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Techno-economic performance and spatial footprint of infrastructure configurations for large scale CO_2 capture in industrial zones A case study for the Rotterdam Botlek area (part A)

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ABSTRACT

This study developed a method to assess the techno-economic performance and spatial footprint of CO_2 capture infrastructure configurations in industrial zones. The method has been successfully applied to a cluster of sixteen industrial plants in the Dutch industrial Botlek area (7.1 $MtCO_2$/y) for 2020–2030. The configurations differ *inter alia* regarding capture technology (post-, pre-, oxyfuel combustion) and location of capture components (centralized vs. plant site). Results indicate that oxyfuel combustion with centralized oxygen production and decentralized CO_2 compression is the most cost effective and realistic configuration when applying CO_2 capture to all industrial plants (61€/tCO_2; 5.8 $MtCO_2$/y avoided), mainly due to relatively low energy costs compared to post- and pre-combustion. However, oxyfuel combustion at plant level is economically preferable when capturing CO_2 from only the three largest industrial plants. For post-combustion, a separated absorber-stripper configuration (73€/tCO_2; 7.1 $MtCO_2$/y avoided) is preferable from a cost perspective, due to economic scale effects of capture equipment. The optimal pre-combustion configuration shows a slightly less favorable performance (81€/tCO_2; 4.4 $MtCO_2$/y avoided). Whereas many industrial plants have insufficient space available for capture equipment, centralized/hybrid configurations show no insurmountable space issues. The deployment of the most favorable configurations is addressed in Part B.

1. Introduction

Nowadays, industry and petroleum refineries are responsible for nearly 40% of the global energy demand and around one-third of the worldwide anthropogenic CO_2 emissions (IEA, 2014a,b). Industrial energy consumption and CO_2 emissions are projected to rise further in the coming decades (IEA, 2014a). Given these projections, decisive action is required to curb anthropogenic greenhouse gas (GHG) emissions in order to achieve the stabilization targets of 450 ppm(v) (IPCC, 2014; ZEP, 2013). Despite the need to implement energy efficiency measures and lower the CO_2 intensity of industrial processes, the International Energy Agency (IEA, 2014a,b) considers the deployment of carbon capture and storage (CCS) in the industry critical to reach these ambitious climate targets.

The potential for and economic costs of CO_2 capture in industry have been investigated in various studies at the broad industrial (e.g., Damen et al., 2009; IEA, 2014a; Saygin et al., 2013), sectoral (e.g., IEA GHG, 2008a; Oda et al., 2007) and industrial plant level (Allam et al., 2005a,b; Berghout et al., 2015, 2013; IEA GHG, 2000, 2008a; Johansson et al., 2013, 2012; Van Straelen et al., 2010; Wilkinson et al., 2003). An extensive literature review on the techno-economic performance of CO_2 capture technologies in industry has shown that the short and midterm (coming 10–15 years) CO_2 avoidance costs range from 25 to 65€$_{2007}$/tCO_2 for the iron and steel sector, 65–135€$_{2007}$/tCO_2 for the cement sector, and

Abbreviations: ADIP-X, mixture of methyldiethanolamine, piperazine and water; ASU, air separation unit; ATR, autothermal reforming; CC, carbon capture; CCS, carbon capture and storage; CHP, combined heat and power; DCC, direct contact cooling; DCCI, downstream capital costs index; DPC, drying purification cooling; EI, electricity import; FGD, flue gas desulphurization; GHG, greenhouse gas; GT, gas turbine; HP, high-pressure; IEA, international energy agency; IPCC, intergovernmental panel on climate change; LHV, lower heating value; MEA, monoethanolamine; MP, medium-pressure; NGCC, natural gas combined cycle; PC, pulverized coal; PPC, process plant cost; ppm(v), parts per million (by volume); PSA, pressure swing adsorption; RAP, Rotterdam aromatics plant; Recsor, remote central solvent regeneration; SCR, selective catalytic reduction; SER, specific energy requiremen; SR, steam reformer; t, metric tonne; TCR, total capital requiremen; TPC, total plant cost; WH, waste heat; WGS, water–gas shift.

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50–120€$_{2007}$/tCO$_2$ for petroleum refineries and petrochemicals, depending on various factors such as capture technology, energy supply options, energy prices and industrial plant configuration (Kuramochi et al., 2012). The roadmaps of the UNIDO/IEA (2011) and the Zero Emissions Platform (2013) state that reducing avoidance costs of CCS in industry is vital to reach the ambitious climate goals. Clustering multiple industrial plants is a possibility to curtail capture and transport costs by exploiting economies of scale compared to individual plant chains, especially for smaller CO$_2$ emitters (Bureau et al., 2011; RCI, 2009). The Green Alliance estimated that the unitary CCS costs of the White Rose demonstration plant in the UK would come down by nearly two-thirds by engaging in a joint cluster approach with industrial CO$_2$ emitters in the Tees Valley (UK) (Benton, 2015).

To date, several studies undertook a feasibility study to assess the economic and practical viability of a distributed CO$_2$ collection and transmission network in an industrial region: the Merseyside and Deeside Basin (UK) (IEA GHG, 2007a), Yorkshire and Humber region (UK) (Yorkshire Forward, 2008), Tees valley (UK) (AMEC, 2010) and Le Havre (France) (Roussanaly et al., 2013). The AMEC study also presented the CO$_2$ capture costs, physical footprint of capture equipment, and land availability on the industrial sites. However, aggregated data from merely one source was used for the input parameters. Also, the study's economic input parameters were rather generic and the underlying cost model was not presented. Nørstebø et al. (2012) and Midthun et al. (2012) established a decision support model for investment in a small Norwegian industrial park, thereby putting strong emphasis on the optimization of post-combustion CO$_2$ capture plants in terms of investment profitability, size and how it should be operated during the life time. Bureau et al. (2011) examined the techno-economic feasibility of innovative amine-solvent based post-combustion capture configurations for the French industrial area Le Havre, including a variation with a flue gas collection network and a separated absorber-stripper configuration with an amine solution circulation. From this specific case, they found *inter alia* that: (i) for several point sources with individual emissions smaller than 500 ktCO$_2$/y, pooling flue gases is more competitive than a standalone capture unit per emission point; (ii) pooling flue gases is an interesting strategy to increase the CO$_2$ volume and/or CO$_2$ concentration in flue gases; (iii) in the case of large CO$_2$ volumes and/or flue gases with low CO$_2$ concentrations, pooling CO$_2$-rich amine solutions is more economic than pooling flue gases. However, as the interest of pooling depends strongly on the capture costs considered and the selected capture process (Bureau et al., 2011), more research is required. Up to this point, an in-depth analysis into the feasibility of different CO$_2$ capture infrastructure configurations for the three main CO$_2$ capture technologies has not been conducted. Aside from the techno-economic aspect, concerns have been expressed regarding possible space limitations for CO$_2$ capture equipment on industrial sites (e.g., Berghout et al., 2013; Hurst and Walker, 2005; Van Straelen et al., 2010). To date, only few studies looked into the quantitative space requirement for CO$_2$ capture retrofit (e.g., GCCSI, 2010; IEA GHG, 2006, 2005a,b; Sinclair Knight Merz, 2009). Florin and Fennell (2009) found a large variation in the required space reported in the literature, which is largely due to assumptions made in each study mainly on plant capacity (net or gross capacity) and the number of utility and/or capture trains. As physical footprints are very case-specific, physical footprints should be based on detailed process simulation and engineering studies. More research is needed to assess potential space limitations on the industrial cluster level. Moreover, a method is needed to assess the techno-economic performance and spatial footprint of CO$_2$ large scale capture infrastructure configurations in industrial zones in a consistent manner.

The objective of this study is to develop this method based on bottom-up analysis and to illustrate this method by investigating different infrastructure configurations for an industrial cluster for the period 2020–2030. The studied configurations differ with respect to the CO$_2$ capture technology (post-, pre-, or oxyfuel combustion), location of the capture components (central, semi-central, or industrial plant site), local pipeline network, and energy supply alternatives for the CO$_2$ capture process (combined heat and power (CHP) plant, natural gas combined cycle (NGCC), gas-fired boiler, electricity import from the grid). The spatial footprint of the CO$_2$ capture infrastructure configurations will be taken into account as well. The time frame of this study is the period 2020–2030. The industrial Botlek area in the Port of Rotterdam in the Netherlands was selected as a case study, because of its high concentration of CO$_2$ point sources representing various industrial sectors, and a wide variety of small and large CO$_2$ emitters, which are dispersed across an area with limited space availability. Furthermore, the port has been appointed by the local government as a potential region for large scale CO$_2$ capture (RCI, 2009). This study did not include the costs for the trunk CO$_2$ pipeline that is planned by the Rotterdam Climate Initiative to run through the Botlek in the future (RCI, 2011). CO$_2$ storage was also excluded from the analysis.

This is the first in a series of two papers investigating the optimal deployment of CCS in industrial zones. While part A focuses on assessing CO$_2$ capture infrastructure configurations across a geographical area, part B centers around their build out over time by investigating deployment pathways, which differ regarding the CCS build out sequence, number of deployment steps, and whether CO$_2$ capture units and energy plants are oversized, or not.

The present paper is structured as follows. Section 2 describes the case study, method and key input data used. Section 3 presents the techno-economic performance and spatial footprints of the CO$_2$ capture infrastructure configurations. A discussion on the method, results and uncertainties is laid out in Section 4. Finally, main conclusions are drawn in Section 5.

2. Method and data

2.1. Case study

The Botlek can be considered as a separate, demarcated area within the port of Rotterdam since it is enclosed by water and it accommodates over thirty companies of which fifteen are currently obliged to compound and update so-called monitoring plans because their annual CO$_2$ emissions are higher than 25 ktCO$_2$/y. The waste processing plant, which also has high CO$_2$ emissions, is exempted from the monitoring obligation because emissions from waste processing are regarded as climate neutral. This study examines sixteen CO$_2$ emitters including the waste processing plant. Fig. 1 and Table 1 present data of their size, locations and potential space for CO$_2$ capture equipment required for clustering multiple industrial plants. The current mass and energy flows of the industrial plants were assumed to remain constant in the short term (2020–2030). Most of the limited space available in the Botlek is leased by either the municipality or the private companies, which have the first right to claim this land and build new installations. These parcels are therefore by no means readily available for CO$_2$ capture installations. However, experience with other regional projects (e.g., the local steam pipe project) has shown that some sort of arrangements can be made between industrial operators and the local authority as long as pipelines or capture installations do not interfere with their core business (Hurenkamp, 2011). In this study, it is assumed that the plot spaces are available for capture equipment and pipelines. The NGCC, CHP plants and hydrogen plants in the Botlek area will hereafter be referred to as *in situ* technology to

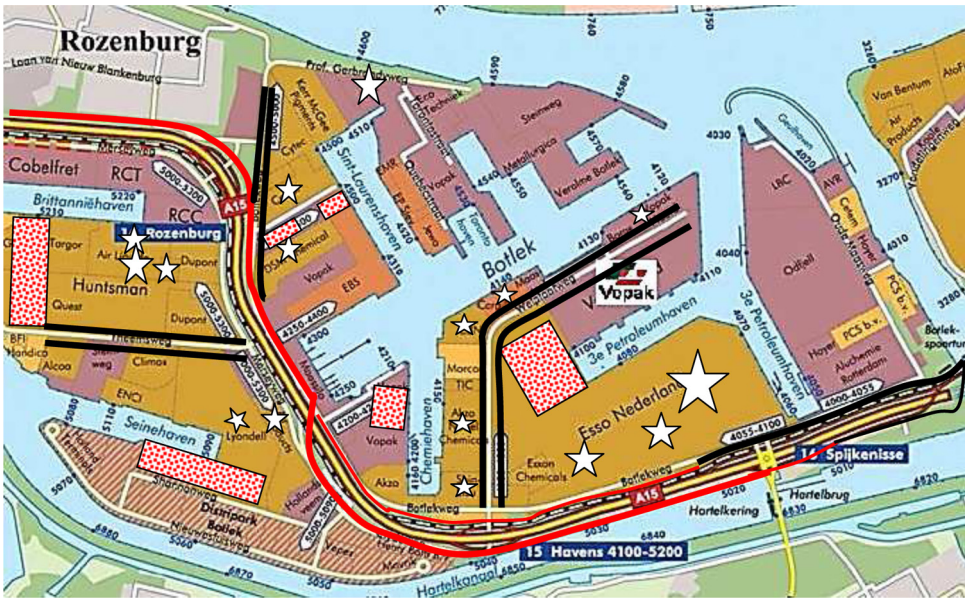

Fig. 1. Botlek area with industrial plants emitting CO_2 (white stars). The size of the white stars reflects the annual amount of CO_2 emissions. The red-dotted rectangulars denote the possible space available for CO_2 capture installations and additional energy plants. A simplistic overview of the pipeline strips is indicated by the red and black lines; the black lines indicate the pipeline strips in which (practically) no space is available for new (large diameter) pipelines, while the red lines indicate the pipeline strips in which space is still available (Pipeliner, 2012a,b,c). The high pressure trunk CO_2 pipeline planned by the Rotterdam Climate Initiative would roughly follow the track of the black–white line on the map. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

Table 1
Point sources in the Botlek, their respective annual CO_2 emissions (both from fuel combustion and core processes) and available space for CO_2 capture equipment required for clustering multiple industrial plants (Based on aerial photographs of Google Maps (2011).

Plant name	Plant type	CO_2 (kt/y)	Plant name	Plant type	CO_2 (kt/y)
Esso	Refinery	2,200	Akzo Nobel	Chemical	181
AVR Rijnmond	Waste processing	1,760	Lyondell	Chemical	133
Air Products (new plant)	Industrial gases	800	DSM	Chemical	101
Eurogen	Utility[a]	465	Shin Etsu	Chemical	80
Rotterdam Aromatics Plant (RAP)	Chemical	411	Evonik Carbon Black	Chemical	61
Air Products (old plant)	Industrial gases	403	Air Liquide	Industrial gases	53
Cabot	Chemical	228	Cargill	Chemical	26
Enecal energy	Utility[b]	204	Biopetrol	Biofuels	18

Total CO_2 emissions 7.1 Mt

Description of location	Amount of space ($\cdot 10^3$ m^2)	Description of location	Amount of space ($\cdot 10^3$ m^2)
Esso refinery site	~200	Distripark Botlek site	~90
Broekman Distriport site	~200	Botlekstraat plot 1	~30
Vopak site	~55	Botlekstraat plot 2	~25

Sources: (Air Liquide, 2010; Air Products, 2010; Akzo Nobel, 2010; AVR, 2010; Biopetrol, 2010; Cabot, 2010; Cargill, 2010; DSM, 2010; Enecal, 2010; Esso, 2010; Eurogen, 2010; Evonik Carbon Black, 2010; Lyondell, 2010; RAP, 2010; Shin Etsu, 2011).

[a] The emission sources are two in situ simple cycle gas turbine CHP plants, mainly fuelled on natural gas.

[b] The emission source is one in situ gas turbine CHP plant that is fueled entirely on natural gas.

distinguish them from newly built NGCC, CHP and hydrogen plants that are required for the CO_2 capture infrastructure configurations.

2.2. CO_2 capture infrastructure configurations

A set of possible CO_2 capture infrastructure configurations was identified based on literature, brainstorm sessions involving the authors and interviews with two CO_2 capture experts. Three main CO_2 capture routes were selected: post-combustion capture based on chemical absorption using an aqueous solvent with a mass fraction of 30% monoethanolamine (MEA), oxyfuel combustion, i.e., fossil fuel combustion with cryogenically produced oxygen, and pre-combustion capture, which is based on the conversion of natural gas into hydrogen and CO_2 in a steam reformer (SR) (see Sections 2.2.1–2.2.3). Chemical absorption using MEA, cryogenic oxygen production and steam reforming are commercial, long-established technologies (Berghout et al., 2013; Kuramochi et al., 2012). Furthermore, two different energy supply options were selected: a boiler/El case and a CHP/NGCC case. The boiler/El case assumes additional heat to be produced in a newly-installed boiler, and/or electricity to be purchased from the grid. In the CHP/NGCC case, additional electricity (and heat) is produced in a newly-installed NGCC(-CHP) plant. The CHP plant is dimensioned to deliver the steam demand required for the CO_2 capture process (post- and pre-combustion configuration); excess electricity is sold to the grid. Two sub-cases were investigated: with CO_2 capture (CC) and without CO_2 capture (vent) from the energy plant. The CO_2 capture infrastructure configurations were distinguished by varying the capture technology and locations of the CO_2 capture units and energy plants. The capture equipment and energy plants were either placed at the industrial plant sites (decentral location), at a central spot, or at several semi-central locations. As a consequence, the capture and energy units vary in scale: smaller scales at the industrial plant sites or larger at central locations where flows from

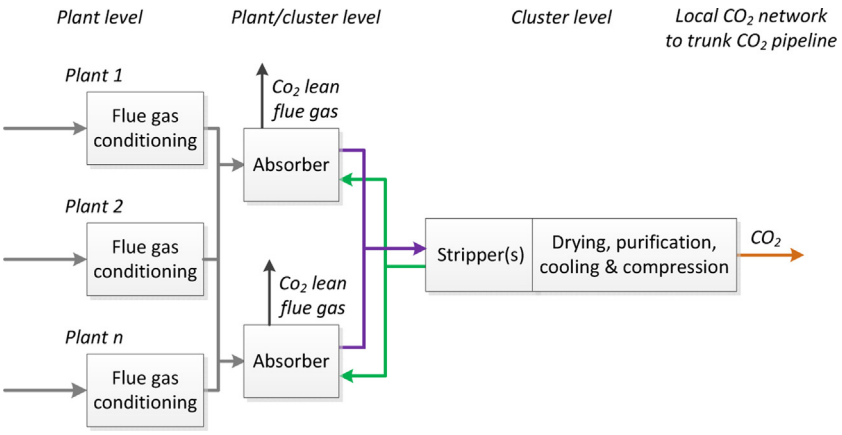

Fig. 2. Schematic overview of the Post-Recsor configuration with a separated absorption and stripping section. The gray, black, purple, green and orange arrows denote the CO_2-rich flue gas, CO_2-lean flue gas, CO_2-rich amine solution, CO_2-lean amine solution, and pure CO_2 flows, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

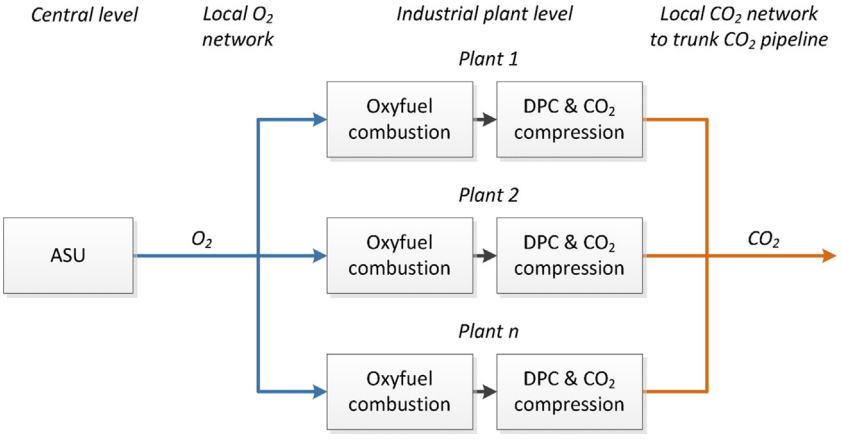

Fig. 3. Schematic overview of the Oxy-Hybrid configuration with centralized oxygen production, and CO_2 drying, purification, cooling and compression at industrial plant level. The blue and orange arrows denote the oxygen and the CO_2 flows, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

different industrial sites are jointly treated. In the central configurations, the capture equipment and energy plants were placed on the available space lying idle north–west of the Esso refinery (see Fig. 1). The configurations are described in further detail for each capture technology.

