sub>2</sub> capture technologies were found to vary significantly by fuel type, energy system type and its application (Chapter 6). For natural gas-fired systems,  $\Delta\eta$  was found to range between 28% for boilers and 11% for natural gas combined cycle- (NGCC-) CHP plants with good heat integration. For coal-fired plants, a  $\Delta\eta$  of 33% was observed for coal boilers and around 23-28% for pulverized coal (PC) power plants.

Distributed energy systems can significantly reduce  $\Delta\eta$  by system integration, which can be achieved differently for industrial plants and district heating plants. For industrial CHP plants, idle capacity can be used to meet CO<sub>2</sub> capture energy requirements. Industrial CHP plants are often operated at partial load conditions. If CO<sub>2</sub> from these plants is to be captured, the increased load due to CO<sub>2</sub> capture improves power plant efficiency and consequently, part of the efficiency reduction would be offset. Chapter 4 shows that  $\Delta\eta$  for industrial NGCC-CHP plants becomes significantly lower than that for CO<sub>2</sub> capture for 400 MW<sub>e</sub> NGCC condensing plants: up to 16% lower for the short term<sup>41</sup> and up to 12% in the mid-term future<sup>41</sup>.

For district heating- (DH-) CHP plants, a significant cost reduction could be expected by recovering some of the energy consumed for CO<sub>2</sub> capture in the form of low-grade heat for DH. Chemical absorption capture for gas- and coal-fired DH-CHP plants and oxyfuel combustion capture for coal-fired DH-CHP plants are investigated in Chapter 6. For

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<sup>41</sup> In Chapter 4, short term was defined around 2010 and mid-term future between 2020 and 2025. As we noted in Chapter 6, however, we consider the estimates to be on the optimistic side taking into account the recent developments in CCS R&D. Therefore, the results for mid-term future would likely be realized in a somewhat longer term.

chemical absorption capture, literature suggests that up to 30% of the solvent regeneration heat can be recovered in the form of hot water for DH without additional capital cost. For oxyfuel combustion capture, it was suggested that 30% of the energy consumption for ASU and CO<sub>2</sub> compression can be recovered and about 5% reduction of capital cost can be achieved by heat integration.

The results with respect to the economic performance show that CO<sub>2</sub> avoidance costs may range between 40-140 €/tonne for plant scales larger than 100 MW<sub>LHV</sub> (fuel input) and 50-160 €/tonne for 10-100 MW<sub>LHV</sub> in the short term (Chapter 6). The more promising options identified are: industrial NGCC-CHP plants with chemical absorption capture in low heat-to-power ratio (HPR) and low load operations, oxyfuel combustion capture from PC-CHP plants, and some refinery gas-fired plants with oxyfuel combustion capture. The effect of scale is mainly attributable to the capital costs for the energy system and CO<sub>2</sub> capture process and not to the technical performance, and it varies largely by CO<sub>2</sub> capture technology. The analysis on PC-CHP plants with chemical absorption capture and oxyfuel capture performed in Chapter 6 is a good example. For 550 MW<sub>e</sub> condensing plants, CO<sub>2</sub> avoidance costs for the two technologies were comparable: 39 €/tonne for chemical absorption and 33 €/tonne for oxyfuel combustion. However, when they are applied to 50 MW<sub>LHV</sub> CHP plants, chemical absorption capture costs 79 €/tonne while oxyfuel combustion capture is about 20 €/tonne cheaper at 58 €/tonne.

The effect of heat integration between CHP plants and CO<sub>2</sub> capture on CO<sub>2</sub> avoidance costs is significant, but it seems to pay off only for industrial CHP plants operated under specific conditions. CO<sub>2</sub> avoidance costs for industrial NGCC-CHP plants may become comparable to those for 400 MW<sub>e</sub> NGCC plants at scales as small as 100 MW<sub>e</sub> in the short term (41-44 €/t CO<sub>2</sub>) and as small as 200 MW<sub>e</sub> in the mid-term future (33-36 €/t CO<sub>2</sub>). For DH-CHP plants, the cost reduction due to the reduction of the energy efficiency ( $\Delta\eta$ ) amounts up to 8-10 €/tCO<sub>2</sub> avoided, but it is outweighed by the lower capacity factor.

For both industrial and DH applications, chemical absorption CO<sub>2</sub> capture may not be suitable for CHP plants operated at high HPR due to conflict for heat. There are two options to meet both electricity and heat demands while capturing CO<sub>2</sub> at high capture efficiency (85-90%): (1) use auxiliary boiler for additional heat, and (2) build a larger CHP plant with increased electricity generation capacity. The first option is not recommended because of increase in CO<sub>2</sub> capture energy penalty as only low-grade heat is generated from a high quality fuel. This is demonstrated in Chapter 4 for industrial NGCC-CHP plants. The second option requires larger capital requirement and the increased electricity generation can also be problematic if the electricity market is unfavorable for exporting the excess electricity to the grid. In addition, for DG systems there is a technical limit to the electrical capacity. Therefore, this option is also unrealistic.

### **Long term performance (after 2030)**

Our literature review has shown that the energy penalty due to CO<sub>2</sub> capture ( $\Delta\eta_{pen}$ ) was found to range around 4-9%. Regarding the economic performance, it was concluded that



in the LT, CO<sub>2</sub> avoidance costs would be around 10-90 €/tonne for plant scales larger than 100 MW<sub>LHV</sub>, 25-100 €/tonne for 10-100 MW<sub>LHV</sub> and 35-150 €/tonne for 10 MW<sub>LHV</sub> or smaller (Chapter 6). The effect of the capacity factor is significant for some technologies that demonstrate considerably higher energy efficiency than conventional technologies but are also more capital cost-intensive, e.g. Solid Oxide Fuel Cell (SOFC) and membrane-based hydrogen production plants. The effect of scale on the cost of CO<sub>2</sub> capture itself is limited because many advanced energy and CO<sub>2</sub> capture systems are highly modular.

Although advanced energy conversion technologies are projected to enable cheap CO<sub>2</sub> capture, their future is highly uncertain. For example, SOFC is expected to enable cheap CO<sub>2</sub> capture. However, SOFC is still in its demonstration phase and a considerable amount of further effort is required for SOFC systems to compete in the energy market. These types of advanced energy conversion technologies would, therefore, become viable only when fitted with CO<sub>2</sub> capture. Chapter 5 investigates whether CO<sub>2</sub> capture using currently commercial or pre-commercial technologies would facilitate the market penetration of SOFC in a carbon-constrained society. The fuel cell stack production cost required for 5 MW<sub>e</sub> industrial SOFC-CHP systems to become economically competitive with and without CO<sub>2</sub> capture was compared under probable CO<sub>2</sub> price conditions. For CO<sub>2</sub> capture, an oxyfuel afterburner using commercially available pressure/vacuum swing adsorption (PSA/VSA) technology for air separation was considered. The economic results showed that in the absence of a CO<sub>2</sub> price, the cell stack production cost (CSPC) required for SOFC-CHP without CO<sub>2</sub> capture to outperform conventional GT-CHP was as low as 140 \$/kW, which is a very ambitious target for the SOFC geometry considered in the study (advanced flat-tubular). For CO<sub>2</sub> prices above 43 \$/tonne, industrial SOFCs were found to be more economical when equipped with CO<sub>2</sub> capture and the effect of CO<sub>2</sub> capture on CSPC became significantly larger at a CO<sub>2</sub> price of 100 \$/tonne. CSPC required to economically compete with the conventional GT-CHP was as high as 930 \$/kW for SOFC-CHP with CO<sub>2</sub> capture, while it was 350/kW for those without CO<sub>2</sub> capture. The results suggest that to enhance the market penetration of SOFC systems in a carbon-constrained society, it is worthwhile for SOFC manufacturers not only to focus on the SOFC systems itself but also on the development of SOFC systems with integrated CO<sub>2</sub> capture.

#### **Potential bottleneck: CO<sub>2</sub> compression and distributed transport**

Although CO<sub>2</sub> transport was not a main issue of the thesis, the costs for processing and delivering the captured CO<sub>2</sub> to a trunk pipeline or other means of collective CO<sub>2</sub> transport is affected by the scale of CO<sub>2</sub> capture. In our analyses, the aim therefore was to quantify these costs as well as the effect of scale. The results from Chapter 6 indicate that CO<sub>2</sub> compression and distributed transport cost may become a bottleneck due to its high cost at smaller scales. For a distance of 30 km, 10 €/tonne CO<sub>2</sub> was calculated for scales below 500 tCO<sub>2</sub>/day and more than 50 €/tonne for scales below 5 tCO<sub>2</sub>/d (equivalent to 1 MW<sub>LHV</sub> natural gas). CO<sub>2</sub> compression accounted for the vast majority of these costs. The cost of distributed transport itself was found to be limited for all scales.

The findings of this chapter highlight the possibilities for cost-effective CO<sub>2</sub> capture in distributed energy systems in case CO<sub>2</sub> transport and storage infrastructure is developed. The costs of CO<sub>2</sub> capture are not prohibitively expensive even in the short term and have a significant cost reduction potential in the long term. CCS should be considered for distributed emission sources; it could also contribute to the economies of scale of CO<sub>2</sub> transport and storage.

#### **7.4 Final remarks and recommendations for future research**

This thesis identified a number of opportunities for CCS in the industry and distributed energy systems. Various technologies are assessed and compared, and promising technologies were identified. The results obtained throughout the thesis strongly indicate that any industrial and distributed CO<sub>2</sub> emission sources should be included for the consideration of CCS both in the short term and the long term future. At the same time, it was shown that the feasibility of CO<sub>2</sub> capture in the industry and distributed energy systems is more case-dependent than in the centralized power sector.

