Grant Agreement Number:
641185

Action acronym:
CEMCAP

Action full title:
CO_2 capture from cement production

Type of action:

H2020-LCE-2014-2015/H2020-LCE-2014-1

Starting date of the action: 2015-05-01
Duration: 42 months

D4.6
CEMCAP comparative techno-economic analysis of CO_2 capture in cement plants

Due delivery date: 2018-08-31
Actual delivery date: 2018-10-31

Organization name of lead participant for this deliverable:
SINTEF Energy Research

Project co-funded by the European Commission within Horizon2020		
Dissemination Level		
PU	Public	x
CO	Confidential, only for members of the consortium	

Deliverable number:	D4.6
Deliverable title:	CEMCAP comparative techno-economic analysis of CO_2 capture in cement plants
Work package:	WP4 Comparative capture process analysis
Lead participant:	SINTEF-ER

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Keywords

CO_2 capture, CCS, cement, clinker, kiln, oxyfuel, chilled ammonia process, membrane, CO_2 liquefaction, calcium looping, techno-economic assessment, SPECCA, cost of clinker, cost of CO_2 avoided.

Abstract

In CEMCAP, four different CO_2 capture technologies for cement plants are investigated: the oxyfuel technology, the chilled ammonia process, membrane-assisted CO_2 liquefaction, and the calcium looping technology. This report presents a techno-economic evaluation of these technologies. Key performance indicators such as specific primary energy consumption for CO_2 avoided (SPECCA), cost of clinker and cost of CO_2 avoided are calculated. The calculations are based on process simulations, with input from experimental work carried out in the project. The SPECCA of the CEMCAP technologies are all lower than 60% of the reference technology (MEA) under the defined base case conditions. The cost of clinker increases with 49-92% compared to the cost of clinker without capture. The cost of CO_2 avoided is between 42 €/t_{CO2} (oxyfuel process) which is approximately halved compared to MEA, and 84 €/t_{CO2} (MAL process), which is on the same level as MEA. The techno-economic KPIs are strongly dependent on factors such as steam source, electricity mix, electricity price, fuel price, *etc.* For the assessment of a CO_2 capture technology for a specific plant, a more plant specific techno-economic evaluation should be performed.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185. This work was supported by the Swiss State Secretariat for Education, Research and Innovation (SERI) under contract number 15.0160

Please cite this report as: *Voldsund, Mari; Anantharaman, Rahul; Berstad, David; De Lena, Edoardo; Fu, Chao; Gardarsdottir, Stefania; Jamali, Armin; Pérez-Calvo, José-Fransisco; Romano, Matteo; Roussanaly, Simon; Ruppert, Johannes; Stallmann, Olaf; Sutter, Daniel, 2018.* CEMCAP comparative techno-economic analysis of CO_2 capture in cement plants *(D4.6)*.
Refer to the CEMCAP community in Zenodo.org for citation with DOI.

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1 INTRODUCTION

This document presents a techno-economic evaluation of the four CO_2 capture technologies that have been investigated in the CEMCAP project. The investigated technologies are:

- Oxyfuel technology
- Chilled ammonia process (CAP)
- Membrane-assisted CO_2 liquefaction (MAL)
- Calcium looping (CaL)
    - Tail-End configuration
    - Integrated Entrained Flow (EF) configuration

The techno-economic evaluation is based on process simulations of the technologies applied to a reference cement plant. Key performance indicators evaluating the energy and economic performance of the technologies are calculated. Different cases in terms of flue gas composition, extent of CO_2 capture, transport option and steam supply are evaluated. The technologies are compared with MEA-based CO_2 absorption which is chosen as the reference technology due to its benchmark status in the literature.

In Section 2, the methodology and assumptions used for technical and economic evaluation of the CO_2 capture technologies as well as the key performance indicators are presented. The reference cement plant is briefly described in Section 3 and the specifications of selected CO_2 capture cases to be evaluated are presented in Section 4. In Sections 5-10, the principles of the different CO_2 capture technologies are explained, together with a brief description of the process simulations, the main results from the process analysis in terms of process performance, utility and material consumption, and the economic evaluation of the investigated CO_2 capture cases. The results are discussed in Section 11 and a sensitivity analysis of the key performance indicators is presented in Section 12. In Section 13, the outlook for further technology development and process improvement is discussed. Finally, concluding remarks are presented in Section 14.

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2 METHODOLOGY

As a basis for the techno-economic analysis, process evaluations with emphasis on energy and environmental performance are first carried out based on process simulations for each of the technologies applied to a reference cement plant. This is used as input for economic evaluations of the technologies. Key performance indicators (KPIs) are finally calculated and used to compare the techno-economic performance of the technologies. This section presents the method and common assumptions used for the process evaluations, the methodology of the economic evaluations, and the definitions of the KPIs.

2.1 Process evaluation

Process modeling and simulations are performed by different partners in the CEMCAP project. An overview of which partner that is responsible for modelling and simulation of each technology is given in Table 2.1. To ensure consistent simulations, common assumptions for process simulations, e.g. various process unit specifications, are used as defined in the CEMCAP framework [VOL, 2018].