2.2.1. Post-combustion configurations

Three main post-combustion configurations were investigated: a decentral configuration with all capture units at the individual plant sites (Post-Decentral), a central configuration with all units at one central location (Post-Central), and a configuration in which the flue gas conditioning and absorption takes place at the industrial plant/semi-central level, and the regeneration, treatment (drying, purification, cooling (DPC)) and compression at semi-central level (Post-Recsor[1]) (see Fig. 2). Configurations with compressed flue gas transport have not been investigated as earlier research found that these configurations are uneconomic (Bureau et al., 2011). In each configuration (incl. Post-Central), the flue gas conditioning units (flue gas desulphurization (FGD), selective catalytic reduction (SCR), and direct contact cooling (DCC)) are installed side by side at plant level to minimize flue gas ducting[2] corrosion problems. Similarly, the stripper, DPC units and compressors are jointly installed at the same location for all configurations to avoid the transport of wet, unpurified CO_2 gas given the corrosive impact of sulfur, water and other contaminants on the pumps and pipelines. Each industrial plant was considered as one point source with one volumetric CO_2 concentration flue gas stream. The latter was derived by taking the average CO_2 concentration of all flue gas streams. For Post-Recsor, however, a distinction was made between different flue gas streams on the industrial plant sites (see Fig. 8).

2.2.2. Oxyfuel combustion configurations

Three main oxyfuel combustion configurations were distinguished by having the Air Separation Unit (ASU) for oxygen production and CO_2 compressors: at plant level (Oxy-Decentral), the ASU central and the CO_2 compression step decentral (Oxy-Hybrid) (see Fig. 3), the ASU and CO_2 compression step central (Oxy-Central). In the Oxy-Central configurations, atmospheric pressure (AP) CO_2 gas is transported to a central location using blowers. For all oxyfuel combustion configurations, DPC takes place at the plant level to prevent the transport of wet flue gas. Several adjustments and additional equipment (e.g., piping, ducting, flue gas recirculation fan) are needed to convert the boilers and furnaces to oxyfuel mode (Berghout et al., 2013; Wilkinson et al., 2003). Based on the plant's fuel mix, each industrial plant was assumed to have one average fuel-oxygen ratio (see Table 4). In principle, oxy-

[1] Recsor stands for REmote Central SOlvent Regeneration.
[2] The term ducting is used for the transport of atmospheric pressure flue gases, whereas pipelines refer to the transport of fluids or gases at higher pressures.

fuel can be applied to all point sources, except to hydrogen plants. The CO_2 emissions of the in situ hydrogen plants are captured using MEA and ADIP-X absorbers and strippers (see Section 2.2.3).

2.2.3. Pre-combustion configurations

One main pre-combustion configuration (Pre-Central) was distinguished by having the hydrogen plant, CO_2 capture and DPC units and compressors at the central level (see Fig. 4). Decentral configurations were not deemed realistic given the relatively high capital cost of (small) SRs. To achieve a reduction level of 90%, the CO_2 is captured both from the high-pressure (HP) process gas using the solvent ADIP-X (methyldiethanolamine mixed with piperazine) and from the atmospheric pressure flue gas with the solvent MEA. The produced hydrogen is routed to the individual plants and used as fuel for the furnaces and boilers. According to Appl (1997), the SR generates HP steam (25 GJ_{LHV}/tH_2; 75 bar, 350 °C) in a waste heat recovery boiler, which could be used for solvent regeneration. However, we were unable to determine whether this waste heat is still available in modern SRs, or whether it is internally integrated. For this study, two subcases were made: with waste heat (WH) availability and without waste heat availability. In case waste heat is available, all heat is used for solvent regeneration and remaining steam was assumed to be sold to other industrial plants. Unlike oxyfuel combustion, no significant modifications are required to convert the boilers and furnaces to hydrogen firing mode (Lowe et al., 2011).

The CO_2 from the *in situ* hydrogen plants in the Botlek was also captured using a MEA and ADIP-X absorber and stripper. Next to natural gas, it was assumed that the chemical fuel gases of the Esso refinery and Rotterdam Aromatics Plant (RAP) can also be converted to hydrogen in the SR. As also indicated by an expert of Air Liquide (Kiewiet, 2012), chemical process gases produced in other industrial plants were assumed to be unsuitable to be used as feedstock for the SR. Also biomass fuel (mainly for the waste processing plant) was not considered for hydrogen substitution, because biomass is already a sustainable fuel. The CO_2 from the additional NGCC/CHP plants was captured using post-combustion capture technology, as this was found to be more cost effective than using hydrogen as a fuel (Berghout et al., forthcoming).

2.3. Performance indicators and data

The relevant formulae to assess the techno-economic performance of the CO_2 capture infrastructure configurations were taken from Berghout et al. (2013). The annually avoided CO_2 emissions Y_a (tCO_2/y) is the main technical indicator in this study. Y_a is expressed using Formula (1):

$$Y_a = Y_{bc} - [Y_{bc} + Y_{ep} - (Y_{bc} \times CR_{ip} + Y_{ep} \times CR_{ep})] - Y_{el,import} \quad (1)$$

where Y_{bc}, Y_{ep} and $Y_{el,import}$ (tCO_2/y) are the respective CO_2 emissions of the industrial plant in the base case, from the energy plants, and from imported electricity. The export of excess electricity from a CHP plant can be seen as a negative value for electricity import. CR_{ip} and CR_{ep} are the capture ratios for the industrial plants and energy plants, respectively.

The CO_2 avoidance cost C_a (€/tCO_2) is the main economic indicator. C_a is expressed using Formula (2):

$$C_a = \frac{\Delta E_{ng\times} P_{ng} + \Delta E_{e\times} P_e + \Delta E_{steam\times} P_{steam} + \alpha \times I + \Delta C_{O\&M}}{Y_a} \quad (2)$$

where ΔE_{ng} and ΔE_e are the net change in annual natural gas (GJ_{LHV}/y) and electricity (GJ_e/y) consumption, ΔE_{steam} is HP steam from the SR that is exported to other industrial plants, P_{ng}, P_e and P_{steam} are the prices of natural gas (€/GJ_{LHV}), electricity (€/GJ_e) and HP steam (€/GJ_{LHV}). I is the investment cost (€), α is the annuity factor, and $\Delta C_{O\&M}$ is the net change in O&M cost (€/y).

For configurations with a CHP that export excess electricity to the grid, the avoidance costs were determined both with and without credits for electricity export. When excluding credits for electricity export, the monetary value and (indirect) CO_2 emissions of both the exported electricity and natural gas related to this electricity, which was determined on an exergy basis (electricity: 1; heat: 0.28) (see Section 2.5), were subtracted from the total emissions and costs.

The investments were considered as Total Capital Requirement (TCR). TCR consists of various components:

- Process Plant Cost (PPC) comprising equipment and installation cost;
- Total Plant Cost (TPC) comprising PPC, engineering fees and contingencies;
- Owner costs (i.e., costs for pre-production, royalties, inventory capital, land and site preparation) and interest during construction.

In this study, all cost figures from data sources were converted to €$_{2012}$. Costs reported in other currencies were first converted to Euro using the year-average exchange rate data of OANDA (2014) for the year the cost data were reported, and were then escalated to the year 2012 using the Downstream Capital Costs Index (DCCI) (IHS CERA, 2014). Table 2 presents the general techno-economic input parameters used in this study. The impact of the input parameter values on the results were assessed by doing a sensitivity

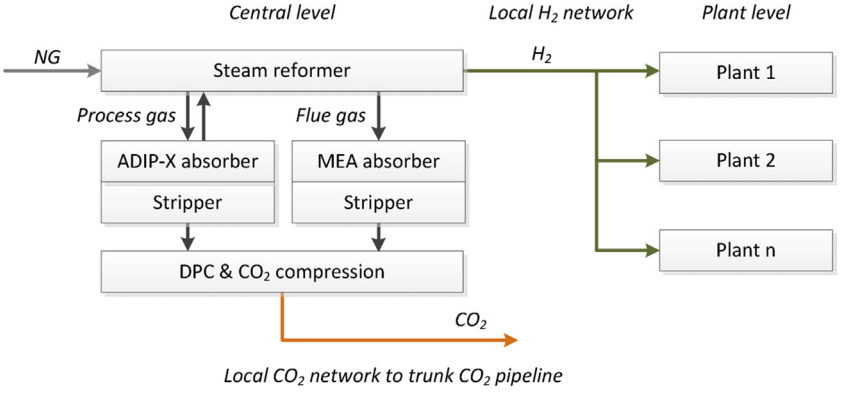

Fig. 4. Schematic overview of the pre-combustion configuration with central hydrogen production, CO_2 capture, CO_2 drying, purification, cooling and compression. NG stands for natural gas. The gray, green and orange arrows denote the natural gas, hydrogen and CO_2 flows, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

analysis. The OPEX and energy use for CO_2 capture were varied with ±30%,; the other value ranges are presented in Table 2. The trunk CO_2 pipeline that is planned to run though the Botlek is assumed to operate at 110 bar to ensure that the CO_2 flow retains its dense phase during transport. This pipeline is beyond the scope of this research.

The key performance data of the CO_2 capture technologies are presented in Tables 3–5. The costs of retrofitting and production loss due to retrofitting were excluded from the analysis. The data sets on the industrial plant characteristics (NEA, 2010) did not provide information on the CO_2 emission patterns throughout the year. Therefore, the annual CO_2 emissions were divided evenly over the year. In other words, the network was not designed to accommodate peak flow rates.

2.4. Scaling and maximum unit size

A generic scaling relation is applied to the capital costs to account for economies of scale (Formula (3)). The investment cost of a component i (M€) is expressed as proposed by Larson et al. (2005):

$$I_i = I_{i,\mathrm{ref}} \times \left(\frac{N_{\mathrm{units},i}}{N_{\mathrm{units},i,ref}}\right)^{SF_n} \times \left(\frac{S_i}{S_{i,\mathrm{ref}}}\right)^{SF_i} \quad (3)$$

where $I_{i,\mathrm{ref}}$ is the reference capital investment (M), $N_{\mathrm{units,i}}$ is the number of parallel process trains per system, $N_{\mathrm{units},I,\mathrm{ref}}$ is the number of parallel process trains in the reference system, S_i is the capacity of a single process train (unit: component dependent), $S_{i,\mathrm{ref}}$ is the reference capacity of a single process train (unit: component dependent), SF_n is the scaling factor for multiple trains (0.9 for all components), and SF_i is the economic scaling factor for component i.

The maximum processing capacity of the capture units was used to determine the number of trains needed for each CO_2 capture infrastructure configuration (see Table 6). No maximum processing unit capacity were assumed for the SCR, FGD, cooling tower, boiler and NGCC(-CHP) units. It was assumed that the maximum capacity of tanks, drums, boilers and the NGCC will not be reached for the throughput and storage quantities discussed in this study.

2.5. Energy plant

A regression curve was constructed by Kuramochi et al. (2010) for the relationship between CHP plant scale and electrical conversion efficiency based on data from the Gas Turbine World Handbook 2007–2008 (GTW, 2007). The derived regression curve can be calculated as follows:

$$\eta_{\mathrm{e,ngcc,cond}} = 0.384 \times x^{0.0619} (R^2 = 0.69) \quad (4)$$

where $\eta_{\mathrm{e,ngcc,cond}}$ is the electrical efficiency of the NGCC at full-load in condensing operation (power-only mode) and X is the capacity of the NGCC plant (MW_e). 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 efficiency in CHP operation ($\eta_{\mathrm{el,ngcc\text{-}chp}}$) was assessed as follows:

$$\eta_{\mathrm{el,ngcc-chp}} = \eta_{\mathrm{ngcc,cond}} - \eta_{\mathrm{th,cc}} \times f_{\mathrm{th,cc}} \quad (5)$$

Table 2
General techno-economic input parameters used in this study.

Parameter	Unit	Value	Sensitivity analysis	References
Annual operating time[a]	h/y	8,000		Own value
Real interest rate	%	10	±30% for total annualized capital cost	Own value
Economic lifetime[a]	years	20		Own value
Total plant cost (TPC)	%-PPC	130		Van Horssen et al. (2009)
Total capital requirement (TCR)	%-TPC	110		Van Horssen et al. (2009)
Calorific value natural gas	MJ_{LHV}/m_3	31.7		Rabou et al. (2006)
Industrial energy price (average for 2020–2030)				
Natural gas (P_{ng})[b]	€/GJ_{LHV}	10	7–13	ECN/PBL (2010); IEA (2010); CBS (2011); own estimations
Electricity (P_{el})[b]	€/GJ_e	22	15–29	
Steam (P_{steam})[c]	€/GJ_{LHV}	12	8–15	Own value
CO_2 emission factor				
Dutch electricity production[d]	$kgCO_2/GJ_e$	63	16–110	Van den Broek et al. (2011)
Natural gas	$kgCO_2/GJ_{LHV}$	56.7		Agentschap NL (2010)
Industrial boiler efficiency[e]	%	85		IEA GHG (2000)
Industrial furnace efficiency[e]	%	80		IEA GHG (2000)
Capital cost boiler	€/kW	65		Grahn et al. (2007)
O&M costs boiler	%-CAPEX	2		Own values
Max. NGCC efficiency[f]	% (LHV)	–	45–60	Own values
Max. NGCC(-CHP) efficiency[g]	% (LHV)	90	75–100	Bolland (1993); Kuramochi et al. (2010)

[a] The values for the annual operating time found in literature range between 7350 h/y and 8500 h/y (e.g., Andersson et al., 2014; Hegerland et al., 2006; Ho et al., 2011; IEA GHG, 2000; Kuramochi et al., 2012; Saygin et al., 2011). The values for the economic lifetime found in literature range between 15 and 25 years (e.g., IEA GHG, 2000; Johansson et al., 2013; Kuramochi et al., 2012; Switzer et al., 2005). The impact of the annual operating time and economic lifetime on the final results is examined indirectly by varying the total annualized capital costs (±30%) in the sensitivity analysis.

[b] Prices were determined based on quarterly energy prices over the period 1997–2008 (CBS, 2011), and extrapolated to the future. The electricity price of 22€/GJ_e is in accordance with the costs of electricity (20–24€/GJ_e) as presented in the Grand Coalition scenario over the period 2020–2030 by Van den Broek et al. (2011). The Grand Coalition scenario is based on the premise that around 50% of the CO_2 is reduced worldwide through renewable energy technologies, power plant CCS, and other CO_2 mitigation options. This scenario was assumed to be a precondition for CCS to be deployed at industrial processes.

[c] It was assumed that the onsite steam production costs in an industrial boiler equal the steam price. Based on a natural gas price of 10€/GJ_{LHV}, boiler capital costs of 85€/kW (Grahn et al., 2007) and boiler O&M costs of 2% of the total investment costs, the production costs of HP steam were calculated to be 11.8€/GJ_{LHV}. This figure is in line with steam prices (11.1€/GJ_{LHV}) indicated for industry (DACE, 2011). For the sensitivity analysis, the steam production costs were varied with the range in natural gas price.

[d] This is the CO_2 emission factor for the electricity mix in 2020–2030 as modelled in the Grand Coalition scenario by Van den Broek et al. (2011). A large uncertainty range of ±75% was assumed to examine the impact on the GHG emission reduction potential.

[e] In case efficiencies were not indicated in the environmental reports or monitoring plants, average efficiencies of 80% for industrial furnaces and 85% for steam boilers were assumed.

[f] The medium value for the newly installed NGCC electrical efficiency depends on the size of the installation and is therefore not indicated in the Table. Usually values of 50–55% were calculated (see Section 2.5).

[g] In accordance to Bolland (1993) and Kuramochi et al. (2010), it was assumed that the overall CHP efficiency (LHV basis) does not exceed 90%.

Table 3
Techno-economic parameters for post-combustion capture. The economic data pertain to a post-combustion system capturing annually 1.0 $MtCO_2$ from flue gas streams of both 4 vol% and 12–14 vol% CO_2 using one absorber and one stripper. Based on: CESAR (2011); IEA GHG (2010); NETL (2010).[a]

	Unit	Value	References
TECHNICAL			
CO_2 capture ratio (CR_{ip})	%	90	CESAR (2011); IEA GHG (2000) NETL (2010)
Regeneration heat CO_2 capture[b]	GJ_{LHV}/tCO_2	3.5–4.0	Nienoord (2012)
Electricity CO_2 capture (pumps and fans)[b]	GJ_e/tCO_2	0.1–0.3	Feron (2005); Peeters et al. (2007)
CO_2 treatment & compression	GJ_e/tCO_2	0.6	CESAR (2011)
CAPEX			
Modifications to stacks	M€/stack	0.1	Hurst and Walker (2005)
SCR/FGD units	$M€/MtCO_2/y$	25	Hurst and Walker (2005)
CO_2 capture equipment[c]	$M€/MtCO_2/y$	76 (4 vol%); 44 (12–14 vol%)	CESAR (2011); IEA GHG (2010); NETL (2010)
CO_2 treatment & compression[c]	$M€/MtCO_2/y$	12	CESAR (2011); IEA GHG (2010); NETL (2010)
OPEX			
Labor	$€/tCO_2$	0.2[d]	CESAR (2011); IEA GHG (2010); NETL (2010)
Administration & overhead	% of labor cost	28	NETL (2010)
Taxes & Insurances	$€/tCO_2$	2.0	NETL (2010)
Maintenance	% of TPC	3.8	NETL (2010)
Water usage	$€/tCO_2$	1.4	NETL (2010)
MEA	$€/tCO_2$	0.5	NETL (2010)
Activated carbon	$€/tCO_2$	0.1	NETL (2010)
Ammonia	$€/tCO_2$	0.2	NETL (2010)
Corrosion inhibitor	$€/tCO_2$	0.0	NETL (2010)
SCR catalyst	$€/tCO_2$	0.1	NETL (2010)
Other chemicals	$€/tCO_2$	0.2	NETL (2010)

[a] The parameters used for economic costs are based on three detailed studies (CESAR, 2011; IEA GHG, 2010; NETL, 2010), which describe post-combustion capture at a pulverized coal-fired (PC) power plant (12–14 vol% CO_2 concentration) and at a natural gas fired combined cycle (NGCC, 3–4 vol% CO_2 concentration). Since the CO_2 concentration in the flue gases can change significantly among and within industrial plants, data from both the capture systems applied at the PC and NGCC power plants were used to account for this factor.