In each chapter, recommendations for further research have been formulated to obtain a better understanding on the feasibility of CO<sub>2</sub> capture in the industry and distributed energy systems. Key recommendations are summarized below:

- Case studies on CCS in industrial plants and distributed energy systems under area-specific energy market conditions;
- In-depth analysis on the industrial production losses due to retrofitting CO<sub>2</sub> capture in industrial processes (in particular for technologies, e.g. oxyfuel combustion, that require modification in the core of the industrial processes);
- Assessment of possibilities for the integration of industrial processes and power plants for cost-effective CO<sub>2</sub> emissions reduction;
- Evaluation of CCS potential for distributed emission sources by incorporating the results obtained in this study into national/regional energy economic models (in particular for the long term);
- Investigation and comparison of the effect of CO<sub>2</sub> capture on air pollutant emissions from industrial processes for different capture technologies (e.g., oxyfuel combustion capture, post-combustion capture using advanced solvents and membranes), including a detailed flowsheet modeling for capture technologies that are integrated into the core of industrial processes;
- Assessment of the economic impact on the changes in non-CO<sub>2</sub> emissions due to CO<sub>2</sub> capture from industrial process, particularly for SO<sub>2</sub> and NH<sub>3</sub> emissions;
- Investigation of possibilities to extend annual operation time for DH CHP plants to enable more cost-effective CO<sub>2</sub> capture;
- Analysis on the minimization of CO<sub>2</sub> compression and distributed transport costs by comparing various system configurations.

## Samenvatting en conclusies

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Dit hoofdstuk geeft een samenvatting van de achtergrond, scope en doelstellingen van dit proefschrift, en de belangrijkste resultaten van het onderzoek zoals gepresenteerd in de voorgaande hoofdstukken.

### 1. Achtergrond

De veranderingen in de atmosfeer, cryosfeer en oceanen tonen duidelijk aan dat er sprake is van opwarming van de aarde. Verschillende bronnen rapporteren dat deze opwarming, die vanaf halverwege de 20<sup>e</sup> eeuw wordt geobserveerd, zeer waarschijnlijk te wijten is aan antropogene broeikasgassen (greenhouse gases, GHG). Van deze antropogene broeikasgassen is koolstofdioxide (CO<sub>2</sub>) de belangrijkste. Het gebruik van fossiele brandstoffen is de primaire bron van de verhoogde atmosferische CO<sub>2</sub>-concentratie sinds de pre-industriële periode. Grote veranderingen in ecosystemen worden voorspeld indien de gemiddelde wereldwijde temperatuur met meer dan 1,5-2,5°C stijgt als gevolg van een hogere atmosferische CO<sub>2</sub>-concentratie. Deze veranderingen zullen waarschijnlijk overwegend negatieve gevolgen hebben voor bijvoorbeeld de biodiversiteit, en de diensten en goederen die gerelateerd aan ecosystemen, zoals de water- en voedselvoorziening.

Recent hebben meer dan honderd landen zich ten doel gesteld de gemiddelde mondiale temperatuurstijging te beperken tot 2°C of lager ten opzichte van het niveau van het pre-industriële tijdperk. Deze maximale temperatuurstijging wordt gebruikt als leidend beginsel voor CO<sub>2</sub>-reductiemaatregelen met als doel de risico's, gevolgen en schade van klimaatveranderingen te beperken. Een portfolio met verschillende reductiemaatregelen is vereist om deze doelstelling te bereiken. Gezien het verwachte grootschalige gebruik van fossiele brandstoffen in de komende decennia wordt CO<sub>2</sub>-afvang en -opslag (carbon capture and storage, CCS) beschouwd als een belangrijke optie om de emissie van broeikasgassen te verminderen.

De term CCS omvat het afvangen van CO<sub>2</sub> uit industriële en energiegerelateerde bronnen, het transport van CO<sub>2</sub> naar een ondergrondse opslagplaats, en de lange termijn afzondering van CO<sub>2</sub> van de atmosfeer. De kosten van CO<sub>2</sub>-afvang zijn bij de meeste grootschalige emissiebronnen de voornaamste kostencomponent van de totale CCS-kosten. Derhalve is er al veel onderzoek gedaan naar de mogelijkheden deze kosten te verminderen.

Grootschalige, centrale elektriciteitscentrales zijn de belangrijkste focus van RDD&D (Research, Development, Deployment and Diffusion; Onderzoek, Ontwikkeling, Uitrol en Verspreiding) op het gebied van CCS. Tevens is het potentieel voor CCS in de industrie en in de olieraffinaderijen naar verwachting zeer significant. De reden dat CCS een groot potentieel heeft in deze sectoren is tweeledig. Ten eerste zijn deze sectoren vandaag de dag goed voor bijna 40% van de totale directe en indirecte wereldwijde CO<sub>2</sub>-emissies. Ten tweede zijn er veel industriële processen die CO<sub>2</sub>-rijke, of zelfs CO<sub>2</sub>-pure, gasstromen

produceren die de kosteneffectiviteit van CCS zouden kunnen verbeteren door het scheiden van CO<sub>2</sub> uit rookgassen makkelijker te maken.

In de toekomst zal zich mogelijkwerwijs een grote transitie voordoen van centrale naar gedecentraliseerde elektriciteitsopwekking (distributed generation, DG), met name in economieën waar deregulering plaatsvindt. Een dergelijke overgang zal leiden tot een toename van CO<sub>2</sub>-emissies afkomstig van kleinere puntbronnen. Volgens schattingen zullen in de toekomst de CO<sub>2</sub>-emissies van gedecentraliseerde energieconversiesystemen significant stijgen naar ongeveer 20% van de mondiale energiegerelateerde CO<sub>2</sub>-emissies in 2030 (Hoofdstuk 6). Grootschalige inzet van warmtekrachtkoppeling (combined heat and power, CHP), een van de belangrijkste DG-opties, leidt tot aanzienlijke verminderingen van het primaire energieverbruik en de CO<sub>2</sub>-emissies. De toename van DG-capaciteit betekent echter ook een toename van emissies van kleinschalige CO<sub>2</sub>-emissiebronnen. Indien een stringente CO<sub>2</sub>-emissiereductie vereist is, zullen voor DG-systemen opties overwogen moeten worden die grootschalige CO<sub>2</sub>-reductie mogelijk maakt, zoals CCS.

In dit kader zijn in dit proefschrift een aantal kennislacunes geïdentificeerd:

**(1) Techno-economische prestatie van CO<sub>2</sub>-afvang uit industriële processen.**

Er is een stijgende trend waarneembaar in het aantal haalbaarheidsstudies over CO<sub>2</sub>-afvang uit industriële processen. Een vergelijking van de verschillende afvangtechnologieën op basis van consistente systeemgrenzen en gestandaardiseerde parameters is echter nog niet uitgevoerd.

**(2) Het effect van CO<sub>2</sub>-afvang op luchtverontreinigende emissies van industriële processen.** Bij de evaluatie van CCS als CO<sub>2</sub>-reductieoptie moet rekening gehouden worden met potentiële effecten op andere emissies. Een dergelijke analyse is uitgevoerd voor de elektriciteitssector, maar nog niet voor de industrie.

**(3) Techno-economische prestatie van CO<sub>2</sub>-afvang uit gedecentraliseerde energieconversiesystemen.** Beperkte aandacht is besteed aan de prestaties van CO<sub>2</sub>-afvangtechnologieën voor gedecentraliseerde energieconversiesystemen, waarbij expliciet rekening wordt gehouden met verschillen in tijdsbestek, soort brandstof, type energiesysteem, schaalgrootte en operationele condities.

## **2. Scope en doelstellingen van dit proefschrift**

Binnen deze context is de doelstelling van dit proefschrift om de technische en economische prestaties van CO<sub>2</sub>-afvang uit industriële installaties en gedecentraliseerde energieconversiesystemen te evalueren. De onderzoeksvragen zijn als volgt:

*1: Wat zijn de technische en economische mogelijkheden voor CO<sub>2</sub>-afvang in de belangrijkste industriële sectoren?*

*2. Wat zijn de mogelijke gevolgen van CO<sub>2</sub>-afvang technologieën op de uitstoot van luchtverontreinigende stoffen (NO<sub>x</sub>, SO<sub>2</sub>, (fijn)stof en NH<sub>3</sub>) van de belangrijkste industriële processen?*

*3: Wat zijn de technische en economische mogelijkheden voor CO<sub>2</sub>-afvang in gedecentraliseerde energieconversiesystemen?*

Onderzoeksvragen 1 en 2 komen aan bod in respectievelijk Hoofdstuk 2 en 3. Onderzoeksvraag 3 komt voornamelijk aan bod in Hoofdstuk 6, en gedeeltelijk in Hoofdstuk 4 en 5 waarin de specifieke vraagstukken met betrekking tot CO<sub>2</sub>-afvang van gedecentraliseerde energieconversiesystemen wordt behandeld. In de volgende paragraaf worden de onderzoeksvragen beantwoord door middel van een samenvatting van de belangrijkste bevindingen uit eerdere hoofdstukken.

In dit proefschrift is een aardgasprijs aangenomen van 7,3-8 €/GJ<sub>LHV</sub> voor de korte termijn (tot 2020) en 8 €/GJ<sub>LHV</sub> voor de middellange (2020-2030) en lange termijn (na 2030). De veronderstelde kolenprijs is 2,6 €/GJ<sub>LHV</sub> voor alle termijnen. Voor de rente wordt uitgegaan van 10-12% voor de industrie en 10% voor andere sectoren. De veronderstelde economische levensduur van de fabrieken is 35 jaar voor kolengestookte energiesystemen en 20-25 jaar voor andere installaties.

### **3. Samenvatting en conclusies**

*Onderzoeksvraag 1: Wat zijn de technische en economische mogelijkheden voor CO<sub>2</sub>-afvang in de belangrijkste industriële sectoren?*

Deze onderzoeksvraag wordt behandeld in Hoofdstuk 2, waarin een technisch-economische evaluatie en vergelijking van CO<sub>2</sub>-afvang technologieën voor de ijzer en staal, cement, en raffinage en petrochemie sectoren wordt gegeven. De evaluatie is gebaseerd op een uitgebreid literatuuronderzoek, waarbij zowel studies uit de industrie als uit onderzoeksinstituten zijn meegenomen. Belangrijke parameters, zoals capaciteitsfactor, energieprijzen, rentetarief, economische levensduur, CO<sub>2</sub>-compressiedruk, en CO<sub>2</sub>-intensiteit van het elektriciteitsnet, zijn gestandaardiseerd om een meer eerlijke vergelijking van de technologieën mogelijk te maken. De analyse focust ook op veranderingen in energie- en massastromen en CO<sub>2</sub>-emissies door het toepassen van CO<sub>2</sub>-afvangtechnologieën.