[b] Regeneration heat and electricity needed for flue gas with volumetric CO_2 concentrations in the range of 4–16%; the specific regeneration energy (GJ_{LHV}/tCO_2; GJ_e/tCO_2) decreases with higher volumetric CO_2 concentrations. It is assumed there is no effect of scale on the specific energy requirement, which is concluded from a literature review (Kuramochi et al., 2013).

[c] All capital costs were standardized to and corrected for the amount of CO_2 captured of 1.0 $MtCO_2$ per year using one absorber and stripper. The cost category CO_2 capture equipment represents the direct contact cooler, absorber and stripper.

[d] The average amount of labor hours were taken from CESAR (2011), IEA GHG (2010) and NETL (2010), and multiplied with a European wage for an operator of 60,000€/FTE/y IEA GHG (2010).

where $\eta_{th,cc}$ is the heat production efficiency and $f_{th,cc}$ is the exergy factor for the heat. $f_{th,cc}$ is assumed to be 0.28 (Bolland and Undrum, 2003; Kuramochi et al., 2010). The required steam for CO_2 capture has a temperature of around 130 °C (Peeters et al., 2007).

In this study, we assumed that the maximum amount of low-temperature steam for CO_2 capture (solvent regeneration) is extracted. At the same time, the maximum total net CHP efficiency is assumed to be 90% (Bolland and Undrum, 2003; Kuramochi et al., 2010). The gas turbine electrical efficiency is assumed to be two-thirds of the NGCC efficiency in condensing mode. For the sensitivity analysis, a value range of 70–100% was used for the maximum total net NGCC-CHP efficiency (see Table 2).

The capital cost of NGCC-CHP plant is calculated using the following equation for condensing NGCC plants derived from the data presented in Gas Turbine World Handbook 2007–2008 (GTW, 2007):

$$C_{ngcc} = \frac{732 + 1.1 \times X}{1 + 0.0052 \times X} (R^2 = 0.97) \quad (6)$$

where C_{ngcc} is the total NGCC equipment cost per installed kWe rated capacity in condensing mode (&z.euro;$_{2008}$/kWe) and X is the rated capacity of the NGCC plant in condensing mode (MWe).

The data and assumptions on stand-alone industrial boilers can be found in Table 2.

2.6. Spatial footprint

The diameter of the absorber and stripper are predominantly determined by the flue gas flow rate and amine solution flow rate, respectively. A regression analysis based on data from process simulation studies was performed to derive a relation between the CO_2 flow rate (kt/d) and the diameter (m) of the absorber and stripper (see Fig. 5). The simulated absorbers contain an aqueous solvent with a mass fraction of 30% MEA designated for a flue gas with a volumetric CO_2 concentration of 12–14%. It was assumed that for each CO_2 flow rate the diameter of a 3–4 vol% CO_2 flue gas absorber is three times larger than for a 12–14 vol% CO_2 flue gas absorber (see Fig. 5). The regression functions were used to calculate the footprints of the absorbers and strippers in this study.

The physical dimensions of other capture components were mainly taken from end-to-end major equipment lists of two FEED studies (E.on, 2011a,b; ScottishPower, 2011) (see Table 7). The footprint of miscellaneous equipment (e.g., pumps) was assumed to be negligible and therefore excluded from the analysis.

Florin and Fennell (2009) reported that using a linear scaling factor to calculate the spatial footprint is overly simplistic. They

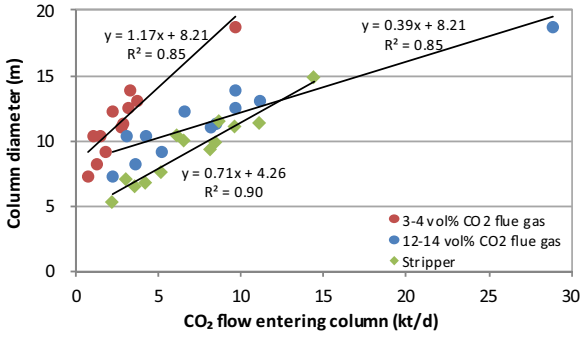

Fig. 5. Diameter (m) of absorber and stripper as a function CO_2 flow entering the column (kt/d). Data were taken from Hurst and Walker (2005), DOE/NETL (2007), Berstad et al. (2011), E.on (2011a) and Nørstebø et al. (2012). The CO_2 flow rate for the absorber was derived from the CO_2 partial pressure (12–14 vol%) in the flue gas entering the absorber.

Table 4
Techno-economic parameters for oxyfuel combustion capture.

	Unit	Value	References
TECHNICAL			
CO_2 capture ratio	%	90	Kuramochi et al. (2012)
Oxygen production	GJ_e/tO_2	0.7	ZEP (2011)
Stoichiometric O_2:CO_2 combustion ratio (weight basis)[b]			
Natural gas		1.77	
Refinery gas		1.45	Allam et al. (2005a); IEA GHG (2000)
Excess oxygen use	%	3	Zanganeh et al. (2004)
Fuel savings furnaces[b]	%	8.3	Allam et al. (2005a,b)
CO_2 treatment & compression	GJ_e/tCO_2	0.5	IEA GHG (2005b)
CAPEX			
Furnace modification[c]	M€/$MtCO_2$/y	1[c]	Allam et al. (2005a,b)
Air Separation Unit (ASU)	M€/ktO_2/d	51[d]	Allam et al. (2005a,b); IEA GHG, (2008b); Meerman et al. (2012b); Spero (2008)
Cooling water system	M€/$MtCO_2$/y	12	Allam et al. (2005a)
Flue gas gathering system	M€/$MtCO_2$/y	7	Allam et al. (2005a)
CO_2 treatment & compression	M€/$MtCO_2$/y	24	Allam et al. (2005a); IEA GHG (2008a)
OPEX			
Labor	€/tCO_2	1.0[e]	Allam et al. (2005a,b)
Administration & overhead	€/tCO_2	0.1	Allam et al. (2005a,b)
Maintenance	€/tCO_2	3.0	Allam et al. (2005a,b)
Taxes & insurances	€/tCO_2	3.0	Allam et al. (2005a,b)
Water	€/tCO_2	1.6	Allam et al. (2005a,b)
Consumables	€/tCO_2	0.3	Allam et al. (2005a,b)

[a] The current specific energy requirement for cryogenic oxygen production was found to be in the range of 0.6–0.8 GJ_e/tO_2 (160–220 kWh_e/tO_2) (ZEP, 2011).
[b] Assumptions were made on the stoichiometric O_2:CO_2 combustion ratio due to insufficient information on the fuel composition in the refinery. In the catalytic cracker, oxygen is used to burn the coke that is deposited on the surface of the catalyst ($C + O_2 \rightarrow CO_2$), resulting in a molar ratio of 1:1, which translates to a mass ratio of 0.73 (32/44). The mass ratio for the combustion of natural gas in the *in situ* NGCC/CHP plants (1.43) was derived from the volumetric O_2:CO_2 ratio as presented by C + B (C+B advies en expertise, 2014). The oxygen requirement for the refinery gases was based on two studies performing detailed analyses into oxyfuel combustion of refinery fuel gases. The mass ratio derived from these studies was 1.44 (Allam et al., 2005a) and 1.49 (IEA GHG, 2000). A value of 1.45 was used in this study. The high stoichiometric O_2:CO_2 mass ratio for refinery fuel gas is mainly due to the high methane and hydrogen concentrations in the fuel streams. However, as fuel gas compositions tend to vary considerably (also within refineries), the impact of the oxygen demand was indirectly examined by varying the energy use for oxygen production in the sensitivity analysis.
[c] Average values derived from the total furnace modification costs as reported in Allam et al. (2005a,b). It was assumed there is no scaling effect for furnace modifications considering the small scale of most furnaces.
[d] Average value based on Allam et al. (2005a,b); Meerman et al. (2013); Spero (2008), and IEA GHG (2008b) and are valid for a cryogenic Air Separation Unit and the oxygen compressor to 30 bar.
[e] No data were available on the amount of operating manpower hours in Allam et al. (2005a,b); only monetary values were presented.

Table 5
Techno-economic parameters for pre-combustion capture.

	Unit	Value	References
TECHNICAL			
Feed natural gas needed for H_2 production	GJ_{LHV}/tH_2	122	NREL (2009)
Fuel natural gas needed for H_2 production	GJ_{LHV}/tH_2	43	NREL (2009)
Electricity needed for H_2 production	GJ_e/tH_2	2.2	NREL (2009)
CO_2 produced			
Production process (CO_2 in process gas)	tCO_2/tH_2	6.9	NREL (2009)
SR furnace (CO_2 in flue gas)	tCO_2/tH_2	3.2	NREL (2009)
HP steam from waste heat SR[a]	GJ_{LHV}/tH_2	30	Appl (1997)
CO_2 capture ratio process gas	%	95	Meerman et al. (2012a)
Heat for regeneration ADIP-X	GJ_{LHV}/tCO_2	1.97	Meerman et al. (2012a)
Power for regeneration ADIP-X	GJ_e/tCO_2	0.04	Meerman et al. (2012a)
Power for CO_2 treatment & compression	GJ_e/tCO_2	0.30	Meerman et al. (2012a)
CAPEX			
SR plant (incl. WGS, PSA, SCR)		87	INL (2010); NREL (2009); Rutkowski (2005); Simbeck (2005)
CO_2 capture equipment for HP process gas	M€/$MtCO_2$/y	9	INL (2010); NREL (2009); Rutkowski (2005); Simbeck (2005)
Drying & compression	M€/$MtCO_2$/y	24	INL (2010); NREL (2009); Rutkowski (2005); Simbeck (2005)
Modified burners	M€/$MtCO_2$/y	0.1	IEA GHG (2000)
OPEX			
Labor	€/tCO_2	0.4[b]	NREL (2009)
Administration & overhead	% of labor costs	20	NREL (2009)
Taxes & insurances	€/tCO_2	3.7	(NREL, 2009)
Maintenance	€/tH_2	4.5	(NREL, 2009)
Demineralized water	€/tCO_2	0.9	(NETL, 2010)
Chemicals[c]	€/tCO_2	2.0	(NETL, 2010)

[a] The SR generates a lot of HP steam (75 bar, 350 °C) in a waste heat recovery boiler. In theory, 30 GJ_{LHV}/tH_2 of waste heat would become available.
[b] The average amount of labor hours were taken from CESAR (2011), IEA GHG (2010) and NETL (2010), and multiplied with a European wage for an operator of 60,000€/FTE/y IEA GHG (2010). Subsequently, the costs were escalated to the year 2012.
[c] SR and WGS catalysts, PSA sorbent, HP process gas solvent, MEA, SCR catalyst.

Table 6
Maximum processing capacity and economic scaling factors of CO_2 capture components.

Component	Unit	Maximum capacity	SF_i	Range sensitivity analysis	References
SCR/FGD units	$ktCO_2/d$	–	0.70	0.6–0.8	IPPC (2006)
Absorber[a]	$ktCO_2/d$	3	0.67	0.6–0.8	Kreutz et al. (2005); Sipöcz et al. (2011)
Stripper[a]	$ktCO_2/d$	10	0.67	0.6–0.8	Assumption
Cooling tower	$ktCO_2/d$	–	0.67	0.6–0.8	Assumption
Compressor	$ktCO_2/d$	8	0.67	0.6–1.0	Knoope et al. (2014); Meerman et al. (2012b)
Air separation unit[b]	ktO_2/d	6	0.65[c]	0.5–0.8	See footnote c
Steam reformer	ktH_2/d	10[d]	0.67	0.6–0.8	Kreutz et al. (2005)
Boiler[e]	MW_{th}	–	0.57	0.5–0.8	DOE/NETL (2002)

[a] Kuramochi et al. (2010) compared values of maximum processing capacity from several studies. Note that the maximum processing capacity of absorbers and strippers, which in turn depends on the maximum manageable diameter, differs per technology provider; the values for absorber diameters reported in literature are within the range of 11–20 m. A conservative assumption was made with respect to the maximum processing capacity of the absorber.
[b] According to White (2009), the largest single-train ASU in operation today is around 3.5 ktO_2/d; Air Products reports ASU units larger than 5 ktO_2/d (IEA GHG, 2007b). In this study, the maximum oxygen production capacity for a single ASU train is assumed to be 6 ktO_2/d for the short term.
[c] Medium value based on Hamelinck and Faaij (2002), IEA GHG (2008a,b), Kreutz et al. (2005), Larson et al. (2005), and Tijmensen et al. (2002).
[d] Personal communication with an expert on industrial gas separation, who stated that 10 ktH_2/d is technically feasible today.
[e] This figure is derived from the data reported for boilers of 4–111 MW_{th} scale generating steam of 18.3 bar, 263 °C (DOE/NETL, 2002). Although unknown, it is assumed that this scaling factor applies to different steam qualities and outside this capacity range as well.

Table 7
Spatial footprint of CO_2 capture components taken from literature.

	Base scale ($S_{i,ref}$)	Unit	m^2	References
Utilities				
NGCC	785	MW_e	2.10^4	IEA GHG (2005a,b)
Boiler	183	MW_{th}	$2{\cdot}10^2$	Switzer et al. (2005)
Post-combustion				
Direct Contact Cooler	56	$kgCO_2/s$	$8{\cdot}10^2$	ScottishPower (2011)
SCR/FGD units[a]	56	$kgCO_2/s$	$4{\cdot}10^2$	E.on (2011a)
Reboiler, heat exchanger, tanks	56	$kgCO_2/s$	$2{\cdot}10^3$	ScottishPower (2011)
CO_2 drying & compression[b]	56	$kgCO_2/s$	$2{\cdot}10^3$	E.on (2011b)
Oxyfuel combustion				
Air Separation Unit (one train)[c]	25	kgO_2/s	$4{\cdot}10^3$	DECC (2009); IEA GHG (2008a)
CO_2 drying & compression	65	$kgCO_2/s$	$3{\cdot}10^3$	DECC (2009)
Pre-combustion				
Steam Reformer	10	kgH_2/s	$5{\cdot}10^3$	Allam et al. (2005b)

[a] Applies to an industrial process flue gas stream with a volumetric CO_2 concentration of 12%.
[b] The spatial footprint was derived by adding op the physical dimensions of six compressors and three dryer beds. The footprints presented by the IEA GHG (2008a,b) are about two times higher. However, as the quality of these figures was unknown, it was decided to use the simulation data from the Eon FEED study (E.on, 2011b).
[c] Due to paucity of data on footprints of ASU's, figures were taken from IEA GHG (2008a,b) and DECC (2009).

suggested to take a modular approach instead and scale foot print with respect to the number of capture trains. According to Blok (2007), the capacity of many types of equipment increases with the third power of the size (determined by volume) while capital costs only increase in a quadratic way (determined by surface area). Therefore, the spatial footprint of the capture components for plant scale k (m^2) was assessed as follows:

$$A_k = \Sigma_i \left[A_{i,\mathrm{ref}} \times \left(\frac{S_i}{S_{i,\mathrm{ref}}} \right)^{SF_i} \right] \quad (7)$$

where $A_{i,ref}$ is the space requirement for component i for the reference capacity (m^2), S_i is the capacity of component i for plant scale k (unit: component dependent), $S_{i,ref}$ is the reference capacity of component i for plant scale k (unit: component dependent), and SF_i is the scaling factor for component i. Based on the reasoning of Blok (2007), a scaling factors of 0.67 (or 2/3) was used.

A 20% margin was added to the computed physical footprints to account for space needed for installation and maintenance. In the results section, the footprints of the direct contact coolers, SCR/FGD units and absorbers are lumped together in one category. The footprints of the reboilers, heat exchanger, tanks and strippers are combined as well.

2.7. Local transport

A detailed pipeline study was carried out for the Post-Recsor (CHP/CC), Oxy-Central (NGCC/CC) and Pre-Central (CHP/CC) configurations (see Pipeliner (2012a,b,c); Pipeliner (2012a,b,c)). The pipelines were designed to meet regulations regarding technical and spatial design, environment, and safety of national, regional and local government agencies (e.g., Arcadis, 2009; CMS, 2013; MIE, 2012; Municipality Rotterdam, 2010,b; NEN, 2012a,b; SenterNovem, 2006). The hierarchy used to determine the design and technical performance of the local pipeline systems is shown in Fig. 6. The optimal pipeline route was determined by studying physical obstacles – such as highways, railways, roads and dams – and available space in communal underground pipelines strips across the Botlek. Pipelines on the industrial plants' sites were included in the analysis. The flow velocity, pressure drop and wall thickness were calculated for each individual pipeline (dashed box); subsequently, the outlet conditions of the transported gas/liquid were used to calculate the technical specifications for the consecutive downstream pipeline. A feedback loop was included to ensure that the maximum allowable velocity was not exceeded. Detailed information on the technical input parameters, material selection and formulae for the technical specifications for the pipelines as well as for the compressors and blowers can be found in Appendix A. The costs of local transport were assessed for compressors, blowers,

Table 8
Economic parameters for ducting and pipelines.