De resultaten laten zien dat in de *ijzer en staalsector* de kosten per ton vermeden CO<sub>2</sub>-emissie (CO<sub>2</sub> avoidance costs) van 40-65 €/tCO<sub>2</sub> bij een CO<sub>2</sub>-vermijdingsfractie (CO<sub>2</sub> avoidance rate) van 0,7-0,8 tCO<sub>2</sub>/t warmgewalste rol, oftewel 40-45% van de totale

koolstofinput van de fabriek, mogelijk is met hoogovens met topgasrecycling (top gas recycling blast furnace, TGRBF) technologie in de korte tot middellange termijn (2020-2025). De onzekerheid in de kostencijfers is hoog ( $\pm 25$  € per ton vermeden CO<sub>2</sub>) doordat de omschakeling van luchtgeblazen hoogoven naar TGRBF leidt tot aanzienlijke veranderingen in de energie- en materiaalstromen. Een luchtgeblazen hoogoven uitbreiden met CO<sub>2</sub>-afvang kan bij vergelijkbare kosten (40-65 € per ton vermeden CO<sub>2</sub>), maar de CO<sub>2</sub>-vermijdingsfractie is dan maar 0,3-0,4 tCO<sub>2</sub>/ per ton warmgewalste rol. CO<sub>2</sub>-afvang van het COREX-proces kan bereikt worden voor minder dan 30 € per ton vermeden CO<sub>2</sub>. Maar, om significante vermindering in CO<sub>2</sub> reducties te bereiken, moet het COREX procesgas eerst aangepast (shift) worden. Geavanceerde end-of-pipe CO<sub>2</sub>-afvang technologieën die op de lange termijn kunnen opduiken (na 2030), zoals membraanscheiding en hydraatkristallisatie, bieden misschien geen significante verbeteringen ten opzichte van conventionele technologieën. Met selectieve koolstofmembranen is CO<sub>2</sub>-afvang bij luchtgeblazen hoogoven mogelijk voor 30 € per ton vermeden CO<sub>2</sub>. De hydraatkristallisatie technologie, die aanzienlijke hoeveelheid energie verbruikt, kan een aantrekkelijke optie worden als er goedkope elektriciteit met een lage koolstofintensiteit beschikbaar is.

Voor het bouwen van nieuwe fabrieken kunnen, in een maatschappij die streeft naar verregaande CO<sub>2</sub> emissiebeperking, smeltreductie processen als het COREX proces door zijn lage kosten voor CO<sub>2</sub>-afvang een aantrekkelijk alternatief zijn voor conventionele hoogovens voor het maken van staal. De resultaten laten ook zien dat het COREX proces met CO<sub>2</sub>-afvang lagere productiekosten en lagere specifieke CO<sub>2</sub>-emissies heeft vergeleken met conventionele technologieën voor staalproductie. Ondanks dat conventionele ijzer- en staalproductie door hoogovens de markt op de lange termijn waarschijnlijk zal domineren, kunnen strenge CO<sub>2</sub>-emissie-eisen de sector dwingen tot grootschalige implementatie van geavanceerde smeltreductie technologieën.

Het direct gereduceerde ijzer (DRI) proces kan een aantrekkelijke optie worden voor staal productie met weinig CO<sub>2</sub>-uitstoot, omdat de CO<sub>2</sub> al verwijderd wordt. Een combinatie van DRI en smeltreductie met CO<sub>2</sub>-afvang kan CCS met lage kosten mogelijk maken op de korte tot middellange termijn indien een CO<sub>2</sub>-transport en -opslag infrastructuur gerealiseerd is.

In de *cementsector* is CO<sub>2</sub>-afvang na verbranding (post-combustion) waarschijnlijk de enige commercieel beschikbare technologie op de korte tot middellange termijn. Met betrekking tot de energieconsumptie voor de regeneratie van het oplosmiddel is, indien aanwezig, de import van laagwaardige stoom van naburige elektriciteitscentrales de meest economische optie. Dit kan leiden tot CO<sub>2</sub>-afvangkosten van ongeveer 65-70 € per ton vermeden CO<sub>2</sub> en een CO<sub>2</sub>-vermijdingsfractie van 0,6 tCO<sub>2</sub> per ton klinker, wat ongeveer 60% van de CO<sub>2</sub>-emissies van de fabriek is. Bij dure geïmporteerde stoom kunnen de CO<sub>2</sub>-vermijdingskosten stijgen tot 90-135 €/tCO<sub>2</sub>. Onze resultaten laten ook zien dat het economischer is om de CO<sub>2</sub> van de stoomproductie (stoomketel of CHP) niet af te vangen. De CO<sub>2</sub>-vermijdingskosten kunnen dalen tot onder 50 €/tCO<sub>2</sub> als geavanceerde

oplosmiddelen (e.g. KS-1) gecombineerd worden met de import van stoom van elektriciteitscentrales.

Op de lange termijn zal verbranding met pure zuurstof (oxyfuel combustion) in klinkerovens en precalciner waarschijnlijk commercieel beschikbaar zijn, wat CO<sub>2</sub>-afvang rond de 40 € per ton vermeden CO<sub>2</sub> mogelijk maakt. CO<sub>2</sub>-afvang door CaO-looping, alleen toegepast bij de precalciner, kan wellicht nog goedkoper (25-30 € per ton vermeden CO<sub>2</sub>). Post-combustion afvang met behulp van geavanceerde oplosmiddelen kan zich mogelijk ook significant ontwikkelen met CO<sub>2</sub>-afvangkosten van 35-55 €/tCO<sub>2</sub> vermeden op de lange termijn. De onzekerheid in de kostencijfers is kleiner vergeleken met die in de ijzer- en staalsector, omdat het toepassen van CO<sub>2</sub>-afvang geen effect heeft op de benodigde grondstoffen en geproduceerde waardevolle eindproducten.

Voor de *olieraffinage en petrochemische sector* duiden onze bevindingen erop dat in de korte tot middellange termijn CO<sub>2</sub>-afvang met zuurstofverbranding met geschatte kosten van 50-60 € per ton vermeden CO<sub>2</sub> economischer is dan post-combustion CO<sub>2</sub>-afvang (>80 € per ton vermeden CO<sub>2</sub>), zowel voor gecombineerde ketels en fornuizen met een schaal van 2Mt CO<sub>2</sub>/jaar, als katalytische krakers met een schaal van 1 Mt CO<sub>2</sub>/jaar. De resultaten zijn echter sterk afhankelijk van de economische evaluatie van de gebruikte stoom. Wij verwachten echter dat het aanpassen van ketels en fornuizen voor CO<sub>2</sub>-afvang met zuurstofverbranding gecompliceerder is dan het aanpassen van stoomketels. Geavanceerde oplosmiddelen die op dit moment in ontwikkeling zijn kunnen de post-combustion CO<sub>2</sub>-afvangkosten aanzienlijk omlaag brengen. Op de langere termijn is CO<sub>2</sub>-afvang met zuurstofverbranding bij een geïntegreerde ASU/CHP (luchtscheidingsinstallatie: Air Separation Unit, ASU) het meest economisch. Een lokale CHP-centrale levert stroom voor de ASU en de CO<sub>2</sub>-compressie terwijl de stoomproductie een gedeelte van de al bestaande stoomketels vervangt. De CO<sub>2</sub>-vermijdingskosten kunnen zakken naar 30 €/tCO<sub>2</sub> of lager, maar door de grote overproductie van elektriciteit is de algehele economische prestatie sterk afhankelijk van de condities van de elektriciteitsmarkt. De CO<sub>2</sub>-vermijdingsfractie die rond de 70-80% ligt, is ook sterk afhankelijk van de credits die worden verkregen door de export van elektriciteit naar het elektriciteitsnet.

Ruwe schattingen zijn gemaakt van het technische potentieel voor wereldwijde CO<sub>2</sub>-emissiereducties in 2030 voor de industriële processen, waarbij gekeken is naar korte tot middellange termijn technologieën. Het totale technische potentieel voor CO<sub>2</sub>-emissiereductie van de onderzochte industriële processen is ongeveer 4 Gt/jaar: 1 Gt/jaar voor de ijzer- en staalsector, ongeveer 2 Gt/jaar voor de cementsector, en 1 Gt/jaar voor de olieraffinage.

Het werkelijke penetratieniveau in 2030 is waarschijnlijk veel kleiner. De CCS roadmap van de IEA schat dat de werkelijke emissiereductie in de industrie en olieraffinage in het BLUE Map scenario ongeveer 0.8 Gt/jaar is: 0.2 Gt/jaar voor de ijzer- en staalsector, 0.15

Gt/jaar voor de cementsector, 0.1 Gt/jaar voor de (petro-)chemische sector (ammonia productie) en 0.3 Gt/jaar voor de brandstof transitie sector.

Ten slotte worden de volgende drie bevindingen benadrukt. Ten eerste is het bij alle industriële sectoren onwaarschijnlijk dat één enkele CO<sub>2</sub>-afvang technologie het afvangen van CO<sub>2</sub> in de toekomst zal domineren. Omdat verschillende afvangtechnologieën de industriële processen verschillend beïnvloeden, zullen de economische prestaties voor een groot gedeelte afhangen van de operationele condities van de individuele fabrieken. Ten tweede kan CO<sub>2</sub>-afvang bij industriële processen de export van elektriciteit naar het elektriciteitsnet significant vergroten (chemische absorptie en zuurstofverbranding bij een geïntegreerde ASU/CHP). Hierdoor zullen de totale CO<sub>2</sub>-afvangprestaties voor een groot gedeelte afhangen van de condities van de lokale elektriciteitsmarkt. Dit maakt een goede integratie tussen industriële installaties en elektriciteitscentrales essentieel voor kosteneffectieve CO<sub>2</sub>-afvang. Tot slot kunnen potentiële synergieën tussen CO<sub>2</sub>-afvangtechnologieën en industriële processen leiden tot een efficiëntere industriële productie. Gebruik maken van zulke synergieën kan een sleutelpositie spelen in een succesvolle uitrol van CCS in de industriële sector.

*Onderzoeksvraag 2: Wat zijn de mogelijke gevolgen van CO<sub>2</sub>-afvang technologieën op de uitstoot van luchtverontreinigende stoffen (NO<sub>x</sub>, SO<sub>2</sub>, (fijn)stof en NH<sub>3</sub>) van de belangrijkste industriële processen?*

Deze onderzoeksvraag wordt behandeld in Hoofdstuk 3 waar de potentiële gevolgen van CO<sub>2</sub>-afvang op de emissies van NECD-stoffen (NO<sub>x</sub>, SO<sub>2</sub>, (fijn)stof, en NH<sub>3</sub>) in Europa (EU-27) bij de belangrijkste industriële processen op de korte termijn (tijdspanne: 2020) kwantitatief is bepaald. De volgende industriële processen zijn onderzocht: cement, olieraffinaderijen (fornuizen en stoomketels), en ijzer en staal (geïntegreerde staalfabrieken). De gevolgen van zowel lokale emissies als emissies gerelateerd aan de toename of afname van centraal opgewekte elektriciteit door het toepassen van CO<sub>2</sub>-afvang zijn beoordeeld. Voor de cementsector en voor fornuizen en ketels die in olieraffinaderijen gebruikt worden is post-combustion CO<sub>2</sub>-afvang met monoethanolamine (MEA) bekeken. Voor de ijzer en staalsector is voor deze analyse gekozen voor zuurstofgeblazen TGRBF met CO<sub>2</sub>-afvang met behulp van vacuüm pressure swing adsorptie (VPSA). Bij post-combustion CO<sub>2</sub>-afvang zijn verschillende hittebronnen voor de regeneratie van het oplosmiddel onderzocht. De belangrijkste verschillen zijn: (1) brandstoftype (aardgas of kolen) en (2) de behandeling van de CO<sub>2</sub> van de warmteproductie-installatie (geëmitteerd of afgevangen met de CO<sub>2</sub> van het industriële proces).

Onze resultaten laten zien dat CO<sub>2</sub>-afvang op de korte termijn een aanzienlijke effect heeft op de SO<sub>2</sub>- en NH<sub>3</sub>-emissies. Als in de EU-27 alle drie de industriële processen volledig worden uitgerust met CO<sub>2</sub>-afvang kunnen de totale SO<sub>2</sub>-emissies van deze processen in de EU-27 dalen met 90%. Ongeveer de helft van die afname komt door lokale



emissiereductie en de rest door vermeden emissies bij gecentraliseerde elektriciteitscentrales door het exporten van elektriciteit naar het elektriciteitsnet. De totale industriële NH<sub>3</sub>-emissies kunnen toenemen met 900%, wat overeenkomt met 12% van de totale Europese NH<sub>3</sub>-emissies. NO<sub>x</sub>-emissies zouden ook kunnen toenemen met maximaal 20%, wat toe te schrijven is aan de cementsector wanneer kolengestookte CHP centrales de energie leveren voor de CO<sub>2</sub>-afvang. Op de korte termijn zijn de totale veranderingen in emissies van NECD-stoffen door CO<sub>2</sub>-afvang toe te schrijven aan de cement en olieraffinage sector en niet aan de ijzer en staalsector.

Wat individuele industriële processen betreft veranderen in de cement sector de NO<sub>x</sub>-emissies tussen een afname van 5% en een toename van 12% vergeleken met de totale EU-27 industriële emissies, afhankelijk van de warmteproductie-installatie voor de CO<sub>2</sub>-afvang. SO<sub>2</sub>-emissies dalen met 20-50% vergeleken met de totale EU-27 industriële emissies als gevolg van CO<sub>2</sub>-afvang. Voor de olieraffinaderijen is de SO<sub>2</sub>-emissiereductie maximaal 20% van de totale EU-27 industriële SO<sub>2</sub>-emissies. De verandering in (fijn)stofemissies in beide sectoren zijn marginaal vergeleken met de totale EU-27 industriële emissies. Uit dit onderzoek komt ook naar voren dat in het geval van post-combustion CO<sub>2</sub>-afvang in de cementsector en olieraffinaderijen, de vermeden emissies, met name SO<sub>2</sub> en (fijn)stof, van gecentraliseerde elektriciteitscentrales door de elektriciteitexport van de CHP centrale, net zo belangrijk zijn als de veranderingen van de lokale emissies.

In de ijzer en staalsector is een beperkt effect van CO<sub>2</sub>-afvang gevonden. SO<sub>2</sub>-emissies kunnen met ongeveer 5% van de totale EU-27 industriële emissies toenemen, maar van de andere NECD-stoffen zouden de emissieniveaus constant blijven. De sectorwijde reductie van SO<sub>2</sub>-emissies is verwaarloosbaar vergeleken met de totale EU-27 industriële emissies. CO<sub>2</sub>-afvang bij hoogovens heeft geen groot effect op de grootste emitterende processen in een geïntegreerde staalfabriek, zoals productie van pellet, sinter en cokes.

De resultaten die in Hoofdstuk 3 gepresenteerd worden zijn voor specifieke CO<sub>2</sub>-afvang technologieën in de cement, olieraffinage, en ijzer en staalsector, en hoeven daarom niet noodzakelijk te gelden voor andere CO<sub>2</sub>-afvang technologieën of sectoren.

*Onderzoeksvraag 3: Wat zijn de technische en economische mogelijkheden voor CO<sub>2</sub>-afvang in gedecentraliseerde energieconversiesystemen?*

Deze onderzoeksvraag wordt behandeld in de Hoofdstukken 4, 5 en 6. In Hoofdstuk 6 worden de technische en economische prestaties van CO<sub>2</sub>-afvang technologieën beoordeeld, waarbij rekening wordt gehouden met de verscheidenheid van verschillende technologieën wat betreft generatortypes, toepassingen, schaalgrootte, gebruikspatronen, en het tijdsbestek waarin de technologieën commercieel kunnen worden toegepast. De resultaten uit Hoofdstukken 4 en 5 behandelen specifieke aspecten van CO<sub>2</sub>-afvang van gedecentraliseerde energieconversiesystemen. De resultaten worden vergeleken met die

voor grootschalige, gecentraliseerde elektriciteitscentrales om het effect van de schaalvoordelen te beoordelen. Daarnaast is de bijdrage aan de totale kosten van CO<sub>2</sub>-compressie, en van het CO<sub>2</sub>-transport van de afvanglocatie naar de centrale CO<sub>2</sub>-pijpleiding, bestudeerd. De studie kijkt naar CHP-installaties, stoomketels en gedecentraliseerde waterstofinstallaties.

### **Prestaties op de korte tot middellange termijn (2020-2025)**

Wat betreft de technische prestaties, blijkt de afname in energie-efficiëntie als gevolg van CO<sub>2</sub>-afvang ( $\Delta\eta$ ) met korte termijn CO<sub>2</sub>-afvang technologieën aanzienlijk te verschillen per brandstoftype, energieconversietechnologie en toepassing (Hoofdstuk 6). Voor aardgasgestookte systemen werd een  $\Delta\eta$  gevonden tussen 28% voor stoomketels en 11% voor NGCC-CHP-installaties met een goede warmte-integratie. Voor kolengestookte installaties, werd een  $\Delta\eta$  van 33% gevonden voor kolenketels en ongeveer 23-28% voor poederkool (pulverized coal: PC) elektriciteitscentrales.

De  $\Delta\eta$  van gedecentraliseerde energieconversiesystemen kan aanzienlijk verminderd worden door systeemintegratie, wat op verschillende manieren bereikt kan worden voor industriële installaties en installaties voor wijkverwarming. Bij industriële CHP-installaties kan onbenutte capaciteit worden gebruikt om energie op te wekken voor de CO<sub>2</sub>-afvang, omdat ze vaak in deellast draaien. Als CO<sub>2</sub> uit deze installaties wordt afgevangen, zal de belasting toenemen, waardoor de efficiëntie verbetert. Hierdoor zal de  $\Delta\eta$  gedeeltelijk gecompenseerd worden. Hoofdstuk 4 laat zien dat  $\Delta\eta$  voor industriële NGCC-CHP-installaties aanzienlijk lager zal worden dan die voor CO<sub>2</sub>-afvang bij een 400 MW<sub>e</sub> NGCC elektriciteitscentrale: tot 16% lager op de korte termijn en tot 12% in de middellange termijn <sup>42</sup>.

Voor stadsverwarming (district heating, DH) CHP-installaties kan een aanzienlijke kostenbesparing verwacht worden door het terugwinnen van een deel van de verbruikte energie voor CO<sub>2</sub>-afvang in de vorm van laagwaardige warmte voor DH. In Hoofdstuk 6 wordt chemische absorptie onderzocht voor gas- en kolengestookte DH CHP-installaties en oxyfuel combustion voor kolengestookte DH CHP-installaties. De literatuur suggereert dat met chemische absorptie tot 30% van de warmte die gebruikt wordt voor de regeneratie van het oplosmiddel kan worden teruggewonnen, zonder extra investeringen, als heet water voor DH. Voor oxyfuel combustion wordt gesuggereerd dat 30% van het energieverbruik van de ASU en CO<sub>2</sub>-compressie kan worden teruggewonnen en de investeringskosten met ongeveer 5% kunnen worden verminderd door warmte-integratie.

De resultaten met betrekking tot de economische prestaties laten zien dat CO<sub>2</sub>-vermijdingskosten kunnen variëren tussen 30-140 €/t voor installaties groter dan 100

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<sup>42</sup> In hoofdstuk 4 is de korte termijn gedefinieerd als 2010 en de middellange termijn tussen 2020 en 2025. Zoals besproken in hoofdstuk 6 lijken deze tijdspannes optimistisch te zijn, gezien de recente ontwikkelingen in CCS R&D. De voorspellingen voor de middellange termijn zullen daarom waarschijnlijk later gerealiseerd worden.