	Unit	Value	References
CAPEX			
Ductwork[a]	€$_{2010}$/m^2	150	DACE (2011)
Carbon steel (CS)			
Price (2012)[b]	€/kg	1.2	Steel prices (2012)
Polyethylene coating	€/m^2	23	Conline Rhenania (2012)
Polypropylene coating	€/m^2	48	Conline Rhenania (2012)
Construction price[c]	€/m^2	1500	Pipeliner (2012a,b,c)
Stainless steel (SS; AISI 304)			
Price (2012)	€/kg	2.9	Steel prices (2012)
Polyethylene coating	€/m^2	28	Conline Rhenania (2012)
Polypropylene coating	€/m^2	56	Conline Rhenania (2012)
Construction price[c]	€/m^2	1800	Pipeliner (2012a,b,c)
Natural gas pipeline for SR[d]	M€	1.1	Gasunie (2012)
Natural gas receiving station[d]	M€	0.2	Gasunie (2012)
Flanges, junctions, appendages, etc.	% CAPEX	10	Pipeliner (2012a,b,c)
Pipeline valves	% CAPEX	20	Pipeliner (2012a,b,c)
OPEX			
O&M costs (excl. energy costs)	% CAPEX	2	Pipeliner (2012a,b,c)

[a] Typical cost for materials and installation of stainless steel ductwork with hangers and supports.
[b] Although prices can differ considerably per carbon steel grade, one price (for grade S355) was used for the sake of consistency.
[c] An aggregated number for construction costs is presented for reasons of confidentiality; this number comprises costs for excavation, welding, labor, pipeline removal, etc.
[d] For the pre-combustion configuration, an additional natural gas pipeline and receiving station has to be installed for the SR hydrogen plant. The total costs were based on an official offer made by the Dutch natural gas company (Gasunie, 2012).

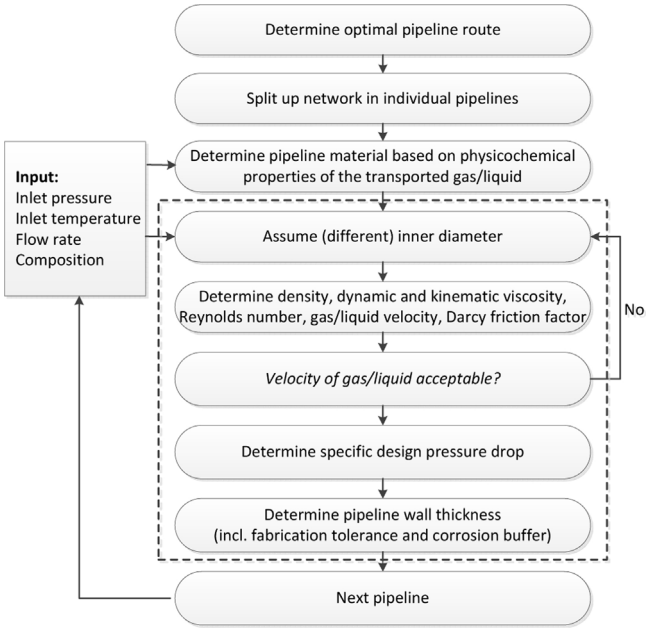

Fig. 6. Hierarchy to determine the design and technical performance of the local pipeline networks. The dashed box contains the steps needed to calculate the technical specifications of each individual pipeline.

and materials and construction (drilling, excavation, etc.) needed for pipeline transport. The main input parameters for the pipeline analysis are presented in Table 8. The total costs of the local network were allocated to the individual plants on basis of their amount of avoided CO_2.

Based on these detailed pipeline studies, a more generic pipeline analysis was carried out for the other configurations using the formulae in Appendix A. However, instead of designing an optimized pipeline network, as was done in the detailed analysis shown in Fig. 6, one fictional pipeline/duct (flue gas, O_2, CO_2, H_2) running from the capture plant to each industrial plant and vice versa, was designed and assessed. Moreover, the assessment was done starting with a typical gas velocity[3] rather than with the pipeline diameter. The costs of the flue gas ducts were assessed using a typical standard cost factor per square meter ductwork (see Table 8). Similar to Bureau et al. (2011), a maximum flue gas duct diameter of 8 inch (2.0 m) was assumed. The costs of the flue gas ducts were assessed using a typical standard cost factor per square meter ductwork (see Table 8).

3. Results

The techno-economic performance of the CO_2 capture infrastructure configurations is presented in Sections 3.1–3.3. More detailed results on the configurations can be found in Appendix B. A comparison among the configurations is made in Section 3.4, whereas the local pipeline networks and general lessons are discussed in Sections 3.5 and 3.6. For configurations with a CHP that export excess electricity to the grid, the presented avoidance costs include credits for electricity sale, unless otherwise stated.

3.1. Post-combustion configurations

Fig. 7 shows the average CO_2 avoidance cost as a function of the annual CO_2 emissions avoided (see caption for explanation), whereas a breakdown of the techno-economic performance is presented in Table 9. An aerial photograph of the industrial Botlek area with the Post-Recsor (CHP/CC) configuration is shown in Fig. 8.

The results indicate that Post-Recsor (CHP/CC) is the most realistic and cost effective (7€/tCO_2) post-combustion configuration for the Botlek. Moreover, it has the highest CO_2 emission reduction

[3] A velocity of 11 m/s was used for flue gas transport, which is at the high end of the range (4.5–12 m/s) that was reported by Lindeburg (2013). The high end value was chosen to minimize the diameter of the flue gas ducts. For the same reason, a velocity of 10 m/s was assumed for gaseous oxygen, hydrogen and CO_2 transport. This value is within the gas velocity range (5–20 m/s) indicated by Knoope et al. (2014) for gaseous CO_2 and by a chemical engineering handbook (Branan, 2005) for gaseous oxygen and hydrogen pipeline transport. For high-pressure (110 bar) CO_2 pipeline transport a velocity of 6 m/s was used, which is in accordance with values presented by Knoope et al. (2014) for liquid CO_2 transport (0.5–6 m/s). A standard Darcy friction factor of 0.02 was used.

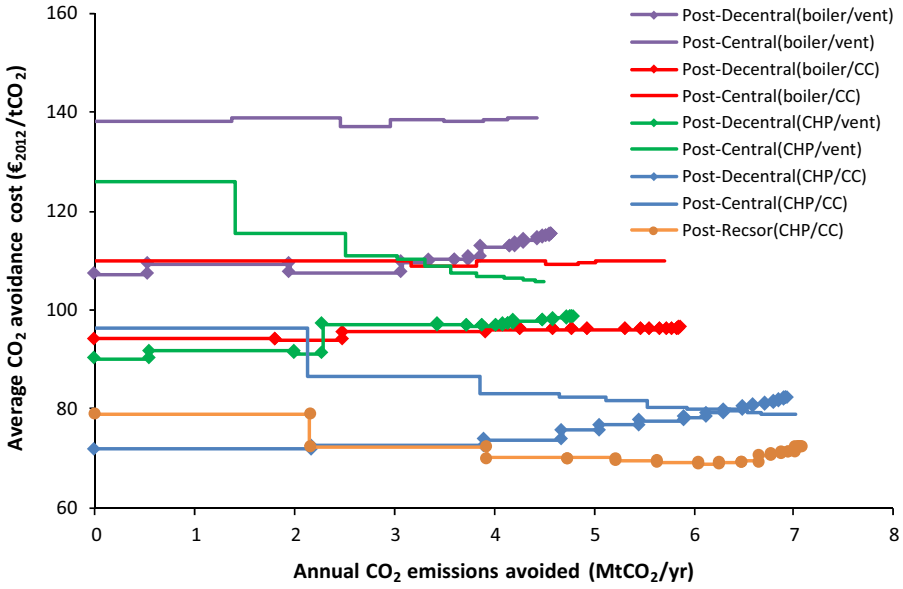

Fig. 7. Average CO_2 avoidance costs as a function of total annual CO_2 emissions avoided for the post-combustion configurations of the sixteen industrial plants in the Botlek. For the Post-Decentral configurations, the annual CO_2 emissions of the industrial plants on the *x*-axis are ordered from the plant with lowest average CO_2 avoidance cost to the plant with the highest average CO_2 avoidance cost. For the Post-Central and Post-Recsor configurations, the plants are ordered from the plant with the highest amount of annual CO_2 emissions avoided (Esso refinery) to the plant with the lowest amount of annual CO_2 emissions avoided (Biopetrol). The costs for the CHP cases include credits for electricity sale. Each energy supply option is assigned a different color; the lines with and without markers (diamonds) represent the decentral and central configurations, respectively. Post-Recsor (CHP/CC) is given a color and type of marker (dots) on its own. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

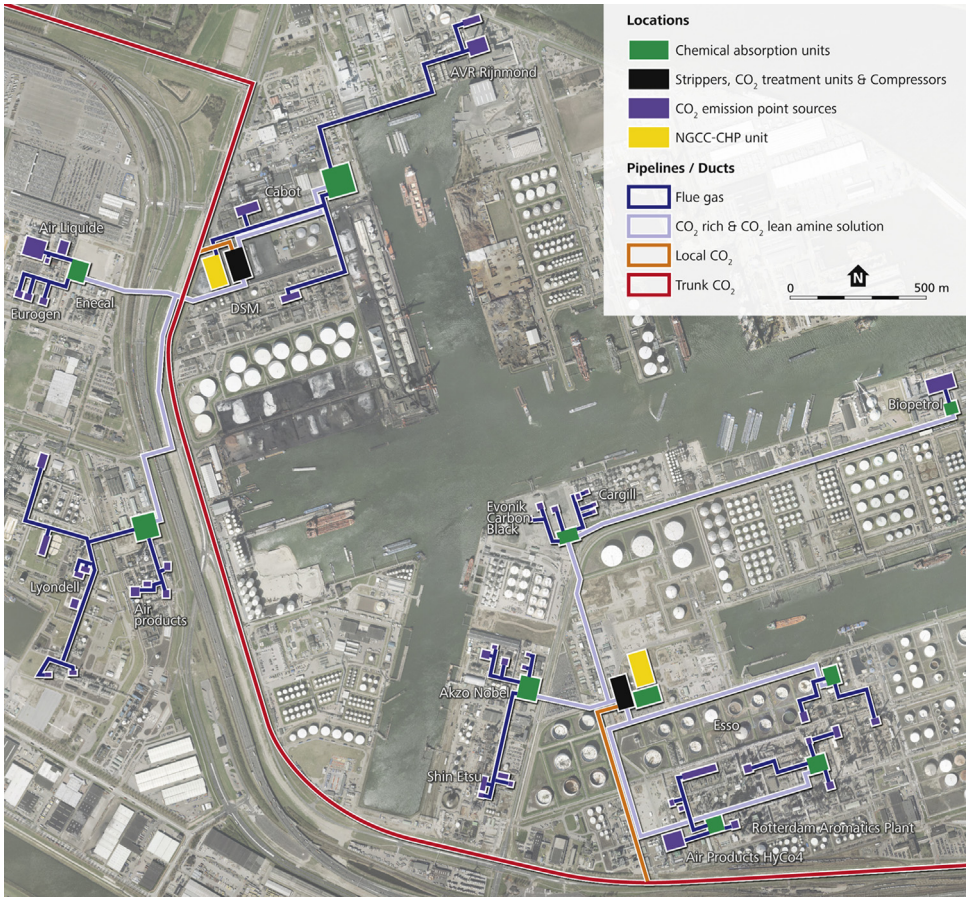

Fig. 8. Aerial photograph of the industrial Botlek area with the Post-Recsor (CHP/CC) configuration.

Table 9
Techno-economic performance of post-combustion configurations in the Botlek.

	Unit	Boiler				CHP				
		Decentral		Central		Decentral		Central		Recsor
		Vent	CC	Vent	CC	Vent	CC	Vent	CC	CC
Technical performance										
CO_2 base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1
CO_2 generated in Botlek	Mt/y	8.7	8.7	8.7	8.7	9.3	10.1	9.7	10.3	10.3
CO_2 captured	Mt/y	6.4	7.8	6.4	7.8	9.1	9.1	6.4	9.3	9.3
CO_2 avoided (w elec. sale)	Mt/y	n/a	n/a	n/a	n/a	4.7	6.9	4.5	7.0	7.1
CO_2 avoided (w/o elec. sale)	Mt/y	4.6	5.9	4.4	5.7	4.5	6.1	3.8	6.1	6.1
Heat demand	PJ_{th}/y	23	24	23	24	23	29	23	28	28
Electricity demand	PJ_e/y	5	6	7	9	5	8	7	11	10
Economic performance										
CAPEX	M€	809	939	629	722	1964	1964	1432	1731	1652
OPEX	M€/yr	52	63	53	64	111	111	86	113	109
Energy costs (w elec. sale)	M€/yr	n/a	n/a	n/a	n/a	181	231	222	237	212
Energy costs (w/o elec. sale)	M€/yr	372	414	486	478	209	318	256	284	262
Average CO_2 avoidance cost (with electricity sale)	€/tCO_2	n/a	n/a	n/a	n/a	96	83	106	79	73
Average CO_2 avoidance cost (without electricity sale)	€/tCO_2	113	100	139	110	106	108	133	91	93
Physical footprint										
Total	$\cdot 10^3$ m^2	34	53	17	20	116	232	101	119	119

potential (7.1 $MtCO_2/y$). The separated absorber-stripper arrangement shows a favorable combination of economic scale effects for the clustered strippers, CO_2 treatment units and compressors, and low local flue gas transport costs due to short distance between the CO_2 emission point sources and partly decentralized absorbers. Furthermore, sufficient space is available as the capture components and energy plants are located at several places in the Botlek and the space consuming flue gas ducts are relatively short (see Fig. 8). The local pipeline network required for Post-Recsor (CHP/CC) is technically and legally feasible, although several new pipeline tracks are needed for the large diameter pipelines (up to one meter) that circulate around 250 Mt of aqueous amine solution per year. Using a CHP plant that captures its own CO_2 – instead of a boiler in combination with electricity import – was found to be most cost effective due to the high energy efficiency of the CHP and credits for excess electricity sale.

Post-Recsor (CHP/CC) is already economically preferable over CO_2 capture at industrial plant site (Post-Decentral (CHP/CC)) when applying CO_2 capture to both the Esso refinery and the waste processing plant, due to economies of scale for both the capture units and CHP plants (see intersection of blue and orange lines in Fig. 7). The economic scale effects are illustrated by the downward trend of the abatement curves of the Post-Recsor and Post-Central configurations with higher amounts of CO_2 avoided, whereas an opposite trend is observed for the decentral configurations. The analysis also shows that it would be very challenging, maybe even impossible, to accommodate all the capture equipment and energy plants on certain industrial plan sites (e.g., Biopetrol, Evonik Carbon Black, Cargill, Lyondell, Air Products (old plant), Enecal, Eurogen, Air Liquide), thus rendering Post-Decentral configurations unrealistic. In contrast, sufficient space is lying idle ($\sim 200 \times 10^3$ m^2) north-west of the Esso refinery for the fully centralized configurations. Furthermore, Post-Central (CHP/CC) shows even slightly lower average avoidance costs (91€/tCO_2) than Post-Recsor (CHP/CC) (93€/tCO_2) when excluding credits for electricity sale. Nevertheless, the Post-Central configurations involve large diameter flue gas ducts (max. 80 inch); some tracks showing over ten ducts running in parallel. Hence, pooling flue gases from multiple industrial plants and routing these flows to central amine absorbers is unlikely from a spatial point of view.

3.2. Oxyfuel combustion configurations

The abatement curves and techno-economic performance of the oxyfuel configurations are shown in Fig. 9 and Table 10, respectively. An aerial photograph of the industrial Botlek area with the Oxy-Hybrid (El) configuration is shown in Fig. 10. As the CO_2 emissions of the *in situ* hydrogen plants cannot be captured via oxyfuel combustion, MEA and ADIP-X absorbers and strippers were used instead (see also Section 3.3).

The results show that Oxy-Hybrid (El) (central oxygen production and decentral CO_2 treatment & compression) is the most realistic oxyfuel configuration. Oxy-Hybrid (El) has an emission reduction potential of 5.8 $MtCO_2/y$ and displays average avoidance costs of 61€/tCO_2 when applying CO_2 capture to all sixteen industrial plants (hereafter referred to as full deployment). Oxy-Central (El) shows slightly lower avoidance costs (60.6€/tCO_2) under full deployment, due to economies of scale for centralized CO_2 treatment & compression units and low energy expenses for the local atmospheric pressure CO_2 gas transport. However, Oxy-Hybrid (El) is economically preferable up to 4.6 $MtCO_2/y$ avoided (five largest industrial plants). Furthermore, the large diameter CO_2 pipelines in the Oxy-Central configurations do not fit in the designated pipeline strips. Instead, Oxy-Hybrid (El), which has a local high pressure (110 bar) CO_2 transport network, is more realistic. For the same reason, medium pressure (~30 bar) instead of atmospheric pressure oxygen transport is needed, which increases the average avoidance costs with around 5€/tCO_2. As can be seen in Fig. 9 and Table 10, the cost difference between atmospheric and medium pressure CO_2 pipeline transport is much smaller (up to 1€/tCO_2). Electricity import was found to be more cost effective than using a NGCC as the latter operates in power-only mode, which is less efficient than CHP mode. However, as the avoidance costs of Oxy-Hybrid (El) and Oxy-Hybrid (NGCC/CC) lie close together, the ranking of the Oxy-Hybrid configurations in terms of avoidance cost will depend strongly on the energy prices and grid CO_2 emission factor (see Section 3.4).

Oxy-Hybrid (El) was found to be more cost effective than CO_2 capture at industrial plant level under full deployment, due to economies of scale for the ASU's, which compensate the costs for local oxygen and CO_2 transport. Moreover, whereas sufficient space is available in the center of the Botlek to install the capture equip-

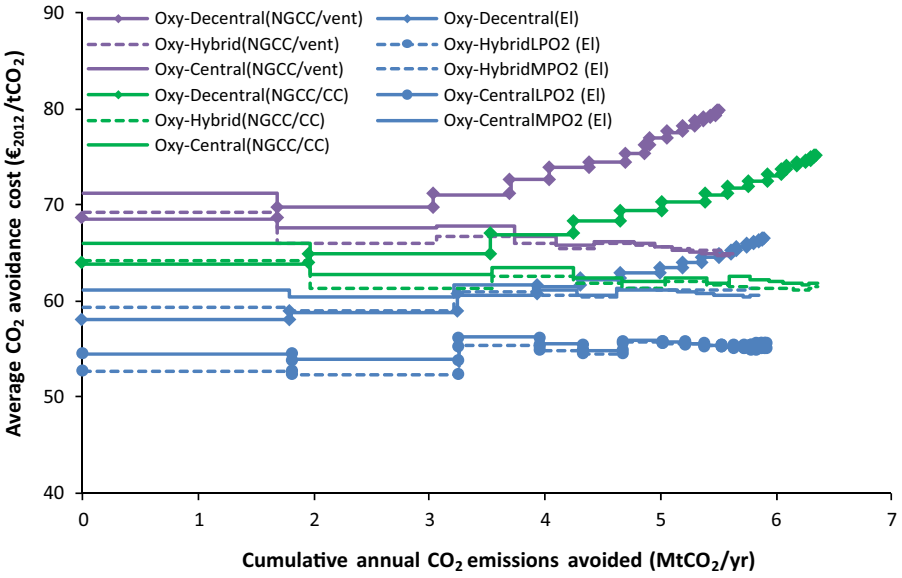

Fig. 9. Average CO_2 avoidance costs as a function of total annual CO_2 emissions avoided for the oxyfuel combustion configurations of the sixteen industrial plants in the Botlek. For the Oxy-Decentral configurations, the annual CO_2 emissions of the industrial plants on the *x*-axis are ordered from the plant with lowest average CO_2 avoidance cost to the plant with the highest average CO_2 avoidance cost. For the Oxy-Central and Oxy-Hybrid configurations, the emissions are ordered from the plant with the highest amount of annual CO_2 emissions avoided (Esso refinery) to the plant with the lowest amount of annual CO_2 emissions avoided (Biopetrol). Each energy supply option is assigned a different color; the dashed lines and solid lines with and without markers (diamonds) represent the hybrid, decentral and central configurations, respectively. LPO_2 (~2 bar) and MPO_2 (~30 bar) stand for low and medium pressure oxygen transport, respectively. All configurations were assessed for MPO_2, except for Oxy-Hybrid LPO_2 (El) and Oxy-Central LPO_2 (El), which are represented by the lines with dot-shaped markers. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

Fig. 10. Aerial photograph of the industrial Botlek area with the Oxy-Hybrid (El) configuration.