$MW_{LHV}$  (brandstofinput) en 50-150 €/t voor installaties van 10-100  $MW_{LHV}$  op de korte termijn (Hoofdstuk 6). Veelbelovende opties zijn: industriële NGCC-CHP-installaties met  $CO_2$ -afvang door middel van chemische absorptie, opererend onder lage belasting en met een lage warmte-elektriciteit ratio, oxyfuel combustion afvang bij PC-CHP-installaties, en sommige gasgestookte centrales bij raffinaderijen met oxyfuel combustion afvang. Het effect van schaalvergroting is vooral toe te schrijven aan de investeringskosten voor de energiecentrale en de  $CO_2$ -afvanginstallatie, en varieert significant met betrekking tot  $CO_2$ -afvangtechnologie. De analyse in Hoofdstuk 6 van de PC-CHP-installaties met  $CO_2$ -afvang op basis van chemische absorptie en oxyfuel combustion is hiervan een goed voorbeeld. Voor 550  $MW_e$  elektriciteitscentrales bleken de  $CO_2$ -vermijdingskosten voor de twee technologieën vergelijkbaar: 39 €/t voor chemische absorptie en 33 €/t voor oxyfuel combustion. Wanneer ze echter worden toegepast bij een 50  $MW_{LHV}$  CHP-installatie is er een duidelijk verschil: de  $CO_2$ -vermijdingskosten bij chemische absorptie zijn 79 €/t, terwijl oxyfuel combustion bijna 20 €/t goedkoper is met 58 €/t.

Warmte-integratie tussen de CHP en  $CO_2$ -afvang installaties heeft een significant effect op de  $CO_2$ -vermijdingskosten, maar dit lijkt alleen economisch aantrekkelijk voor industriële CHP-installaties. De  $CO_2$ -vermijdingskosten voor industriële NGCC-CHP-installaties kunnen op de korte termijn vergelijkbaar worden met die van een 400  $MW_e$  NGCC elektriciteitscentrale bij een schaalgrootte van 100  $MW_e$  (41-44 €/t  $CO_2$ ), en vergelijkbaar voor een 200  $MW_e$  CHP centrale op de middellange termijn (33-36 €/t  $CO_2$ ). Voor DH CHP-installaties kunnen de  $CO_2$ -vermijdingskosten verlaagd worden met maximaal 8-10 €/t  $CO_2$  als gevolg van een kleinere vermindering van de energie-efficiëntie ( $\Delta\eta$ ). Dit zal tevens gepaard gaan met een lagere capaciteitsfactor, wat warmte-integratie voor deze situatie oninteressant maakt.

Voor zowel industriële als DH toepassingen is  $CO_2$ -afvang met chemische absorptie niet geschikt bij CHP-installaties met een hoge warmte-elektriciteit ratio, door de hoge warmtevraag van zowel de eindgebruiker als het afvangproces. Derhalve wordt dit afgeraden. Er zijn twee mogelijkheden om  $CO_2$ -afvang met een hoog afvangrendement (85-90%) te realiseren waarbij zowel aan de elektriciteit- als aan de warmtevraag wordt voldaan: (1) het gebruik van een extra stoomketel; (2) het installeren van een CHP-installatie met extra elektrisch vermogen. De eerste optie wordt niet aanbevolen, omdat dit tot een grotere energieconsumptie van de  $CO_2$ -afvang leidt doordat er slechts laagwaardige warmte gegenereerd wordt uit een energetisch hoogwaardige brandstof. Dit is aangetoond in Hoofdstuk 4 voor industriële NGCC-CHP-installaties. Nadelen van de tweede optie zijn grotere investeringen, en mogelijke problemen bij levering van extra elektriciteit aan het elektriciteitsnet indien er sprake is van ongunstige marktcondities. Tenslotte bestaat er voor DG-systemen een technische limiet aan de elektrische capaciteit.

### **Prestaties op de lange termijn (na 2030)**

Volgens de literatuur is de vermindering van de energie-efficiëntie ( $\Delta\eta_{pen}$ ) als gevolg van  $CO_2$ -afvang ongeveer 4-9%. Voor sommige geavanceerde energieconversiesystemen zijn in onze analyse huidige state-of-the-art commerciële systemen als referentie gebruikt,

omdat deze geavanceerde systemen inherent met CO<sub>2</sub>-afvang zijn ontworpen en er geen identieke centrales bestaan zonder CO<sub>2</sub>-afvang. Met betrekking tot de economische prestaties werd geconcludeerd dat de CO<sub>2</sub>-vermijdingskosten ongeveer 10-90 €/t voor een centrale met een schaalgrootte groter dan 100 MW<sub>LHV</sub> zullen zijn, 25-100 €/t voor 10-100 MW<sub>LHV</sub> en 35-150 €/t voor centrales van 10 MW<sub>LHV</sub> en kleiner (Hoofdstuk 6). Het effect van de capaciteitsfactor op de kosten is significant voor technologieën die zowel een hoge efficiëntie hebben als kapitaal intensief zijn ten opzichte van conventionele technologieën, zoals solid oxide fuel cell (SOFC) brandstofcellen en waterstofproductie met behulp van membranen. Het effect van schaalvergroting op de kosten van de CO<sub>2</sub>-afvang zelf is beperkt, omdat veel geavanceerde energieopwekkingstechnologieën en CO<sub>2</sub>-afvanginstallaties zeer modulair zijn.

Hoewel verwacht wordt dat geavanceerde energieconversietechnologieën in de toekomst goedkope CO<sub>2</sub>-afvang mogelijk maken, is hun toekomst hoogst onzeker. Een voorbeeld is de SOFC, een type brandstofcel dat naar verwachting goedkope CO<sub>2</sub>-afvang mogelijk zal maken. Echter, SOFC is nog in de demonstratiefase en er zullen aanzienlijke inspanningen nodig zijn om SOFC-systemen te kunnen laten concurreren op de energiemarkt. Dit type geavanceerde energieconversietechnologieën zouden wel eens alleen levensvatbaar kunnen zijn indien ze worden uitgerust met CO<sub>2</sub>-afvang. Hoofdstuk 5 onderzoekt of het afvangen van CO<sub>2</sub> met de huidige commerciële en pre-commerciële technologieën de marktpenetratie van SOFC in een maatschappij met een CO<sub>2</sub> emissiebeperking faciliteert. De productiekosten per SOFC membraammodule die benodigd zijn om concurrerend te zijn, zijn vergeleken voor een SOFC CHP-systeem met en een SOFC CHP-systeem zonder CO<sub>2</sub>-afvang, voor verschillende CO<sub>2</sub>-prijsscenario's. Voor CO<sub>2</sub>-afvang is uitgegaan van een oxyfuel naverbrander met commercieel verkrijgbare pressure/vacuüm swing adsorptie (PSA/VSA)-technologie voor luchtscheiding (air separation unit: ASU). De economische resultaten geven aan dat bij het ontbreken van een CO<sub>2</sub>-prijs de kosten van de membraammodules 140 \$ kW<sup>-1</sup> moeten zijn om beter te presteren dan een conventionele GT-CHP. Dit is een zeer ambitieuze kostendoelstelling voor het type membraammodules (advanced flat tubular) waarvan in deze studie is uitgegaan. Bij CO<sub>2</sub>-prijzen van meer dan 43 \$/t zijn industriële SOFCs goedkoper met CO<sub>2</sub>-afvang dan zonder. Dit effect van CO<sub>2</sub>-afvang op de productiekosten per membraanmodule (cell stack production costs, CSPC) is aanzienlijk groter bij een CO<sub>2</sub>-prijs van 100 \$/t. De cel stack productie kosten die nodig zijn om te concurreren met de conventionele GT CHP was 930 \$ kW<sup>-1</sup> voor SOFC-CHP met CO<sub>2</sub>-afvang, terwijl het 350 \$ kW<sup>-1</sup> is voor dit type installaties zonder CO<sub>2</sub>-afvang. De resultaten suggereren dat, om de marktpenetratie van SOFC-systemen in een maatschappij met een CO<sub>2</sub> emissiebeperking te verbeteren, het voor SOFC-fabrikanten de moeite waard is niet alleen op de SOFC-systemen zelf te richten, maar ook op de ontwikkeling van SOFC-systemen met geïntegreerde CO<sub>2</sub>-afvang.

### **Potentiële knelpunten: CO<sub>2</sub>-compressie en lokaal transport**

Hoewel CO<sub>2</sub>-transport niet een van de belangrijkste onderwerpen van de scriptie is, is het wel meegenomen, omdat de kosten voor de compressie en het transport van de afgevangen CO<sub>2</sub> naar een hoofdpijpleiding, of andere middelen van collectief transport, beïnvloed

worden door de schaal van de CO<sub>2</sub>-afvang. Het doel van de analyse was deze kosten te kwantificeren, evenals het effect van schaalgrootte hierop. Uit de resultaten van Hoofdstuk 6 blijkt dat de kosten van CO<sub>2</sub>-compressie en lokaal transport een knelpunt kunnen vormen bij afvang op kleinere schaal. Voor een afstand van 30 km werden kosten van 10 €/t CO<sub>2</sub> berekend voor hoeveelheden onder 500 tCO<sub>2</sub>/dag en meer dan 50 €/t voor hoeveelheden van minder dan 5 tCO<sub>2</sub>/d (het equivalent van 1 MW<sub>LHV</sub> aardgas). De CO<sub>2</sub>-compressie is verantwoordelijk voor het overgrote deel van deze kosten. De kosten van het lokale transport bleken beperkt te zijn voor alle schaalgroottes.

De resultaten van dit hoofdstuk geven aan dat er mogelijkheden bestaan voor kosteffectieve CO<sub>2</sub>-afvang in gecentraliseerde energieconversiesystemen als er een infrastructuur wordt aangelegd voor CO<sub>2</sub>-transport en -opslag. In een dergelijke situatie is CO<sub>2</sub>-afvang niet onbetaalbaar, zelfs niet op de korte termijn. Op de lange termijn bestaan er aanzienlijke mogelijkheden tot kostenreductie. CCS moet daarom ook worden overwogen voor gedecentraliseerde CO<sub>2</sub>-bronnen; het kan tot schaalvoordelen voor CO<sub>2</sub>-transport en -opslag leiden.