Table 10
Techno-economic performance of oxyfuel combustion configurations in the Botlek. LP and MP stand for low- and medium-pressure pipeline transport, respectively.

	Unit	Decentral ASU			Central ASU				Central ASU			
		Decentral compression			Decentral compression				Central compression			
		El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC
Technical												
CO_2 base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1
CO_2 generated	Mt/y	6.8	7.7	8.1	6.8	6.8	6.8	7.6	6.8	6.8	7.6	7.6
CO_2 captured	Mt/y	6.1	6.1	7.0	6.1	6.1	6.1	6.9	6.1	6.1	6.1	6.9
CO_2 emis. avoid.	Mt/y	5.9	5.5	6.3	5.8	5.8	5.6	6.4	5.9	5.8	5.6	6.4
Heat demand	PJ_{th}/y	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Elec. demand	PJ_e/y	8.0	8.0	8.0	7.0	8.0	8.0	8.0	7.0	8.0	8.0	8.0
Economic												
CAPEX	M€	1347	1712	1900	1014	1068	1278	1390	974	1017	1263	1397
OPEX	M€/yr	58	72	81	57	58	67	76	59	60	69	77
Energy costs	M€/yr	175	166	173	153	175	147	152	153	175	147	152
Aver. avoid. cost	€/tCO_2	66	80	75	56	61	65	61	55	61	65	62
Footprint												
Total	$\cdot 10^3$ m^2	50	75	82	34	39	48	52	33	33	47	52

ment, several industrial plant sites show too little space for capture equipment (see Fig. 10). The footprints of the decentral CO_2 compressors in the Oxy-Hybrid configurations are small enough to be placed on most plant sites. The few premises with limited plot space (i.e., Enecal and Eurogen) may need to place (and share) compression units at other plant sites. Despite the cost advantage of Oxy-Hybrid (El) under full deployment, Oxy-Decentral (El) (see Fig. 9) shows slightly lower average avoidance costs up to about a cumulative amount of 3.9 $MtCO_2/y$ avoided, because of the oxygen transport costs between the central ASU's and industrial plants. When applying decentralized CO_2 capture to small industrial plants as well, the average avoidance costs increase rapidly.

3.3. Pre-combustion configurations

The average abatement curves and techno-economic performance of the pre-combustion configurations are shown in Fig. 11 and Table 11, respectively. An aerial photograph of the industrial Botlek area with the Pre-Central (El) configuration is shown in Fig. 12. As some industrial plants combust little natural gas, and large amounts of chemical fuel gases that are not suitable for replacement with hydrogen, the total maximum amount of CO_2 that can be avoided via pre-combustion capture (4.9 $MtCO_2/y$; see Table 11) is considerably smaller than for post- and oxyfuel combustion capture (Post-combustion: 7.1 $MtCO_2/y$; Oxyfuel: 6.4 $MtCO_2/y$). As the fuel gases in the Cabot plant cannot be replaced with hydrogen, CCS was not applied to this plant. A drop in the abatement curves can be observed for the second and fifth plant, which are the *in situ* Air Liquide and Air Products hydrogen plants in the Botlek. These plants show lower avoidance costs, because they do not need to be fueled with hydrogen, but require merely additional absorbers (MEA and ADIP-X) and strippers for CO_2 capture.

Pre-Central WH (El) is the most economic configuration (81€/tCO_2; 4.4 $MtCO_2/y$) in case the maximum amount of waste heat from the SR is available for solvent regeneration, whereas Pre-

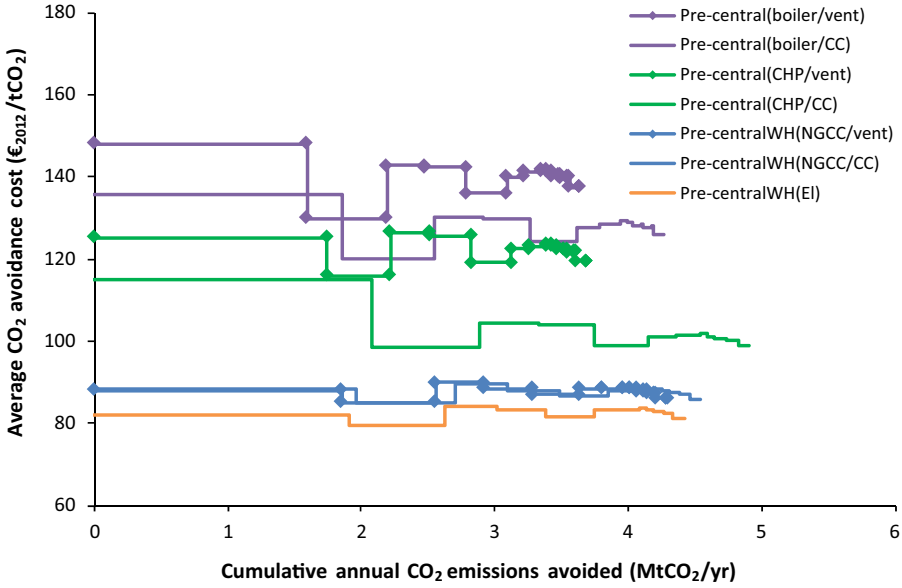

Fig. 11. Average CO_2 avoidance costs as a function of annual total CO_2 emissions avoided for the pre-combustion configurations of fifteen industrial plants in the Botlek. Pre-combustion capture was not applied to the CABOT plant as the fuel gases cannot be replaced. In the abatement curves, the plants are ordered from the plant with the highest amount of annual CO_2 emissions avoided (Esso refinery) to the plant with the lowest amount of annual CO_2 emissions avoided (Biopetrol). The costs for the CHP cases include credits for electricity sale. Each energy supply option (both vent and CC) is assigned a different color; lines with and without markers (diamonds) represent the vent and CC cases, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

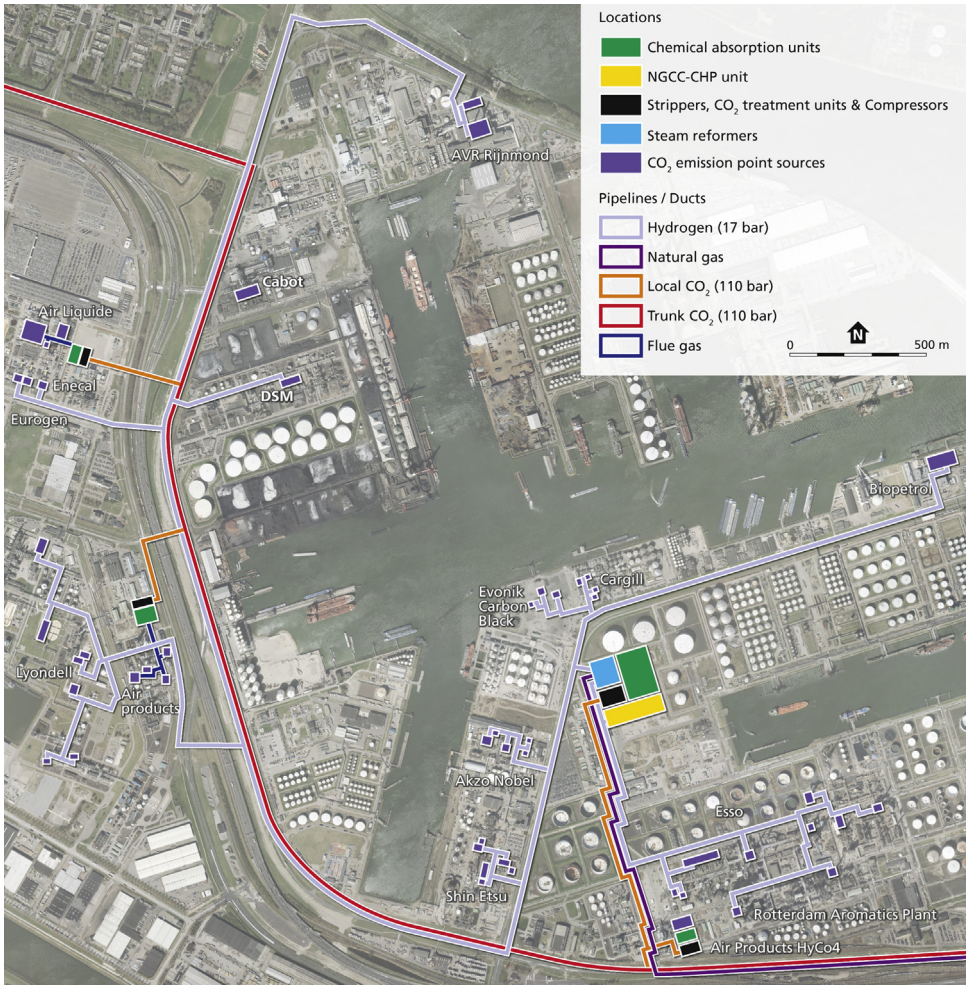

Fig. 12. Aerial photograph of the industrial Botlek area with the Pre-Central (CHP/CC) configuration.

Central (CHP/CC) is most cost effective (99€/tCO_2; 4.9 $MtCO_2$/y) in case waste heat is unavailable, regardless of the amount of CO_2 avoided. Both options show low spatial footprints (Pre-Central WH (El): 23×10^3 m^2; Pre-Central (NGCC/CC): 47×10^3 m^2) and capture equipment can be easily located in the center of the Botlek (see Fig. 12). As the hydrogen exits the SR at around 17 bar, no additional compression is required, resulting in low transport costs and small diameter pipelines. A potential bottleneck indicated by the national

Table 11
Techno-economic performance of pre-combustion configurations in the Botlek. The Pre-Central WH (boiler) configuration requires no additional energy production for CO_2 capture; hence, the differentiation in venting and CO_2 capture (CC) of the energy plant emissions is not applicable.

	Unit	Boiler			CHP			
		No waste heat		Waste heat	No waste heat		Waste heat	
		Vent	CC	El	Vent	CC	Vent	CC
Technical performance								
CO_2 base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1
CO_2 eligible for capture	Mt/y	4.9	4.9	4.9	4.9	4.9	4.9	4.9
CO_2 generated in Botlek	Mt/y	8.0	8.0	7.2	8.4	8.7	7.5	7.8
CO_2 captured	Mt/y	4.9	5.5	4.9	4.9	6.2	4.9	5.2
CO_2 avoided (w elec. sale)	Mt/y	n/a	n/a	n/a	3.7	4.9	n/a	n/a
CO_2 avoided (w/o elec. sale)	Mt/y	3.5	4.3	4.4	3.3	4.4	4.3	4.5
Heat demand	PJ_{th}/y	12	12	12	12	14	12	14
Electricity demand	PJ_e/y	3	3	3	3	4	3	3
Economic performance								
CAPEX	M€	665	739	650	903	1054	755	795
OPEX	M€/y	57	65	57	67	84	61	65
Energy costs (w elec. sale)	M€/y	n/a	n/a	n/a	268	277	n/a	n/a
Energy costs (w/o elec. sale)	M€/y	365	385	226	284	302	220	233
Average CO_2 avoidance cost (w elec. sale)	€/tCO_2	n/a	n/a	n/a	120	99	n/a	n/a
Average CO_2 avoidance cost (w/o elec. sale)	€/tCO_2	137	126	81	137	115	86	86
Physical footprint								
Total	$\cdot 10^3$ m^2	23	25	23	42	47	30	31

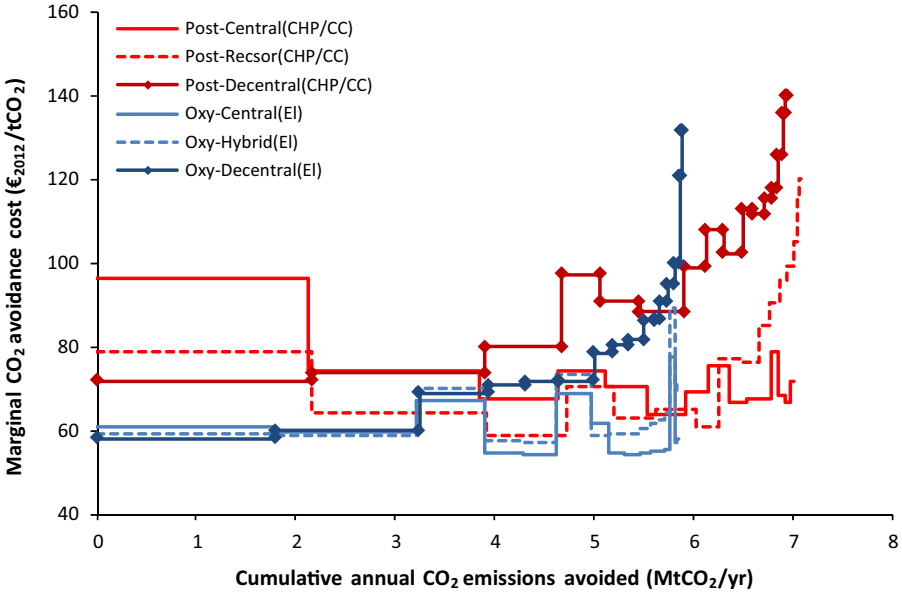

Fig. 13. Marginal CO_2 avoidance costs as a function of annual total CO_2 emissions avoided for the most costs-effective CO_2 capture infrastructure configuration in the Botlek. In the abatement curves, the plants are ordered from the plant with the highest amount of annual CO_2 emissions avoided (Esso refinery) to the plant with the lowest amount of annual CO_2 emissions avoided (Biopetrol). The avoidance costs of Post-Central (CHP/CC) and Post-Recsor (CHP/CC) include credits for excess electricity sale. Different colors are assigned to the post-(red) and oxyfuel (blue) configurations; the dashed lines and solid lines with and without markers (diamonds) represent the Post-Recsor/Oxy-Hybrid, decentral and central configurations, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

natural gas supplier is that it would currently be unable to supply the high natural gas demand for the SR in the Pre-Central configurations due to infrastructure limitations (Gasunie, 2012). However, this issue can be tackled by taking timely action.

The avoidance costs and amount of avoided CO_2 emissions of the Pre-Central WH configurations lie relatively close together (81–86€/tCO_2 and 4.3–4.5 $MtCO_2$/y under full deployment), which illustrates the relatively low impact of the energy supply mode on the performance of the configurations. The small impact is due to the fact that the NGCC is operated in power only mode and no electricity is exported to the grid, which shows limited advantages compared to the boiler case with electricity import. Conversely, the energy supply mode has a larger impact on the Pre-Central configurations (w/o waste heat availability), which is shown by the large differences in avoidance costs and avoided emissions under full deployment (see Table 11). Using a CHP with CO_2 capture is economically preferable, due to the operation of a larger and more efficient energy plant and credits for electricity sale. Also without electricity credits, the CHP/CC case (115€/tCO_2) is the preferred option.

3.4. Sensitivity analysis and comparison of configurations

A sensitivity analysis was carried out to determine the impact of the economic scaling factors, energy prices, pipeline/ducting costs and grid CO_2 emission factor on the ranking of the configurations within each capture technology in terms of avoidance costs under full deployment. The ranges of these parameters can be found in Tables 2 and 6. The CAPEX of the pipelines/ducting were varied with ±30%.

The sensitivity analysis shows that for each capture technology the ranking of the configurations in terms of average avoidance costs remains the same under any combination of the varied economic scaling factors and pipeline/ducting costs. Table 12 shows the most cost effective configuration for each capture technology (with credits for electricity sale) under medium conditions and when changing one input parameter (energy price or grid CO_2 emission factor). The results show that Post-Recsor (CHP/CC) is always the most cost effective post-combustion configuration. Oxy-Hybrid (EI) and Oxy-Hybrid (NGCC/CC) show the lowest average avoidance costs for oxyfuel combustion, which illustrates that the optimal energy supply mode depends heavily on the energy prices and grid emission factor. Note that the more cost effective oxyfuel configurations with low pressure O_2 and CO_2 pipeline transport are not presented in Table 12 as these were not considered realistic from a spatial perspective (see Section 3.2). The configurations Pre-Central WH (EI) and Pre-Central (NGCC/CC) always show the lowest costs, regardless of the energy prices and grid emission factor.

Fig. 13 presents the *marginal* abatement curves of the most costs-effective post- and oxyfuel combustion configurations in the Botlek (Pre-Central was excluded because of the higher avoidance costs and lower emission reduction potentials compared to the other technology configurations). The peaks in the abatement curves are due to either CO_2 capture from an *in situ* hydrogen plant or the addition of an extra CO_2 capture component (i.e., absorber, stripper, ASU), which results in lower economic scale effects (the amount of captured CO_2 is divided over the total amount of capture units), and therefore increases the specific capital costs (€/tCO_2). The graph shows that the marginal avoidance costs of the oxyfuel configurations are lower than for the post-combustion configurations up to about 3.2 $MtCO_2$/y avoided. For emission reductions between 3.2 and 5.8 $MtCO_2$/y, Oxy-Central (EI) and Oxy-Hybrid (EI) are overall most cost effective. Post-Recsor (CHP/CC) and Post-Central (CHP/CC) do, however, have a larger emission reduction potential, mainly because of the energy supply mode (CHP/CC). Starting CO_2 capture using oxyfuel combustion at the largest point sources seems to be the best option, considering the possibility that CO_2 capture may not be applied to all industrial plants in the future, but only to large emitters. Although Oxy-Decentral (EI) is more cost effective than Oxy-Hybrid (EI) for the first three industrial plants, the latter is recommended as the cost difference is rather small and a hybrid configuration is more realistic from a spatial perspective.