#### **4. Slotopmerkingen en aanbevelingen voor toekomstig onderzoek**

In dit proefschrift zijn aantal mogelijkheden besproken voor implementatie van CCS in de industrie en in gedecentraliseerde energieconversiesystemen. Verschillende technologieën zijn beoordeeld en vergeleken, en veelbelovende technologieën zijn geïdentificeerd. De resultaten van het onderzoek wijzen uit dat alle industriële en decentraliseerde CO<sub>2</sub>-bronnen moet worden meegenomen wanneer CCS wordt overwogen, zowel op de korte als lange termijn. Tegelijkertijd komt uit de resultaten naar voren dat de haalbaarheid van CO<sub>2</sub>-afvang in de industrie en gedecentraliseerde energieconversiesystemen meer afhankelijk is van de specifieke situatie dan bij grote gecentraliseerde elektriciteitscentrales.

In elk hoofdstuk zijn aanbevelingen voor verder onderzoek gedaan om een beter inzicht te krijgen in de haalbaarheid van CO<sub>2</sub>-afvang in de industrie en gedecentraliseerde energieconversiesystemen. De belangrijkste aanbevelingen zijn hieronder samengevat:

- Case studies naar CCS voor de industrie en voor gedecentraliseerde energieconversiesystemen met regio-specifieke marktomstandigheden;
- Een grondige analyse van de productievermindering in de industrie als gevolg van het achteraf implementeren van CO<sub>2</sub>-afvang (met name voor technologieën, bijvoorbeeld oxyfuel combustion, waarvoor een fundamentele wijziging van industriële processen nodig is);
- Inventarisatie van de mogelijkheden voor de integratie van industriële processen en elektriciteitsopwekkingseenheden voor kosteneffectieve CO<sub>2</sub>-emissiereductie;

- Onderzoek naar het potentieel van CCS voor decentraliseerde emissiebronnen door de resultaten van dit proefschrift te gebruiken in nationale/regionale energie-economische modellen (met name voor de lange termijn);
- Onderzoek naar het effect van CO<sub>2</sub>-afvang op de emissie van luchtverontreinigende stoffen voor technologieën als oxyfuel combustion afvang en post-combustion afvang met behulp van geavanceerde oplosmiddelen en membranen. In aanvulling hierop het in detail modelleren van technologieën met stroomdiagrammen voor de CO<sub>2</sub>-afvangtechnologieën die geïntegreerd zijn in de industriële processen;
- Evaluatie van de economische impact van de uitstoot van vervuilende stoffen, behalve CO<sub>2</sub>, als gevolg van CO<sub>2</sub>-afvang uit industriële processen, met name voor SO<sub>2</sub>- en NH<sub>3</sub>-emissies;
- Onderzoek naar mogelijkheden voor het uitbreiden van het aantal draaiuren per jaar van DH CHP-installaties om daarmee CO<sub>2</sub>-afvang meer kosteneffectief te maken;
- Analyse van mogelijkheden voor het minimaliseren van de CO<sub>2</sub>-compressiekosten- en de lokale CO<sub>2</sub>-vervoerskosten door verschillende systeemconfiguraties te vergelijken.

## 要旨

本章は、本論文の背景、目的および主な研究結果をまとめたものです。

### 1. 背景

地球の温暖化は大気、雪氷圏や海洋の変化の観測結果から明白である。20 世紀後半の世界平均地球温暖化はほぼ間違いなく人為起源の温室効果ガスによるものであることが結論付けられている。二酸化炭素 (CO<sub>2</sub>) は最も重要な人為起源の温室効果ガスである。工業化以後の大気中の CO<sub>2</sub> 濃度上昇は主に化石燃料の使用によるものである。世界の平均気温上昇が 1.5-2.5°C を超えると、生物多様性と生態系に多大な悪影響を及ぼすことが予想されている。

今日までに世界 100 カ国以上が気候変動のリスク、影響、被害を最小限に緩和する指針として地球の平均温度上昇を 2°C に制限することで合意している。この目標を達成するためには様々な気候変動緩和策が必要になる。その中でも CO<sub>2</sub> 回収・地下貯留技術 (CCS) は、引き続き大規模な化石燃料の使用が予想される今後数十年において温室効果ガス排出削減のための重要な選択肢と考えられている。

CCS は二酸化炭素の回収、輸送および地下貯留の三部から構成される。そのうち CO<sub>2</sub> 回収のコストが最も大きいため、コスト削減のための様々な技術の研究・開発が行われている。

CO<sub>2</sub> 回収技術に関しては、これまでは主に大規模発電所からの回収を主眼に研究・開発がなされてきたが、産業部門においても大きな CCS の可能性を秘めている。産業部門からの CO<sub>2</sub> 排出は直接または間接的に世界全体の人為起源の CO<sub>2</sub> 排出の約 40% を占める。また、様々な工業プロセスにおいて CO<sub>2</sub> 回収の費用対効果を高めることができる高濃度の CO<sub>2</sub> を含んだガス流が多く存在する。これまでに様々な CO<sub>2</sub> 回収技術とその経済性について報告されている。

さらに、今日多くの経済圏において分散型発電の普及がみられる。分散型発電の主要なオプションであるコージェネレーション (CHP) の広範囲な展開は化石燃料の消費と CO<sub>2</sub> 排出の大幅削減につながるが、これは同時に中小規模排出源からの CO<sub>2</sub> 排出の増大を意味する。分散型エネルギーシステムからの CO<sub>2</sub> 排出量は、2030 年には現在の世界のエネルギー関連の CO<sub>2</sub> 排出量の約 20% を占めると推定される。将来においては分散型エネルギーシステムにおいても CCS を考慮する必要がある。

しかし、これまで以下の事項については科学的検証がなされていない：

- 工業プロセスからの CO<sub>2</sub> 回収の技術性および経済性の公平な比較・評価。工業プロセスからの CO<sub>2</sub> 回収に関して、近年様々な技術が提案され、それらの技術性・経済性が報告されている。しかし、報告されている技術性・経済性はそれぞれ前提条件が異なるため、結果をより公平に比較・評価するにはシステム境界とパラメータを標準化させなければならない。
- 産業部門からの CO<sub>2</sub> 回収による大気汚染物質排出量の変化。産業部門は長年に渡り NO<sub>x</sub> や SO<sub>2</sub> 等の大気汚染物質の主な排出源となってきた。ゆえに CCS の評価は CO<sub>2</sub> の削減だけでなく、その他の排出物質に与える影響も加味しなければならない。

分散型エネルギーシステムからの CO<sub>2</sub> 回収の技術性・経済性。近年コージェネなど高いエネルギー変換効率を可能にする分散型エネルギーシステムが注目されているが、これらシステムからの CO<sub>2</sub> 回収の技術性・経済性の比較・評価はこれまでされていない。

## 2. 本論文の目的

本論文では、産業部門および分散型エネルギーシステムからの CO<sub>2</sub> 回収の技術的、経済的可能性を検証する。以下3つの研究問題を設定した。(1) 主要産業部門における CO<sub>2</sub> 回収のための技術的および経済的可能性は？ (2) 主要産業部門からの CO<sub>2</sub> 回収は大気汚染物質（窒素酸化物（NO<sub>x</sub>）、二酸化硫黄（SO<sub>2</sub>）、浮遊性粒子状物質（PM）及びアンモニア（NH<sub>3</sub>））の排出量にどのような影響を及ぼすか？ (3) 分散型エネルギーシステムからの CO<sub>2</sub> 回収の技術的および経済的可能性は？

分散型エネルギーシステムに関して本論文では電気容量にして 1-50 MW<sub>e</sub> 程度のシステムを研究の対象とした。また本論文における経済性分析は以下の条件において行われた。天然ガス価格は短期（2020年頃）においては 7.3-8€/GJ<sub>LHV</sub><sup>43</sup>、中長期では 8€/GJ<sub>LHV</sub> を想定している。石炭価格は短中期・長期共に 2.6€/GJ<sub>LHV</sub> を想定している。金利は産業部門においては 10-12%、その他の部門では 10% を想定している。各設備の経済的耐用年数は、石炭火力発電所および石炭ボイラーは 35年、他の設備では 20-25年を想定している。

## 3. 要約および結論

### (1) 主要産業部門における CO<sub>2</sub> 回収のための技術的および経済的可能性は？

主要産業部門における CO<sub>2</sub> 回収のための技術的および経済的可能性は第2章で取り上げられている。鉄鋼、セメント、石油精製および石油化学部門からの CO<sub>2</sub> 回収を評価した。評価は広範な文献レビューに基づいている。技術の公正な比較を可能にするため、重要なパラメータ（プラント稼働率、エネルギー価格、金利、経済的耐用年数、CO<sub>2</sub> 圧縮圧力、使用端 CO<sub>2</sub> 排出原単位）は標準化されている。

鉄鋼部門については主に高炉法と熔融還元法を取り扱った。高炉法に関しては短中期的（2020-2025）には炉頂ガス循環高炉（TGRBF）技術により熱延コイル 1 トン当たり 0.7-0.8 トンの CO<sub>2</sub>（全炭素入力 of 40-45 パーセント）を、CO<sub>2</sub> 抑制 1 トン当たり 40-65 € にて削減できることが分かった。しかし TGRGF 技術は既存の高炉技術に比べてマテリアル・エネルギーフローが大きく変化するため、経済性に関する不確実性も高いことが示された。従来型の高炉ガスからの CO<sub>2</sub> 回収も同様の削減費用（抑制 1 トン当たり 40-65 €）で行えることが示されたが、CO<sub>2</sub> の削減率は熱延コイル 1 トン当たり約 0.3-0.4 トンと低くなる。熔融還元法に関しては COREX 法からの CO<sub>2</sub> 回収は抑制 1 トン当たり 30 € 以下で行えるが、大幅な CO<sub>2</sub> 削減を達成するためには、水性シフト反応により COREX 副生ガスに含まれる一酸化炭素を CO<sub>2</sub> に転換しなければならない。長期的（2030 年以降）には、膜分離や水和物結晶等の技術が提案されているが、従来技術に比べ大きな経済的利点をもたらさない可能性が示唆された（CO<sub>2</sub> 抑制 1 トン当たり 30-55 €）。選択的炭素膜を従来型の高炉ガスに用いた場合、CO<sub>2</sub> 抑制 1 トン当たり 30 € の削減費用になることが示唆された。水和物結晶化技術は低コスト、低炭素の電力が利用可能になった場合に魅力的な選択肢となるかもしれない。

二酸化炭素排出が社会的・経済的に制限された中で製鉄所を新設する場合、COREX 法等の熔融還元技術は低コストの CO<sub>2</sub> 排出削減を可能にするため既存の高炉法に基づいた一貫型製鉄の強力な競争相手になることが示唆される。本研究の結果は CO<sub>2</sub> 回収を行う場合、COREX 法は高炉法に比べて低コスト・低 CO<sub>2</sub> 原単位となることを示唆している。さらに、

<sup>43</sup> LHV: 真発熱量



長期的には溶融還元技術は効率的な製鉄の一部として CO<sub>2</sub> 分離を組み込むことにより更なる低コスト・低 CO<sub>2</sub> 原単位を実現できることが示唆された。従来型の一貫型製鉄が主流でありつづけることが予想されるものの、大幅な CO<sub>2</sub> 排出削減が求められる場合は鉄鋼部門が従来型の一貫型製鉄から溶融還元法へ中長期的に移行するかもしれない。

セメント部門では、短中期的には化学吸収法が唯一実現可能な技術になると考えられる。化学溶媒の再生エネルギーを最も経済的に供給するには隣接する発電所から低位熱を購入するのが理想的である。この方法では CO<sub>2</sub> 排出を 1 トン当たり約 70 € にて約 60 % 削減することが示された。上記以外の方法で溶媒再生熱を供給する場合 CO<sub>2</sub> 削減費用は 1 トン当たり 90-130 € まで上昇することが示された。また、KS-1 等の最先端の商用化学溶媒と発電所からの低位熱を用いた場合 CO<sub>2</sub> 削減費用は 1 トン当たり 50 € 以下になる可能性が示された。

長期的には CO<sub>2</sub> 削減費用を 1 トン当たり 25-55 € 程度にできることが示された。クリンカ焼成キルンにおける燃料の純酸素燃焼によって CO<sub>2</sub> 削減費用を 1 トン当たり 40 € まで削減することができることが分かった。CaO-ループ法を用いる場合では更なる低コスト (CO<sub>2</sub> 削減費用 1 トン当たり 25-30 €) の実現が示唆された。化学吸収法も長期的には 1 トン当たり 35-55 € の CO<sub>2</sub> 削減費用までコストを低減できることが分かった。セメントプラントからの CO<sub>2</sub> 回収はマテリアルフローに影響を与えないことが示唆されるゆえ、CO<sub>2</sub> 削減費用の不確実性は鉄鋼部門に比べて小さいことも示された。

石油精製および石油化学部門においては、短中期的にはボイラー・ヒーターおよび流動接触分解装置 (FCC) の両方において純酸素燃焼回収 (CO<sub>2</sub> 削減費用 1 トン当たり 50-60 €) が化学吸収 (CO<sub>2</sub> 削減費用 1 トン当たり >70 €) より経済的であることが分かった。しかし、化学吸収法のコストは使用される蒸気のコスト次第である。また、既存の設備を改造して純酸素燃焼回収を行う場合は改造中の工業生産の損失が大きくなる可能性がある。化学吸収法については最先端の化学溶媒を用いることにより CO<sub>2</sub> 削減費用を大きく低減させられることが示唆される。長期的には酸素分離膜 (OCM) を用いた純酸素燃焼 CO<sub>2</sub> 回収が最も経済的であることが示唆された (削減費用 1 トン当たり 30 € 以下)。しかし、この方法では酸素分離後の高温・高圧の空気を有効利用するために大量の余剰電力が生産される。よってこの CO<sub>2</sub> 回収法の経済性は電力市場の状況 (発電コストおよび使用端 CO<sub>2</sub> 排出原単位) に大きく依存することが示唆された。

産業部門からの CO<sub>2</sub> 回収については発電部門同様、経済性で明らかに勝る技術はないことが示唆された。また、工業プラントからの CO<sub>2</sub> 回収に消費されるエネルギーを効率よく供給する場合に多大な余剰電力が生産されることが予想され、その際電力市場に多大な影響を与える可能性が示唆された。さらに、長期的には CO<sub>2</sub> 回収による工業プロセスの効率化が産業部門における CCS の大規模な展開の鍵を握ると考えられる。

(2) CO<sub>2</sub> 回収は主要産業部門からの大気汚染物質 (窒素酸化物 (NO<sub>x</sub>)、二酸化硫黄 (SO<sub>2</sub>)、浮遊粒子状物質 (PM) 及びアンモニア (NH<sub>3</sub>)) の排出量にどのような影響を及ぼすか?

第 3 章では近未来 (2020 年頃) に EU-27 の主要産業部門 (セメント、石油精製および鉄鋼) において CO<sub>2</sub> 回収が展開された場合、大気汚染物質の排出量がどのような変化するかを検証した。CO<sub>2</sub> 回収技術はセメント部門と石油精製部門には化学吸収法を、鉄鋼部門には炉頂ガス循環高炉 (TGRBF) を採用した。化学吸収法では異なる熱エネルギー供給オプションを検討した。主な違いは (1) 熱供給プラントの燃料 (石炭または天然ガス) と (2) 熱供給システム (CHP) により生成された CO<sub>2</sub> の取り扱い (放出または回収) である。

本研究では CO<sub>2</sub> 回収が SO<sub>2</sub> と NH<sub>3</sub> の排出に多大な影響を与える可能性があることが明らかになった。全三産業部門全てのプラントにて CO<sub>2</sub> 回収が行われた場合、EU-27 における産業

部門からの SO<sub>2</sub> 排出は最大 80%減少することが示唆された。この減少のうち約半分は工業プラントからの排出の減少、残りの半分は CO<sub>2</sub> 回収に伴う余剰電力生産による電力部門からの SO<sub>2</sub> 排出の抑制によることも分かった。一方、NH<sub>3</sub> の排出は 800%近く増加することが示唆された。これは EU-27 からの全排出量の約 10%に相当する。NO<sub>x</sub> の排出に関しては、セメント部門からの CO<sub>2</sub> 回収に石炭燃料の熱供給プラント (CHP) が使用された場合に約 20%増加することが示唆された。これら大気汚染物質の排出量の変化はセメント部門および石油精製部門に依るものであり、鉄鋼部門からの CO<sub>2</sub> 回収による影響は限定的である。

部門別では、セメント部門からの NO<sub>x</sub> 排出の変化は産業部門全排出量に比べ 5%の減少から 12%の増加の範囲に及ぶことが示唆された。この大きな幅は化学吸収剤の再生エネルギー供給オプションに依る。SO<sub>2</sub> 排出の減少量は産業部門の全排出量の 20-50%程度であることが示された。石油精製部門においては SO<sub>2</sub> 排出量の増加は産業部門全排出量の 5%程度であることが示された。PM に関してはセメント・石油精製両部門からの排出の変化は限定的であることも分かった。さらに、化学吸収法が採用されたこの二部門については、余剰電力生産による電力部門からの SO<sub>2</sub> および PM 排出の抑制の影響が大きいことも示された。ただし、第 3 章で得られた結果は特定の CO<sub>2</sub> 回収技術を想定したものであり、他の CO<sub>2</sub> 回収技術を採用した場合には結果は必ずしも有効ではない。

### (3) 分散型エネルギーシステムからの CO<sub>2</sub> 回収の技術的および経済的可能性は？

この研究問題は、第 4-6 章で検討されている。第 6 章では広範な文献レビューに基づき様々な CO<sub>2</sub> 回収技術の技術性・経済性について発電技術、用途、プラントの大きさ、稼動パターン、CO<sub>2</sub> 回収技術が商業化される時期等の要素を考慮に入れた上で比較・評価を行った。第 4、5 章は分散型エネルギーシステムに関する特定のケースを扱った。スケールメリットの影響を評価するため、本研究で得られた結果は大型発電所の技術性・経済性と比較された。本研究では CHP、ボイラーおよび分散型水素プラントを考慮した。

### 既存の技術・短中期的評価 (2020-2025)

短中期に商業化されると思われる CO<sub>2</sub> 回収技術に関しては、システムエネルギー効率の低減 ( $\Delta\eta$ ) は燃料、エネルギーシステムの種類および用途によって大きく異なる。天然ガス燃料の場合、ボイラーは約 28 %、コンバインドサイクル (NGCC) では熱統合が最適された場合 11 % を示した。石炭燃料の場合、 $\Delta\eta$  はボイラーでは約 33 %、発電所の場合は 23-28 % を示した。

分散型エネルギーシステムからの CO<sub>2</sub> 回収は熱統合により  $\Delta\eta$  を格段に低下させることができることが示唆された。工業用 CHP については部分負荷にて低効率で稼動している場合が多く、余剰発電・蒸気生産能力を CO<sub>2</sub> 回収エネルギー供給に用いればプラントの総効率を上昇させることができるため、正味の  $\Delta\eta$  を低減させることができる。第 4 章では工業用 NGCC-CHP にて CO<sub>2</sub> の化学吸収を行う場合、400 MW<sub>e</sub> の NGCC 発電所からの化学吸収より  $\Delta\eta$  が既存の技術 (短期) を用いた場合 16 %、現在開発中の技術 (中期) を用いた場合 12 % 低くなることが示された。

地域暖房用コージェネ (DH-CHP) の場合、CO<sub>2</sub> 回収エネルギーの一部を低位熱として回収できる。文献を検証した結果、化学吸収法の場合は消費される熱エネルギーの約 30 % が追加の設備投資なく回収できることが示唆された。純酸素燃焼の場合も、酸素分離と CO<sub>2</sub> 圧縮に消費される電気エネルギーの 30 % を低位熱として回収できることが示唆された。また純酸素燃焼の場合、熱統合により設備費も約 5 % 削減できることが示唆された。