Fig. 14 presents the techno-economic performance of Post-Recsor (CHP/CC), Oxy-central (NGCC/CC) and Pre-Central WH (EI) and Pre-Central (CHP/CC). These configurations were selected as

Table 12
Most cost effective CO_2 capture infrastructure configuration for each capture technology under full deployment in terms of average avoidance costs (with credits for electricity sale) depending on variations in the energy prices and grid CO_2 emission factor. The minus indicates that the best configuration under medium conditions is still most cost effective when changing the input parameter.

	Best configuration under medium conditions	Natural gas price		Electricity price		Grid CO_2 emission factor	
		Low	High	Low	High	Low	High
Post-combustion	Post-Recsor (CHP/CC)	–	–	–	–	–	–
Oxyfuel combustion	Oxy-Hybrid (El)	Oxy-Hybrid (NGCC/CC)	–	–	Oxy-Hybrid (NGCC/CC)	–	Oxy-Hybrid (NGCC/CC)
Pre-combustion (waste heat)	Pre-Central WH (El)	–	–	–	–	–	–
Pre-combustion(no waste heat)	Pre-Central (NGCC/CC)	–	–	–	–	–	–

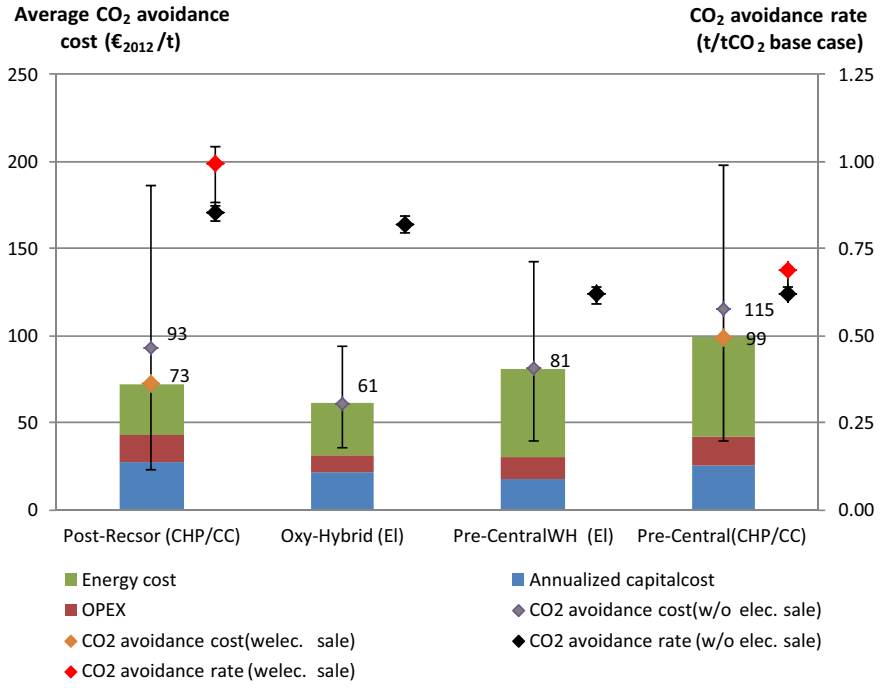

Fig. 14. CO_2 avoidance costs and CO_2 avoidance rate for Post-Recsor (CHP/CC), Oxy-Hybrid (El), Pre-Central WH (El) and Pre-Central (NGCC/CC) under full deployment. The error bars show the sensitivity of the results with respect to the uncertainty in the input parameters. The error bars for the costs of Post-Recsor (CHP/CC) and Pre-Central (CHP/CC) apply to average avoidance costs with credits for electricity sale (orange diamonds). (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

the best configurations from both an economic and spatial perspective. Figure 14 also shows the combined effect of the uncertainty in the input parameters on the average avoidance cost and CO_2 avoidance rate. The high and low end sides of the error bars represent the extreme pessimistic and extreme optimistic case, in which all parameters are set a their least and most favorable values, respectively.

Oxy-Hybrid (El) shows the lowest average avoidance costs (61€/tCO_2). The combination of high OPEX and capital costs renders Post-Recsor (CHP/CC) less attractive than Oxy-Hybrid (El), even when electricity sale credits are included. Although Pre-Central WH (El) and Pre-Central (NGCC/CC) have (relatively) low CAPEX, the high energy expenses, especially for hydrogen production, and low avoided CO_2 emissions result in high avoidance cost, even when the maximum amount of waste heat from the SR is available for solvent regeneration. The uncertainties in the avoidance costs of Oxy-Hybrid (El) are considerably less than for the other configurations. Nevertheless, there is still significant overlap between the error bars (see Fig. 14). Post-Recsor (CHP/CC) shows the largest amount of CO_2 emissions avoided, followed by Oxy-Central (El), Pre-Central (CHP/CC) and Pre-Central WH (El). All four CO_2 capture infrastructure configurations are realistic from a spatial viewpoint. The higher footprint of Post-Recsor (CHP/CC) (115×10^3 m^2) compared to the other three configurations ($23 \times 10^3 \times 47 \times 10^3$ m^2) is due to the high number of absorbers and strippers, which are located at several places in the Botlek.

The impact of the individual key input parameter values on the economic performance of Post-Recsor (CHP/CC), Oxy-Hybrid (El) and Pre-Central WH (El) is shown in Fig. 15.

The avoidance costs of Post-Recsor (CHP/CC) and Pre-Central WH are influenced more strongly and by more input parameters than Oxy-Hybrid (El) (see Fig. 15). The avoidance cost of Post-Recsor (CHP/CC) is mainly influenced by the energy prices, whereas Oxy-Hybrid (El) shows large sensitivity to energy use for CO_2 capture[4] and the electricity price. The avoidance cost of Pre-Central WH (El) is strongly determined by the SR production efficiency and natural gas price. These two strong sensitivities are also observed for the Pre-Central configurations without waste heat availability. A

[4] This applies to fuel savings due to oxyfuel combustion, energy use for CO_2 capture from the *in situ* hydrogen plants, and for oxygen and CO_2 compression.

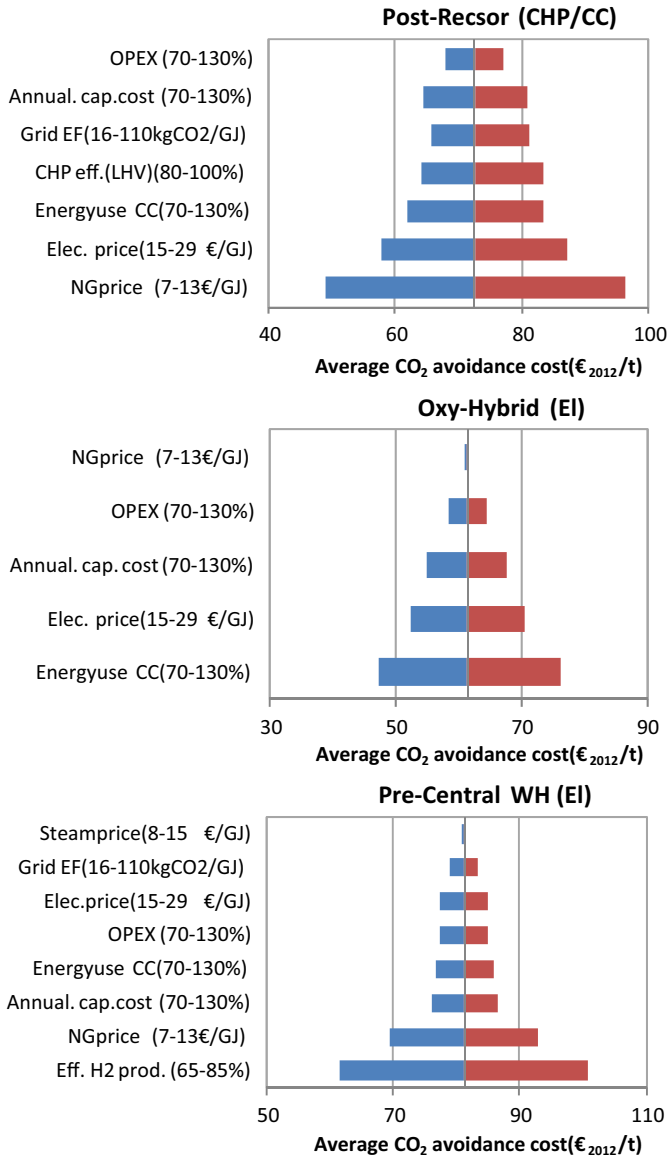

Fig. 15. Sensitivity of CO_2 avoidance costs of the Post-Recsor (CHP/CC), Oxy-Hybrid (El) and Pre-Central WH (El) configurations on variations in the key input parameters. Note that the values on the *x*-axes are different.

low natural gas price or high electricity price would make Post-Recsor (CHP/CC) the configuration with the lowest avoidance cost (49€/tCO_2 with low natural gas price; 57€/tCO_2 with high electricity price) when electricity credits are included; however, without these credits, Oxy-Hybrid (El) (either with NGCC/CC or electricity import) is economically preferable over Post-Recsor (CHP/CC), regardless of the energy prices. A combination of low capital costs for pre-combustion capture and high energy use for oxyfuel would set Pre-Central WH (El) on an equal cost footing as Oxy-Hybrid (El). Furthermore, Pre-Central WH (El) would be the most cost effective configuration under, for instance, a high SR production efficiency and low natural gas price; however, the combinations of input parameters rendering Pre-Central WH (El) the configuration with the lowest avoidance cost are rather unlikely.

3.5. Local pipeline networks

The required local pipeline networks were found to be feasible from a technical, spatial, legal and safety perspective, except for the large scale flue gas duct network for Post-Central. Although new pipeline routes are needed to accommodate some of the new pipelines (e.g., for the aqueous amine solution circulation), no insurmountable issues were identified. However, the new pipeline routes involve additional costs for drilling, excavation and crossings of numerous bottlenecks like other pipelines, roads and railway tracks. These costs may increase as the demand for space in the pipeline strips is expected to grow in the near future, which would entail alternative pipeline routing and more bottlenecks to overcome (Pipeliner, 2012a,b,c). Part of these spatial issues can be averted by increasing the operating pressure and thus the diameter of the pipelines (e.g. for oxygen pipelines). Although the additional capital and energy costs to deal with these spatial issues can be substantial, they are relatively minor compared to the total capture costs. For each industrial area, the optimal local transport network will depend on local conditions, not only related to space, but also to safety and environmental regulations.

3.6. General lessons

The method developed in this study has proven useful to assess the performance of infrastructure configurations for large scale CO_2 capture in industrial zones. Considering the large advantages of centralized and hybrid configurations compared to decentral CO_2 capture, the findings of this study are expected to apply to other industrial areas as well, even when these have a different mix of industrial plants (e.g., including cement or iron & steel plants). Especially areas with multiple small industrial plants would benefit from joint CO_2 capture initiatives. Nevertheless, more generic insight is needed under which exact conditions (e.g., type and number of plants, point source size, distance among plants) certain configurations become preferable over other configurations. Clustering capture equipment for industrial areas with a higher number of industrial plants and CO_2 emissions than the Botlek will likely not provide much more cost benefits, due to the maximum capacities, and thus economies of scale, of the capture units. After all, Post-Recsor (CHP/CC) and Oxy-Hybrid (El) in the Botlek already show multiple absorbers, strippers, compressors and/or ASU's. The feasibility of CO_2 capture infrastructure configurations in industrial areas with even lower plot space availability than the Botlek remains to be seen for Post-Recsor. For Oxy-Hybrid and Pre-Central, however, spatial footprints can be reduced significantly by purchasing oxygen and hydrogen from industrial gas producers and import it via pipelines into the industrial area. A factor that could affect the selection of the capture technology is the availability of (low cost) waste heat in an industrial area. For example, the techno-economic performance of Post-Recsor (CHP/CC) in the Botlek could improve considerably by utilizing waste heat for solvent regeneration that comes from the recently installed steam pipeline (40 bar, 420 °C) across the Botlek (Visser and Smit Hanab, 2014). Unfortunately, data limitations on waste heat prohibited the assessment on the performance improvement potential of Post-Recsor (CHP/CC).

Finally, as the average CO_2 avoidance costs differ significantly per industrial plant, a full-fledged network across the area emerging simultaneously is not very likely to occur. If CCS will come into existence, it will probably take off with several key anchor projects and the rollout of a basic transport network. Subsequently, a sequential connection of other industrial plants may follow. Such a gradual deployment will pose all kind of new challenges for the industrial plants and authorities, and a fine-tuned strategy will have to be formulated to address these challenges in an adequate fashion. More insight is needed on the conditions needed to deploy large scale CO_2 capture infrastructure configurations in an optimal manner over time. In part B of this analysis, several deployment pathways for Post-Recsor (CHP/CC) and Oxy-Hybrid (El) are investigated, which differ *inter alia* regarding the sequence in which industrial plants start capturing CO_2, the number of deployment

steps, and whether CO_2 capture units and energy plants are oversized in an early stage to anticipate future capture demand, or not.

4. Discussion

This section provides a discussion on the main limitations of this study as well as recommendations for further research.

4.1. Base case CO_2 emissions and fluctuations in industrial activity

In this study, the CO_2 emissions in the period 2020-2030 were assumed to be similar to current emission levels. In reality, the amount of CO_2 emissions will likely alter as a result of different industrial production rates, energy efficiency improvements and changes in the portfolio of industrial plants located in the Botlek. Also, the CO_2 capture infrastructure configurations were not designed to cope with temporal fluctuations in industrial activity, due to lack of data on the plants' activity during the year. Including industrial activity patterns would render the configurations more expensive because of the need for either larger capture equipment and pipelines or venting part of the CO_2 to the atmosphere during activity peaks. Similarly, during low industrial activity, part of the capture capacity will lay idle, resulting in higher specific capital costs per tonne CO_2. Follow-up research is needed to examine the impact of temporal fluctuations as well as flexibility issues on the performance of the configurations.

4.2. Data uncertainty and techno-economic improvement potential of studied configurations

The sensitivity analysis has shown that the position of Oxy-Hybrid as the most favorable infrastructure configuration is fairly robust to a change in the CO_2 capture performance data. Nevertheless, more detailed analyses and higher data quality on flue gas transport and gas velocities as well as on the improvement potential of the infrastructure configurations is desirable to obtain more accurate results for the techno-economic performance. The performance of flue gas transport depends strongly on the network design as well as on the velocity and properties of the flue gas. As the specific capital costs for flue gas transport were found to be around 50% lower than indicated by Bureau et al. (2011), these parameters should be examined in further detail. Also, advanced solvents as well as higher oxygen and hydrogen production efficiencies provide opportunities for cost reductions in the short term. The hydrogen production efficiency can be lower than indicated in this study by excluding the PSA step that is conventionally applied in steam reformers to obtain high purity (>99.9%) hydrogen. For the purpose of pre-combustion capture, lower purity hydrogen (95%) will suffice. Data limitations prohibited the analysis of this option. Also, the potential to utilize waste heat from the SR as well as from the recently installed steam pipeline across the Botlek should be investigated in further detail. Although several industrial plants in the Botlek are already integrated to a large extent in terms of energy and mass flows, the advent of CO_2 capture may increase the potential for further integration, both within and among industrial plants. For example, argon coming from the ASU's can be sold to external parties to improve the economics of oxyfuel combustion.

4.3. Physical footprint and other practical issues

An innovative modular approach was used to assess the physical footprint of CO_2 capture equipment. This approach needs to be validated with technology providers. However, even when using a linear relation between equipment capacity and surface area, which results in lower spatial footprints, several industrial plant sites would still have insufficient space available for capture equipment. For several capture components, large variations in input data were observed, which is mainly due to the disparate nature of the data sources. For other components, only limited and aggregated data was available. More data is needed to provide more accurate estimations. Despite these uncertainties, the findings of this study with respect to space limitations on the industrial plant sites will likely remain unaffected. We assumed that the plot spaces in the Botlek will be available for CO_2 capture installations. Although experts expected this to be true, further research is needed to obtain confirmation. The scope for further improvements in spatial footprints (e.g., by using more advanced packings to reduce the absorber diameter (Feron and Jansen, 2002; Gronvold et al., 2005)) should be examined to assess the potential for footprint reductions in the future. Several other issues related to CO_2 capture in industry have not been addressed in this study, such as (production losses due to) retrofit and the impact of CO_2 capture on the reliability of the industrial processes, operational difficulties when using oxygen or hydrogen in *in situ* NGCC plants, plant operators lacking skills to handle CO_2 capture units, and the possibility to switch back to conventional production (without CO_2 capture).

5. Conclusions

This study developed a method to assess the techno-economic performance and spatial footprint of infrastructure configurations for large scale CO_2 capture in industrial zones. A group of sixteen industrial plants in the Dutch industrial Botlek area (CO_2 emissions: 7.1 $MtCO_2/y$) was selected as a case study. The time frame of the study is the period 2020–2030. The configurations differ with respect to CO_2 capture technology (post-, pre-, or oxyfuel combustion), location of capture components (centralized vs. decentralized), local pipeline network, energy supply for CO_2 capture (NGCC(-CHP), or gas-fired boiler and/or electricity import), and whether CO_2 is captured from the energy plant (CC) or released to the atmosphere (vent).

This analysis has shown that centralized/hybrid CO_2 capture infrastructure configurations provide good opportunities to lower the costs for CO_2 capture in industry and circumvent spatial limitations on industrial plants' sites. Oxy-Hybrid (El) with centralized oxygen production and decentralized CO_2 compression was found to be the most cost effective and realistic configuration when applying CO_2 capture to all industrial plants (61€/tCO_2; 5.8 $MtCO_2/y$ avoided), mainly because of the low energy costs compared to the other capture technologies. For oxyfuel combustion, the economic scale effects of centralized oxygen production units outweigh higher costs for local oxygen and CO_2 transport. The CO_2 treatment and compression units in the Oxy-Hybrid (El) configuration are small enough to be placed on most plant sites. Yet while Oxy-Hybrid (El) is more cost effective on a large scale, oxygen production and CO_2 capture at plant level is still economically preferable for the first three plants. For post-combustion capture, using a separated absorber-stripper configuration (Post-Recsor (CHP/CC): 73€/tCO_2; 7.1 $MtCO_2/y$ avoided) is preferable both from an economic and spatial perspective. These performance figures include economic and CO_2 emission credits for excess electricity sale in the CO_2 avoidance costs. When including these credits, Post-Recsor (CHP/CC) can be more cost effective than Oxy-Hybrid (El) depending on the energy prices. Although cost effective, a fully centralized post-combustion configuration is unrealistic due to the high spatial footprint of large diameter flue gas ducting running across the Botlek. The performance of the pre-combustion configurations depends strongly on the availability of waste heat (WH) from the steam reformer for solvent regeneration. Pre-Central WH (El) shows lower avoidance costs (81€/tCO_2) than Post-Recsor (CHP/CC) (93€/tCO_2) in case

electricity credits are excluded. However, the CO_2 emission reduction potential (4.4 $MtCO_2/y$) is considerably lower than for the post- and oxyfuel configurations as not all fuels in the industrial plants can be replaced with hydrogen. Pre-Central WH (El) can be more cost effective than Oxy-Hybrid (El) when varying several input parameters; yet, the combinations of input parameters rendering Pre-Central WH (El) the configuration with the lowest avoidance cost are rather unlikely.

For all capture technologies, the optimal energy supply mode in terms of costs depends on the energy prices and whether or not economic and CO_2 emission credits for electricity sale are taken into account. In general, when there is both heat and electricity demand for CO_2 capture, using a CHP plant is most cost effective, whereas electricity import is preferable when only electricity is required. Capturing CO_2 from the energy plant shows often lower avoidance costs than venting the CO_2 emissions to the atmosphere.

Although our approach regarding spatial footprints requires expert validation and more reliable data, the findings in this study are not expected to change. Building an extensive local pipeline network across the Botlek was found to be technically and legally feasible, albeit limited space availability in the designated pipeline strips would make several detours and higher operating pressures necessary, which increase local transport costs. However, these costs are minor compared to the total capture costs. For each industrial area, the optimal local transport network will depend on local conditions, not only related to space, but also to safety and environmental regulations. Considering the large advantages of centralized/hybrid configurations compared to decentral CO_2 capture, the findings of this study are expected to apply to other industrial areas as well, especially areas with small industrial plants as these benefit most from economic scale effects of centralized capture units and energy plants. Oxy-Hybrid (El) is the preferable configuration from an economic and spatial perspective. Furthermore, the avoidance cost of Oxy-Hybrid (El) shows a higher robustness to changing energy prices than the other CO_2 capture configurations, which poses lower risks regarding costs. However, alternative industrial areas should be examined to validate this study's findings. The method developed in this study has proven useful for such an analysis and provides valuable lessons for the application of CCS in industrial zones.

Finally, more insight is needed on how to deploy these central/hybrid capture infrastructure configurations in an optimal manner over time. In part B of this analysis, several deployment pathways for Post-Recsor (CHP/CC) and Oxy-Hybrid (El) are investigated, which differ *inter alia* regarding the sequence in which industrial plants start capturing CO_2, the number of deployment steps, and whether CO_2 capture units and energy plants are oversized in an early stage to anticipate future capture demand, or not.

Acknowledgement

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Appendix A.

This appendix presents information on the material selection and technical specifications of pipelines. More detailed information can be found in the detailed pipeline studies underlying this paper (Pipeliner, 2012a,b,c).

A-I Pipeline materials

An analysis was made on the material selections for the pipelines, including all relevant aspects such as corrosion, embrittlement, mechanical stress, possible ruptures, the sound barrier of moving gases, and possibly impingement. The selected pipeline materials are discussed below. An external coating and cathodic protection are applied to all pipelines to prevent corrosion. Polypropylene is used for pipelines that are installed via the so-called pipe ramming method, whereas polyethylene is used in other cases (see Pipeliner, 2012a,b,c).

Flue gas ductwork

The conditions of the flue gas are typically similar to the outlet gas of a flue gas desulphurization (FGD) unit (Billingham et al., 2012). A wide range of corrosion resistant alloys have been used and proposed for FGD ducting, depending on the specific requirements regarding temperature, stress, welding and corrosion resistance. A typical stainless steel duct material (AISI 304) was assumed in this study.

Amine pipelines

The composition, temperature, pressure, density and flow rates of the CO_2-rich and CO_2-lean amine flows were based on Fischer et al. (2005), who modelled a post-combustion capture system using MEA. The use of amine systems is known to cause significant problems in terms of pipeline corrosion (Chakravarti et al., 2001), which will be enhanced by the presence of CO_2. The addition of bases to lower the acidity of the solution has not shown to be an effective measure (Cummings et al., 2005). Carbon steel seems only to be an option if the velocity of the gas remains below $\pm$0.5 m/s; the intactness of the formed oxidized layer, which protects the pipeline wall against corrosion, cannot be guaranteed above this velocity. However, such a low velocity would result in a too large diameter to fit in the designated pipeline strips that still have space for pipelines. The use of a coating on the inside of the pipelines was not feasible for the small diameters considered in this study. On top of that, damage to the coating layer could still result in corrosion (Pipeliner, 2012a). Other pipeline materials considered in this study are fiberglass, specially treated polyethylene and concrete. However, these options were technically infeasible as fiberglass pipelines cannot undergo the drilling techniques required to install the pipelines in the Botlek area, polyethylene pipelines lose strength at relatively high temperatures (the temperatures of the amine solutions are nearly 60 °C), and concrete pipelines cannot be used for the required pressures in this study. Therefore, stainless steel (AISI 304) was used instead.

Oxygen pipelines

Gaseous oxygen transported by pipeline generally contains negligible amounts of water and, thus, no special precautions against corrosion are required (EIGA, 2012). It is, therefore, in principle not a problem to use carbon steel as a pipeline material (EIGA, 2012). The issue with pure oxygen in a carbon steel pipe is, however, not corrosion per se but several concerns such as contamination of the oxygen, plugging of valves and connections due to iron oxide and debris, and a possible combustion and/or explosion if the pipeline is not sufficiently clean (NASA, 1996). Stainless steel is recommended over carbon steel though stainless steel is still flammable in pure oxygen (NASA, 1996). The use of more expensive stainless steel may therefore be a more reasonable choice. The use of polymer pipelines was also investigated. However, as polymer has a lower ignition temperature than steel and fire would propagate rather

fast along the length of the pipeline, stainless steel (AISI 304) was assumed for oxygen transport pipelines.

Hydrogen pipelines

Gaseous hydrogen is usually transported in carbon steel pipelines (EIGA, 2004). The most important criteria for the material selection are stress corrosion, cracking, and embrittlement (EIGA, 2004). Low strength carbon steel grades (e.g., X52) show resistance to hydrogen embrittlement. Polymer pipelines were not considered suitable given the hazardous nature and high operational pressures of hydrogen gas. Due to the small size of hydrogen molecules, seamless steel pipes and specific welding techniques are recommended for hydrogen transport to prevent leakage. A carbon steel grade of X52 was used in this study.

CO_2 pipelines

The technical specifications of the CO_2 streams were such that no significant corrosion would occur in carbon steel pipelines (Ramírez et al., 2011). Given the low level of impurities after the CO_2 treatment step (<10 ppmv SO_x, <10 ppmv NO_x, no free water), carbon steel (X60 or X70) can be used for both low and high pressure pipelines, as this is more economical than stainless steel (Ramírez et al., 2011).

A-II Formulae pipeline calculations

The amount of gas is determined using the ideal gas law:

$$P \times V = n \times R \times T \tag{8}$$

where P is pressure (Pa), V is volume (m^3), n is the amount of gas (mole), T is the temperature (K) and R is the gas constant ($8.314\,J \times mol^{-1} \times K^{-1}$).

The density of the gas ρ (kg/m^3) can be derived from the following formula:

$$\rho = \frac{P \times M}{R \times T \times 1000} \tag{9}$$

where M is the molar mass (flue gas: 29.4 g/mol; amine solution; 25.1 g/mol; O_2: 32.0 g/mol; H_2: 2.0 g/mol; CO_2: 44.0 g/mol) (Fischer et al., 2005).

The dynamic viscosity $_{\mu ref}$ ($\mu Pa \times s$) of the gas at the input temperature T (K) is determined using the formula of Sutherland:

$$\mu = \mu_{ref} \times \left(\frac{T}{T_{ref}}\right)^{3/2} \times \frac{T_{ref} + S}{T + S} \tag{10}$$

where $_{\mu ref}$ is the reference viscosity (O_2: 20.18 $\mu Pa \times s$; H_2: 8.76 $\mu Pa \times s$; CO_2: 14.8 $\mu Pa \times s$ (LMNO, 2014)) at temperature T_{ref} (O_2: 292.25 K; H_2: 293.85 K; CO_2: 293.15 K (LMNO, 2014)), and S is the Sutherland's constant for the gaseous material (O_2: 127 K; H_2: 72 K; CO_2: 240 K (LMNO, 2014))[5].

Subsequently, the kinematic viscosity v (m^2/s) is determined using the following formula:

$$v = \frac{\mu}{\rho} \tag{11}$$

The Reynolds number Re (−) can be calculated as follows:

$$Re = \frac{\rho \times v \times D_i}{\mu} \tag{12}$$

[5] The dynamic viscosity of flue gas (17 $\mu Pa \times s$) was determined by using the mole based average of the dynamic viscosities of its components (CO_2, N_2, O_2, H_2O).

where v is the mean velocity of the gas (m/s) and D_i is the inner diameter of the pipe (m).

To limit the risk of fire with oxygen transport, the maximum velocity (10 m/s at a pressure of 40 bar) was taken from EIGA (2012).

The mean velocity of the gas v is a function of the flow rate Q (m^3/s) and the inner diameter.

$$v = \frac{Q}{1/4 \times \pi \times D_i^2} \tag{13}$$

For a Reynolds number that is larger than 4000, the Darcy friction factor f(−) can be determined using the Swamee–Jain equation:

$$f = \frac{0.25}{\left[\log_{10} \times \left(\frac{\epsilon}{3.7 \times D_i} + \frac{5.74}{Re^{0.9}}\right)\right]^2} \tag{14}$$

where ϵ is the roughness height (m). A value of 0.001 meter was assumed for both carbon and stainless steel.

For a Reynolds number that is smaller than 2000, the Darcy friction factor f can be calculated as follows:

$$f = \frac{64}{Re} \tag{15}$$

The pressure drop in a pipeline ΔP (Pa) was calculated using the following formula:

$$\Delta P = f \times \rho \times \frac{L}{D_i} \times \frac{v^2}{2} \tag{16}$$

where L is the pipeline distance (m).

When calculating each pipeline within the local network, the pressure drop and change in density of the previous pipelines was taken into account as well.

The design pressure of the pipelines is assumed to be 10% higher than the operational inlet pressure in order to give some operational freedom (Knoope et al., 2014).

Once the pressures in the pipelines are known, the material-diameter-wall thickness combinations of the pipelines are determined:

$$\sigma_p = \gamma_p \times \frac{P_d \times D_{ao}}{2 \times d_{min}} \leq \frac{R_{eb}}{\gamma_m} \tag{17}$$

where σ_p is the maximum working stress that material can tolerate (MPa), γ_p is the (adapted) partial factor on the internal pressure (1.88) (Municipality Rotterdam, 2010), γ_m is the partial material factor (1.1) (NEN, 2012a), P_d is the design pressure (MPa), D_{ao} is the average outside diameter of the pipeline (10^{-3} m), d_{min} is the minimum wall thickness (10^{-3} m), and R_{eb} is the yield strength of the pipeline material ($N/10^{-3}m^2$).

The tolerance in the pipeline wall thickness and the corrosion buffer is calculated as follows:

$$d_{min} = \frac{d}{(1 + ft)} + C \tag{18}$$

where d is the calculated pipeline wall thickness (10^{-3} m), ft is the fabrication tolerance (12.5%, assumption), and c is the corrosion buffer c (assumed to be 1.0×10^{-3} m).

A-III Power for gas compression and pumping

The power requirement for blowers and compressors to route the gaseous flue gas, oxygen and CO_2 though the ductwork and pipelines is calculated using the following formula (adapted from Damen et al., 2007b):

$$C_c = \frac{Z \times N \times R \times T_1 \times k}{M_{i\times}(k-1)} \times \left[\left(\frac{p_2}{p_1}\right)^{\frac{(k-1)}{N \times k}} - 1\right] \times \frac{m_i}{\eta_{is} \times \eta_m} \tag{19}$$

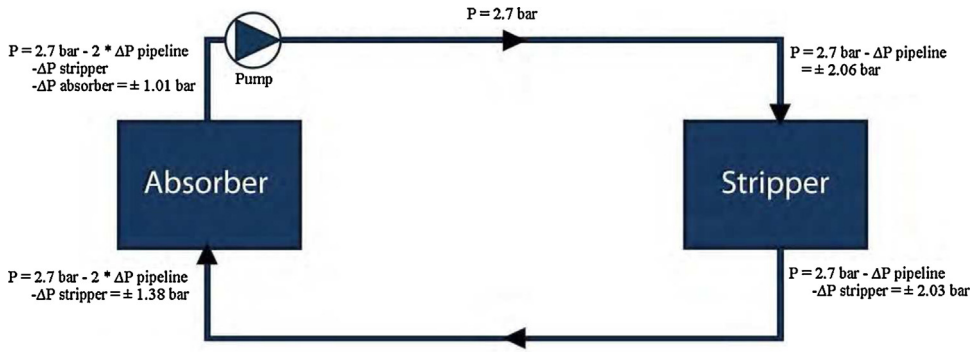

Fig. 16. Schematic overview of amine recirculation system for a design pressure of 6 bar and the associated pressure drops in the absorber, stripper and amine pipelines.

where C_c is the compression power (kW), Z is the compressibility factor (H_2: 1.001; O_2: 1; CO_2: 0.625)[6], N is the number of compressor/blower stages (compressor: 4, blower: 1), R is the universal gas constant (8.3145 J/(mol × K), T_1 is the suction temperature (H_2: 293 K; O_2: 303 K (IEA GHG, 2000); flue gas[7]: 328 K; CO_2: 308 K (Fischer et al., 2005)), k is the specific heat ratio (H_2: 1.38; O_2: 1.39; flue gas[8]: 1.39; CO_2: 1.29 (Air Liquide, 2014)), M_i is the molar mass of gas i (flue gas: 29.4 g/mol (Fischer et al., 2005); amine solution; 25.1 g/mole; O_2: 32.0 g/mol; H_2: 2.0 g/mol; CO_2: 44.0 g/mol), p_1 is the suction pressure (H_2: 15×10^5 Pa (IEA GHG, 2000); O_2: 1.7×10^5 Pa (Allam et al., 2005a); CO_2: 1.92×10^5 Pa (Fischer et al., 2005); flue gas: 1.01×10^5 Pa (Fischer et al., 2005)), $p2$ is the discharge pressure (Pa), mi is the mass flow rate of gas i (kg/s), $\eta$is is the isentropic efficiency (75%), ηm is the mechanical efficiency (90%).

The specific energy requirement (SER) for the four stage O_2 compression from 1.38 to 30 bar (0.35 GJ_e/tO_2) is based on a vendor quote from a compressor technology provider (Dresser-Rand, 2015). The power requirement for pumps to circulate the amine solutions between the absorbers and strippers is as follows:

$$P = Q \times \rho \times g \times H \tag{20}$$

where P is the pump power (kW), Q is the flow rate (m^3/s), ρ is the density of the amine solution (953 kg/m^3 for rich amine solution and 958 kg/m^3 for lean amine solution (Fischer et al., 2005)), g is the gravitational constant (9.81 m/s^2), and H is the head difference (m). An efficiency of 65% was assumed to obtain the SER for pumping.

Fig. 16 shows the amine recirculation system with the pressure drops in the supply and retour pipelines as well as in the absorber and stripper. The pressure drop in both the absorber and stripper was assumed to be 0.1 bar/660 m^3 for a packing consisting of cascade mini rings (Fischer et al., 2005).

A-IV Costs for pipelines, blowers and compressors

The pipeline material costs were obtained using the following Formula:

$$C_m = \left(\frac{\pi D_i \times L \times d \times \rho_s \times P_s}{0.95} + \pi D_o \times L \times P_c\right) \times 1.1 \times 1.2 \tag{21}$$

where D_i is the inner pipeline diameter (m), D_o the outer pipeline diameter (m), L the pipeline length (m), d the wall thickness (m), ρ_s the density of steel (7800 kg/m^3, assumption), Psthe steel price (€/kg), and Pc the price for coating (€/m^2). Specific cost factors were applied to account for cutting losses (0.95), flanges, junctions, appendages, etc. (1.2) and valves (1.1).

The construction costs Cc were derived as follows:

$$C_c = C_p \times OD \times L \tag{22}$$

where C_p is the specific construction cost (€/mL/mD), which contain inter alia costs for excavation, welding, labor and removal of the pipeline at the end of the lifetime.

The annual O&M costs are assumed to be 2% of the capital cost. The material and construction costs for flue gas ducts were assessed using a standard specific cost factor (see Table 8).For the pre-combustion case, an additional natural gas pipeline and receiving station has to be installed for the SR hydrogen plant with a flow rate capacity of 3.2×10^4 N m^3/h. The total costs were based on an official offer made by the Dutch natural gas company (Gasunie, 2012) and are presented in Table 8. More detailed information on the technical specifications and costs can be found in (Pipeliner, 2012c).

The blower capital costs Cblower were determined by using an equation function of the gas flow rate Q (m^3/s) and power capacity P (kW), which was derived by using data from process simulation software AspenPlus® in the COCATE project (Bureau et al., 2011).

$$C_{blower} = \frac{1413 \times e^{0.325 \times \ln(P \times Q \times 3600)}}{10^6} \tag{23}$$

The compressor capital costs for CO_2 compression to 110 bar are presented in Tables 3–5. The capital cost for the four stage compressor trains to boost the O_2 from 1.38 to 30 bar (22.6 M€/MtO_2/y) is based on a vendor quote from a compressor technology provider (Dresser-Rand, 2015).

[6] Compressibility Factor H_2 and O_2 at 1.013 bar and 15 °C (Air Liquide, 2014); average compressibility factor CO_2: [Z(Pin: 1 bar) + Z(Pout: 150 bar)]/2 = 0.625 at 25–40 °C (Mccollum and Ogden, 2006). The compressibility factor changes with the addition of impurities. For example, with 1.5% SO_2 added, the compressibility factor of CO_2 changes by 2–5% (Mccollum and Ogden, 2006). As it was difficult to find compressibility factors for non-pure CO_2 streams, the value for pure CO_2 was used in this study.

[7] Temperature of flue gas exiting the direct contact cooler.

[8] The specific heat ratio (k) for flue gas is calculated to be 1.39 for a composition of 10 mol% CO_2, 3 mol% O_2 and 86 mol% N_2 and 1 mol% Ar at 1.03 bar and 25 °C.

Appendix B.

See Tables 13–15.

Table 13
Techno-economic performance of post-combustion configurations in the Botlek.