短期的な CO<sub>2</sub>削減費用は 100MW<sub>LHV</sub> 以下の場合 1 トン当り 30-140 €、10-100MW<sub>LHV</sub> の場合 1 トン当り 50-150 € になることが示唆された。数多くの提案されている技術のうち、以下のものが他より経済性に優れていることが示唆された(1) 工業用 NGCC-CHP から化学吸収(低負荷・低熱効率の場合)、(2) CHP から純酸素燃焼回収、(3) 石油精製・副生ガスの純酸素燃焼。スケールメリットの影響はエネルギーシステムおよび CO<sub>2</sub> 回収技術の種類に大きく依存する。第 6 章にその例が示されている。500 MW<sub>e</sub> の大型発電所の場合、CO<sub>2</sub> 削減費用は化学吸収法で 1 トン当り 39 €、純酸素燃焼法で 1 トン当り 33 € と算出された。しかしながら 50MW<sub>LHV</sub> 規模の CHP に小規模化した場合、CO<sub>2</sub> 削減費用は 1 トン当り 79 € と 58 € となり、大きな差が示された。

前述の通り、CHP と CO<sub>2</sub> 回収技術の熱統合の効果は大きいものの、本研究の結果はこの効果が工業用 CHP の場合にのみ有効であることが示唆された。既存の化学吸収技術を用いた場合(短中期)、100 MW<sub>e</sub> 規模の工業用 NGCC-CHP からの CO<sub>2</sub> 削減費用(1 トン当り 41-44 €) は 400 MW<sub>e</sub> の大型発電所からのそれと同等であることが示唆された。また現在開発中の化学吸収技術を用いた場合(長期)でも、200 MW<sub>e</sub> 規模の工業用 NGCC-CHP からの CO<sub>2</sub> 削減費用(1 トン当り 41-44 €) は 400 MW<sub>e</sub> の大型発電所からのそれと同等であることが示唆された。一方、DH-CHP の場合、熱統合により CO<sub>2</sub> 削減費用を 1 トン当り 10 € ほど低減できる。しかし低い稼働率により CO<sub>2</sub> 抑制 1 トン当りの設備費減価償却が大きくなるため、その効果は相殺されてしまう。

CHP の熱・電力供給比が高い場合、化学吸収法は熱エネルギーを巡る競争を引き起こすので推奨できない。熱エネルギーの競争が起きた場合、以下の 2 通りの方法で熱需要を満たすことが想定される:(1) 予備のボイラーを用いる、(2) より大きな CHP の設置。(1) の方法では上質の燃料で低位熱のみを生産することになるため、エネルギー変換効率が悪く  $\Delta\eta$  が大きくなってしまいうため推奨できない。これは第 4 章で工業用 NGCC-CHP について検証されている。(2) の方法では多大な余剰電力を生産することになり、電力輸出の大きなインセンティブがなければならない。また分散型電源は出力レベルに技術的限界があるため、この方法を採用できるケースは限定されるかもしれない。

### 先端技術・長期的評価(2030 年以降)

長期的に商業化されるであろう先端技術に関しては  $\Delta\eta$  は 4-9 % 程度になることが示された(第 6 章)。経済性については第 6 章にて、CO<sub>2</sub> 削減費用は 100 MW<sub>LHV</sub> 以下の場合 1 トン当り 10-90 €、10-100 MW<sub>LHV</sub> の場合 1 トン当り 25-100 €、10 MW<sub>LHV</sub> 以下の場合 1 トン当り 35-150 € 程度になると結論づけられた。固形酸化物形燃料電池(SOFC)や水素分離型リフォーマー(water gas shift membrane reformer: WGSMR)などの最先端技術を用いた場合、エネルギー効率と同時に設備費も高くなることが想定されるため、プラントの稼働時間が経済性に大きく影響することが示唆された。これら最先端技術の多くはモジュール化されたものが多いため、設備を小型化した場合の費用対効果への影響は限定的である。

SOFC 等の最先端エネルギー変換技術は CO<sub>2</sub> 削減費用を大幅に低減させられる可能性がある一方、コストや信頼性の面で既存の技術に劣っている。ゆえにこれら最先端エネルギー変換技術は CO<sub>2</sub> 回収なしには既存技術と経済的に競争できない可能性がある。そこで第 5 章では、CO<sub>2</sub> 回収が SOFC の商業化初期の競争力をどの程度強化できるかを検証した。本研究では 5 MW<sub>e</sub> の工業用 SOFC-CHP を取り扱った。CO<sub>2</sub> 回収には燃料極のオフガスの純酸素燃焼を、酸素分離には圧力・真空スイング吸着法(PSA/VSA)を採用した。経済性の分析では SOFC が既

存の CHP 技術（ガスタービン）との競合に必要な電池スタック生産コストを比較した。CO<sub>2</sub> 価格がゼロの場合、電池スタック生産コストは 140 \$/kW 以下でなければガスタービン CHP と競合できないことが分かった。この値は想定した SOFC の形状（flat-tubular）にとっては非常に達成困難なコストレベルである。そして SOFC-CHP は CO<sub>2</sub> 価格が 1 トン当り 43 \$ 以上の場合は CO<sub>2</sub> を回収した方が経済性がよくなり、また CO<sub>2</sub> 価格が高くなるほど CO<sub>2</sub> 回収 SOFC-CHP の優位性は高まることが分かった。CO<sub>2</sub> 価格が 1 トン当り 100 \$ の場合、SOFC がガスタービンと競合するには CO<sub>2</sub> 回収なしでは電池スタック生産コストを 350 \$/kW 以下に抑えなければならないが、CO<sub>2</sub> を回収する場合は大幅に高い 930 \$/kW でも競争力を維持できることが分かった。

### 小規模 CO<sub>2</sub> 回収の障害：CO<sub>2</sub> 圧縮と末端輸送

小規模 CO<sub>2</sub> 回収の場合、CO<sub>2</sub> の圧縮と回収地から幹線パイプラインまでの末端輸送のコストが非常に高くなる可能性がある。本研究では末端輸送距離を 30 km と仮定した場合、圧縮・末端輸送コストは一日 500 トンの回収規模では 1 トン当り 10 €、一日 5 トンの回収規模（天然ガス 1 MW<sub>LHV</sub> に相当）では 1 トン当り 50 € にも上ることが示された。このコストは主に CO<sub>2</sub> 圧縮によるものであることも同時に示された。

第 6 章で得られた結果は、分散型エネルギーシステムからの CO<sub>2</sub> 回収コストは分散型電源の設備規模がある程度低い範囲（数 MW<sub>e</sub> 規模）においても将来想定される CO<sub>2</sub> 価格と比べて法外に高いわけではないことを示している。CO<sub>2</sub> 輸送および貯留のインフラが整備され次第、大型排出源だけでなく分散型エネルギーシステムからの CO<sub>2</sub> 回収も短中期および長期において考慮されるべきであることを示唆している。

## 4. 最後に

本論文では、産業部門と分散型エネルギーシステムからの CO<sub>2</sub> 回収の技術性および経済性を比較・検討し、いくつかの有望な技術が同定された。また本研究にて得られた結果は短中期および長期両方において CO<sub>2</sub> の輸送・貯留のインフラが整備され次第、全ての工業プラントおよび分散型エネルギーシステムにおいて CO<sub>2</sub> 回収の実現性を考慮されるべきであることを強く示唆している。同時に、工業プラントおよび分散型エネルギープラントからの CO<sub>2</sub> 回収の実現性は大型発電所からのそれに比べ、よりケース・バイ・ケースであることも示唆された。

本論文では各章の終わりに更なる研究のための提言をしてきた。それらをまとめて以下に示す。

- 工業プラントおよび分散型エネルギーシステムにおける CCS の実現性のケーススタディ
- 工業プロセスの核心部分を改造する必要がある CO<sub>2</sub> 回収技術（純酸素燃焼など）を採用した場合の工業生産ロスの分析
- CO<sub>2</sub> 排出削減の費用対効果の向上のための工業プロセスや発電所の統合の可能性の検討と評価
- エネルギー・経済モデル（MARKAL 等）を用いた特定の国・地域における分散型エネルギーシステムからの CO<sub>2</sub> 回収・地下貯留の可能性の評価
- 本研究で検討された技術以外の CO<sub>2</sub> 回収技術を採用した場合の工業プラントからの大気汚染物質排出への影響および詳細なプロセスモデリング

- CO<sub>2</sub>回収による工業プラントからの SO<sub>2</sub> と NH<sub>3</sub> 排出の変化による、経済的影響の評価
- 地域暖房コージェネシステムから経済的に CO<sub>2</sub>を回収するために年間稼働時間を延長する可能性の模索
- 分散型 CO<sub>2</sub> 排出源からの CO<sub>2</sub>回収における、CO<sub>2</sub>の圧縮と末端の輸送コストの最小化させるための分析



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## Curriculum Vitae

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Takeshi Kuramochi (倉持 壯) was born on July 8, 1981 in Kanagawa, Japan. He spent most of his childhood in Tokyo. He entered the University of Tokyo (Science II) in April 2000, where he majored in Chemistry and Biotechnology in the School of Engineering. He received a Bachelor of Engineering degree in March 2004.



In September 2004, he moved to the Netherlands to study in the Master's programme Sustainable Development (Track Energy and Resources) at Utrecht University. He received a Master of Science degree in 2006. His MSc thesis research was on the sector-based approach for greenhouse gas emissions reduction for the post-Kyoto protocol period under the supervision of Dr. Michel den Elzen (Netherlands Environmental Assessment Agency) and Dr. Jeroen van der Sluijs (Utrecht University).

Since December 2006, he works as a researcher in Group Science Technology and Society in the Copernicus Institute, Faculty of Science at Utrecht University. Since then, he has worked in the following projects: the Dutch research programme on CO<sub>2</sub> capture technology development (CAPTECH), the Dutch policy research programme on air and climate (BOLK), and the Dutch national research programme on CO<sub>2</sub> Capture and Storage (CATO). Part of the results from these projects are presented in this dissertation.