	Unit	Boiler				CHP				
		Decentral		Central		Decentral		Central		Recsor
		Vent	CC	Vent	CC	Vent	CC	Vent	CC	CC
Technical performance										
Annual CO_2 produced base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1
Annual CO_2 captured (without CC energy plant)	Mt/y	6.4	6.4	6.4	6.4	9.1	6.4	6.4	6.4	6.4
Heat for capture	TJ_{th}/y	23.2	23.6	23.2	23.6	23.2	29.1	23.2	28.5	28.5
Electricity for capture	TJ_e/y	4.5	6.2	7.1	8.8	5.4	7.6	7.1	10.8	9.6
Energy plant performance										
Heat output	MW	804	820	804	820	804	1009	804	989	989
Electrical output	MW	0	0	0	0	489	612	613	754	754
Fuel input	MW	946	964	946	964	1357	1801	1575	1937	1937
Excess electrical output	MW	0	0	0	0	331	455	366	507	546
Energy plant emissions										
CO_2 generated	Mt/y	1.5	1.6	1.5	1.6	2.2	2.9	2.6	3.2	3.2
CO_2 captured	Mt/y	0.0	1.4	0.0	1.4	0.0	2.6	0.0	2.8	2.8
Remaining emissions	Mt/y	1.5	0.2	1.5	0.2	2.2	0.3	2.6	0.3	0.3
CO_2 emissions										
Total annual CO_2 generated	Mt/y	8.7	8.7	8.7	8.7	9.3	10.1	9.7	10.3	10.3
Total annual CO_2 captured	Mt/y	6.4	7.8	6.4	7.8	9.1	9.1	6.4	9.3	9.3
Emissions from imported electricity	Mt/y	0.4	0.4	0.4	0.6	-0.6	-0.8	-0.7	-0.9	-1.0
Total annual CO_2 emissions	Mt/y	2.5	1.3	2.7	1.4	2.3	0.2	2.6	0.1	0.0
Total CO_2 emissions avoided (with elec. sale)	Mt/y	n/a	n/a	n/a	n/a	4.7	6.9	4.5	7.0	7.1
Total CO_2 emissions avoided (w/o elec. sale)	Mt/y	4.6	5.9	4.4	5.7	4.5	6.1	3.8	6.1	6.1
Number of units										
Absorbers	–	18	19	7	8	18	19	7	10	11
Stripper	–	16	16	2	3	16	16	2	3	4
CO_2 compressors	–	16	16	3	3	16	16	3	4	4
Boilers/NGCC-CHP	–	16	16	1	1	16	16	1	1	2
Economic performance										
CO_2 capture										
CAPEX										
Modification to stacks	M€	7	7	7	7	7	7	7	7	7
SCR/FGD units	M€	93	107	93	107	93	119	93	121	121
Absorbers	M€	229	245	159	188	229	280	159	221	232
Strippers	M€	107	123	60	75	107	140	60	84	105
CO_2 compressors & treatment units	M€	87	100	53	61	87	114	53	73	78
Total process plant costs (PPC)	M€	524	582	372	438	524	660	372	506	543
Total plant cost (TPC)	M€	681	757	484	569	681	858	484	658	706
Total capital requirement (TCR)	M€	750	832	533	626	750	943	533	723	776
OPEX	M€/yr	51	62	51	62	72	72	51	73	73
Energy plant										
CAPEX	M€	58	59	21	21	972	972	824	931	813
OPEX	M€/yr	1	1	0	0	39	39	33	37	33
Flue gas ducting										
Length	km	7.1	7.1	14.4	14.4	7.1	7.1	14.4	14.4	7.1
CAPEX	M€	35.8	36.7	74.6	74.6	35.8	36.7	74.6	74.6	36.7
Capacity blowers	MW	49	49	98	98	49	49	98	98	49
OPEX	M€/yr	0.9	0.9	1.9	1.9	0.9	0.9	1.9	1.9	0.9
Amine pipelines (4 bar)										
Length	km	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	14.6
CAPEX	M€	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	24.2
Capacity pumps	MW	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1
OPEX	M€/yr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.7
CO_2 pipeline (110 bar)										
Length	km	24.9	24.9	1.2	1.2	24.9	24.9	1.2	1.2	1.2
CAPEX	M€	10.2	1.3	1.1	1.3	10.2	11.5	1.1	1.6	1.6
Capacity compressors	MW	123	150	123	150	123	173	123	177	177
OPEX	M€/yr	0.3	0.3	0.0	0.0	0.3	0.3	0.0	0.1	0.1
Total costs										
CAPEX	M€	809	939	629	722	1964	1964	1432	1731	1652
OPEX	M€/yr	52	63	53	64	111	111	86	113	109

Table 13 (*Continued*)

	Unit	Boiler				CHP				
		Decentral		Central		Decentral		Central		Recsor
		Vent	CC	Vent	CC	Vent	CC	Vent	CC	CC
Energy costs										
Natural gas	M€/yr	272	278	272	278	519	519	453	558	558
Electricity	M€/yr	100	136	214	200	-288	-288	-232	-321	-346
Total (with credits for electricity sale)		n/a	n/a	n/a	n/a	181	231	222	237	212
Total (without credits for electricity sale)	M€/yr	372	414	486	478	209	318	256	284	262
Average CO_2 avoidance cost (with elec. sale)	€/tCO_2	n/a	n/a	n/a	n/a	96	83	106	79	73
Average CO_2 avoidance cost (w/o elec. sale)	€/tCO_2	113	100	139	110	106	108	133	91	93
Physical footprint										
Scrubbers	$\cdot 10^3$ m^2	10.0	26.2	4.9	5.6	10.0	29.9	4.9	6.2	6.5
Strippers	$\cdot 10^3$ m^2	13.2	15.2	6.1	7.0	13.2	16.8	6.1	7.9	7.9
CO_2 compressors & treatment units	$\cdot 10^3$ m^2	9.6	11.0	5.9	6.7	9.6	12.6	5.9	8.0	8.0
Energy plant (Boiler/NGCC)	$\cdot 10^3$ m^2	1.1	1.1	0.5	0.5	83.7	172.9	84.3	96.9	96.9
Total	$\cdot 10^3$ m^2	34	53	17	20	116	232	101	119	119

Table 14
Techno-economic performance of oxyfuel combustion configurations in the Botlek.

	Unit	Decentral ASU			Central ASU				Central ASU			
		Decentral CO_2 compression			Decentral CO_2 compression				Central CO_2 compression			
		El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC
Technical performance												
Annual CO_2 produced base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1
Annual CO_2 produced oxyfuel combustion	Mt/y	6.8	6.8	6.8	6.8	6.8	6.8	6.8	6.8	6.8	6.8	6.8
Annual CO_2 captured (without CC energy plant)	Mt/y	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1
Heat for capture	TJ_{th}/y	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Electricity for capture	TJ_e/y	8.0	8.0	8.0	7.0	8.0	8.0	8.0	7.0	8.0	8.0	8.0
Oxygen production	Kt/d	12	12	14	12	12	12	12	12	12	12	14
Air Separation Units	MW_e	93	93	110	93	93	93	93	93	93	93	108
Energy plant performance												
Electrical output	MW	0	277	288	0	0	277	287	0	0	277	287
Fuel input	MW	0	576	600	0	0	509	527	0	0	509	527
Energy plant emissions												
CO_2 generated	Mt/y	0.0	0.9	1.0	0.0	0.0	0.8	0.9	0.0	0.0	0.8	0.9
CO_2 captured	Mt/y	0.0	0.0	0.9	0.0	0.0	0.0	0.8	0.0	0.0	0.0	0.8
Remaining emissions	Mt/y	0.0	0.9	0.1	0.0	0.0	0.8	0.1	0.0	0.0	0.8	0.1
CO_2 emissions												
Decrease emissions industrial plants	Mt/y	0.0	0.0	0.0	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total annual CO_2 generated	Mt/y	6.8	7.7	8.1	6.8	6.8	6.8	7.6	6.8	6.8	7.6	7.6
Total annual CO_2 captured	Mt/y	6.1	6.1	7.0	6.1	6.1	6.1	6.9	6.1	6.1	6.1	6.9
Emissions from imported electricity	Mt/y	0.6	0.0	0.0	0.5	0.6	0.0	0.0	0.5	0.6	0.0	0.0
Total annual CO_2 emissions	Mt/y	1.2	0.7	0.7	1.2	1.3	0.7	0.7	1.2	1.3	0.7	0.7
Total CO_2 emissions avoided	Mt/y	5.9	5.5	6.3	5.8	5.8	5.6	6.4	5.9	5.8	5.6	6.4
Number of units												
Air Separation Units	–	13	13	16	2	2	2	2	2	2	2	3
CO_2 compressors	–	16	16	16	16	16	16	16	3	3	3	3
NGCC plants	–	0	16	16	0	0	1	1	0	0	1	1
Economic performance												
CO_2 capture												
CAPEX												
Furnace modifications	M€	5	5	5	5	4	5	5	5	5	5	5
Boiler modifications	M€	11	11	11	11	11	11	11	11	11	11	11
Air separation units	M€	510	510	590	300	300	300	332	300	300	300	367
Gas gathering system	M€	43	43	49	43	43	43	43	43	43	43	43
Cooling system	M€	74	83	91	74	83	74	74	74	74	74	74
Scrubbers	M€	33	33	33	33	33	33	33	33	33	33	33
Strippers	M€	16	16	16	16	16	16	16	16	16	16	16
CO_2 compressors & treatment units	M€	168	168	185	168	168	170	182	103	103	103	111
Total process plant costs (PPC)	M€	859	868	979	650	658	652	695	585	585	585	660
Total plant cost (TPC)	M€	1117	1129	1272	845	856	847	904	760	760	760	858
Total capital requirement (TCR)	M€	1229	1242	1400	930	942	932	994	836	836	836	944
OPEX	M€/yr	55	55	63	55	55	55	62	55	55	55	62

Table 14 (*Continued*)

	Unit	Decentral ASU			Central ASU				Central ASU			
		Decentral CO_2 compression			Decentral CO_2 compression				Central CO_2 compression			
		El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC	El LP O_2	El MP O_2	NGCC/vent	NGCC/CC
Energy plant												
CAPEX	M€	0	359	371	0	0	246	252	0	0	246	252
OPEX	M€/yr	0	14	15	0	0	10	10	0	0	10	10
Oxygen pipeline (2 and 30 bar)												
Length	km	8.5	8.5	8.5	20.0	20.0	20.0	20.0	20.0	20.0	20.0	20.0
CAPEX	M€	107	100	117	72	115	88	131	72	115	115	131
Capacity compressors/blowers	MW	47	47	47	12	47	47	47	12	47	47	47
OPEX	M€/yr	2.7	2.5	2.9	1.8	2.9	2.2	3.3	1.8	2.9	2.9	3.3
CO_2 pipeline (2 and 110 bar)												
Length (2 bar pipelines)	km	0.0	0.0	0.0	0.0	0.0	0.0	0.0	23.7	23.7	23.7	23.7
Length (110 bar pipelines)	km	24.9	24.9	24.9	24.9	24.9	24.9	24.9	1.2	1.2	1.2	1.2
CAPEX	M€	11.4	11.4	11.9	11.8	11.8	11.8	12.5	66.0	66.0	66.0	70.3
Capacity compressors/blowers	MW	112	112	128	112	112	112	126	5	5	5	5
OPEX	M€/yr	0.3	0.3	0.3	0.3	0.3	0.3	0.3	1.6	1.6	1.6	1.7
Total costs												
CAPEX	M€	1347	1712	1900	1014	1068	1278	1390	974	1017	1263	1397
OPEX	M€/yr	58	72	81	57	58	67	76	59	60	69	77
Energy costs												
Natural gas	M€/yr	0	166	173	0	0	147	152	0	0	147	152
Electricity	M€/yr	175	0	0	153	175	0	0	153	175	0	0
Total	M€/yr	175	166	173	153	175	147	152	153	175	147	152
Average CO_2 avoidance cost	€/tCO_2	66	80	75	56	61	65	61	55	61	65	62
Physical footprint												
NGCC plants	$\cdot 10^3$ m^2	0	29	30	0	0	14	15	0	0	14	15
Scrubbers	$\cdot 10^3$ m^2	29	29	34	18	18	18	20	18	18	18	22
Strippers	$\cdot 10^3$ m^2	2	2	2	2	2	2	2	2	2	2	2
Air separation units	$\cdot 10^3$ m^2	3	3	3	3	3	3	3	3	3	3	3
CO_2 compressors & treatment units	$\cdot 10^3$ m^2	16	16	18	16	16	16	17	10	10	10	11
Total	$\cdot 10^3$ m^2	50	75	82	34	39	48	52	33	33	47	52

Table 15
Techno-economic performance of pre-combustion configurations in the Botlek.

	Unit	Boiler			NGCC(-CHP)			
		No waste heat		Waste heat	No waste heat		Waste heat	
		Vent	CC	El	Vent	CC	Vent	CC
Technical performance								
Annual CO_2 produced base case	Mt/y	7.1	7.1	7.1	7.1	7.1	7.1	7.1
Annual CO_2 base eligible for capture	Mt/y	4.9	4.9	4.9	4.9	4.9	4.9	4.9
Hydrogen production per day	kt/d	1.2	1.2	1.2	1.2	1.2	1.2	1.2
Heat for capture	TJ_{th}/y	11.7	11.9	11.9	11.7	14.4	11.7	14.4
Electricity for capture	TJ_e/y	2.5	3.3	2.5	2.5	4.2	2.5	2.8
Energy plant performance								
Heat output	MW	405	413	0	405	499	0	0
Electrical output	MW	0	0	0	286	352	88	108
Fuel input	MW	477	486	0	768	946	174	219
Excess electrical output	MW	0	0	0	198	264	0	0
Energy plant emissions								
CO_2 generated	Mt/y	0.8	0.8	0	1.3	1.5	0.3	0.4
CO_2 captured	Mt/y	0	0.7	0	0.0	1.4	0.0	0.3
Remaining emissions	Mt/y	0.8	0.1	0	1.3	0.2	0.3	0.0
CO_2 emissions								
Total annual CO_2 generated	Mt/y	8.0	8.0	7.2	8.4	8.7	7.5	7.8
Total annual CO_2 captured	Mt/y	4.9	5.5	4.9	4.9	6.2	4.9	5.2
Emissions from imported electricity	Mt/y	0.2	0.2	0.2	-0.4	-0.5	0.0	0.0
Total annual CO_2 emissions	Mt/y	3.3	2.6	2.5	3.2	2.0	2.6	2.6
Total CO_2 emissions avoided (with elec. sale)	Mt/y	n/a	n/a	n/a	3.7	4.9	n/a	n/a
Total CO_2 emissions avoided (w/o elec. sale)	Mt/y	3.5	4.3	4.4	3.3	4.4	4.3	4.5
Number of units								
Steam reformer	–	1	1	1	1	1	1	1
Absorbers process gas	–	4	4	4	4	4	4	4
Stripper process gas	–	2	2	2	2	2	2	2
Absorbers flue gas	–	2	3	2	2	3	2	2
Stripper flue gas	–	1	1	1	1	1	1	1
CO_2 compressors	–	2	3	2	2	3	2	2
Boilers/NGCC(-CHP)	–	1	1	1	1	1	1	1

Table 15 (*Continued*)

	Unit	Boiler			NGCC(-CHP)			
		No waste heat		Waste heat	No waste heat		Waste heat	
		Vent	CC	El	Vent	CC	Vent	CC
Economic performance								
CO_2 capture								
CAPEX								
Steam reformer	M€	251	251	251	251	251	251	251
Modification to burners	M€	0.5	0.6	0	0	1	0	1
SCR/FGD units	M€	17	22	17	17	27	17	19
Absorbers process gas	M€	15	15	15	15	15	15	15
Strippers process gas	M€	10	10	10	10	10	10	10
Absorbers flue gas	M€	44	63	44	44	76	44	50
Strippers flue gas	M€	19	24	19	19	30	19	21
CO_2 compressors & treatment units	M€	83	99	83	83	107	83	86
Total process plant costs (PPC)	M€	440	486	440	440	517	440	455
Total plant cost (TPC)	M€	572	632	572	572	673	572	591
Total capital requirement (TCR)	M€	629	695	629	629	740	629	650
OPEX	M€/yr	56	64	56	56	72	56	60
Energy plant								
CAPEX	M€	14	21	0	252	291	105	123
OPEX	M€/yr	0.3	0.4	0	10.1	11.6	4.2	4.9
Natural gas pipeline (40 bar)								
Length	km	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CAPEX	M€	1.3	1.6	0.7	1.4	1.8	0.9	1.0
OPEX	M€/yr	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Hydrogen pipeline (15 bar)								
Length	km	12.2	12.2	12.2	12.2	12.2	12.2	12.2
CAPEX	M€	20	20	20	20	20	20	20
OPEX	M€/yr	0.5	0.5	0	0.5	0.5	0.5	0.5
CO_2 pipeline (110 bar)								
Length	km	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CAPEX	M€	0.8	1.1	0.8	1.1	1.1	0.8	0.9
Capacity compressors	MW	73	80	73	73	86	73	76
OPEX	M€/yr	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total costs								
CAPEX	M€	665	739	650	903	1054	755	795
OPEX	M€/yr	57	65	57	67	84	61	65
Energy costs								
Natural gas	M€/yr	309	312	172	393	444	222	233
Steam export	M€/yr	0	0	-2	0	0	-2	-2
Electricity	M€/yr	56	74	56	-125	-167	0	0
Total (with credits for electricity sale)	M€/yr	n/a	n/a	n/a	268	277	n/a	n/a
Total (without credits for electricity sale)	M€/yr	365	385	226	284	302	220	233
Average CO_2 avoidance cost (with elec. sale)	€/tCO_2	n/a	n/a	n/a	120	99	n/a	n/a
Average CO_2 avoidance cost (w/o elec. sale)	€/tCO_2	137	126	81	137	115	86	86
Physical footprint								
Steam reformer	$\cdot 10^3$ m^2	9.3	9.3	9.3	9.3	9.3	9.3	9.3
Scrubbers	$\cdot 10^3$ m^2	4.0	4.5	4.0	4.0	4.2	4.0	4.2
Strippers	$\cdot 10^3$ m^2	5.1	5.6	5.1	5.1	6.1	5.1	5.3
CO_2 compressors & treatment units	$\cdot 10^3$ m^2	4.4	5.3	4.4	4.4	5.8	4.4	4.6
Energy plant (Boiler/NGCC(-CHP))	$\cdot 10^3$ m^2	0.3	0.3	0.0	19.1	21.9	6.6	7.5
Total	$\cdot 10^3$ m^2	23	25	23	42	47	30	31

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