Table 2.1. Project partners responsible for process modelling of CEMCAP technologies.

CO_2 capture technology	Project partner(s) responsible for process simulations
MEA-based CO_2 absorption (reference technology)	SINTEF ER
Oxyfuel	VDZ (core process), SINTEF ER (heat recovery cycle and CO_2 purification unit)
Chilled ammonia process	ETH (core process), SINTEF ER (CO_2 compression/liquefaction)
Membrane-assisted CO_2 liquefaction	SINTEF ER
Calcium looping (tail-end and integrated EF)	PoliMi (core process and heat recovery cycle), SINTEF ER (CO_2 purification unit)

Based on the process modelling results, the energy and environmental performance of the processes are evaluated. Common assumptions used in the process evaluations are summarised in Table 2.2. It should be noted that the electricity generation efficiency and specific CO_2 emissions have a large impact on the results, and the numbers used as basis in CEMCAP are average values for the European electricity mix in 2014. Details on technology specific assumptions and process simulation approaches for the different technologies are presented in Sections 5-10.

Table 2.2. Main assumptions for the process evaluations.

Parameter	Value
Clinker / cement factor	0.737
Coal LHV [kJ/kg]	27,150
Electricity generation efficiency, η_{el} [%]	45.9
Electricity generation specific CO_2 emissions, e_{el} [kg_{CO2}/MWh]	262
Electric heater efficiency [%]	95
Natural gas specific CO_2 emissions [kg_{CO2}/GJ]	56.1
Natural gas boiler efficiency [%]	90
ASU power demand [QUE, 1996] [kWh/t_{O2}]	226
ASU cooling demand [IEA, 2005] [kJ/kg_{O2}]	566
ASU dehydration heat demand [IEA, 2005] [kJ/kg_{O2}]	58.3

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CO_2 dehydration with TEG heat demand [kJ/kg$_{CO2}$]	2.62
CO_2 dehydration with TEG power demand [kJ/kg$_{CO2}$]	0.045
Cooling water system power demand [kW/kW]	0.02

2.2 Economic evaluation

The economic evaluation consists of estimation of two main parts: (i) the capital costs (CAPEX) which is expressed in terms of total plant cost (TPC), and (ii) the operating costs (OPEX). The cost estimation of the investigated CO_2 capture technologies and compilation of the equipment list on which the capital costs are based are carried out by different partners in the CEMCAP project, as listed in Table 2.3.

Table 2.3. Project partners responsible for cost estimation of CEMCAP technologies.

CO_2 capture technology	Project partner(s) responsible for equipment list	Project partner(s) responsible for cost estimation
MEA-based CO_2 absorption (reference technology)	SINTEF ER	SINTEF ER
Oxyfuel	Italcementi, VDZ, SINTEF ER	SINTEF ER
Chilled ammonia process	GE	GE (CAPEX), SINTEF ER (OPEX)
Membrane-assisted CO_2 liquefaction	SINTEF ER	SINTEF ER
Calcium looping (tail-end and integrated EF)	PoliMi (core process and heat recovery), SINTEF ER (CO_2 purification unit)	PoliMi (CAPEX – core process and heat recovery cycle, OPEX), SINTEF ER (CAPEX – CO_2 purification unit)

As detailed in the CEMCAP framework [VOL, 2018], a bottom-up approach is used for estimation of TPC for the CO_2 capture technologies. The cost estimates of all the CO_2 capture technologies are performed for "nth of a kind" plants, i.e. for commercial plants built after initial technology demonstration plants. A breakdown of the costing approach is illustrated in Figure 2.1 while Figure 2.2 gives an overview of the different cost factors, variable and fixed, which make up the total operating costs. An excel model for cost estimation following this approach is developed and used for estimation of economic key performance indicators for the technologies. The model is available online [DEL, 2018]. The main economic assumptions common for all technologies and capture cases are presented in Table 2.4.

For the MEA, oxyfuel, membrane-assisted liquefaction and calcium-looping technologies, estimation of total equipment costs (TEC) and installation costs for most standard process equipment is done using Aspen Process Economic Analyzer® (APEA). For the calcium-looping technologies, the costs of the heat recovery steam cycle components are estimated using the Thermoflex® software. The estimation is based on key characteristics of each equipment from process simulations and design criteria, such as pressure, temperature, flows and materials. Estimates for other, non-standard components are based on information provided by the CEMCAP industry partners and literature. This includes e.g. several non-commercial process equipment in the oxyfuel and calcium-looping systems, membrane packages and multi-stream plate and fin heat exchangers used in CO_2 purification units (CPU). More details on design criteria for standard process equipment and cost estimation methodologies for non-standard equipment can be found in CEMCAP report D4.4 *"Cost of critical components in CO_2 capture processes"* [CIN, 2018].

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Estimation of TEC and installation costs for the chilled ammonia CO_2 capture process is performed by GE using their proprietary tool QFACT which is based on an extensive database of executed projects. The unit costs are lumped into TEC and installation costs as to not disclose GE confidential information about cost structure and/or pricing strategy.

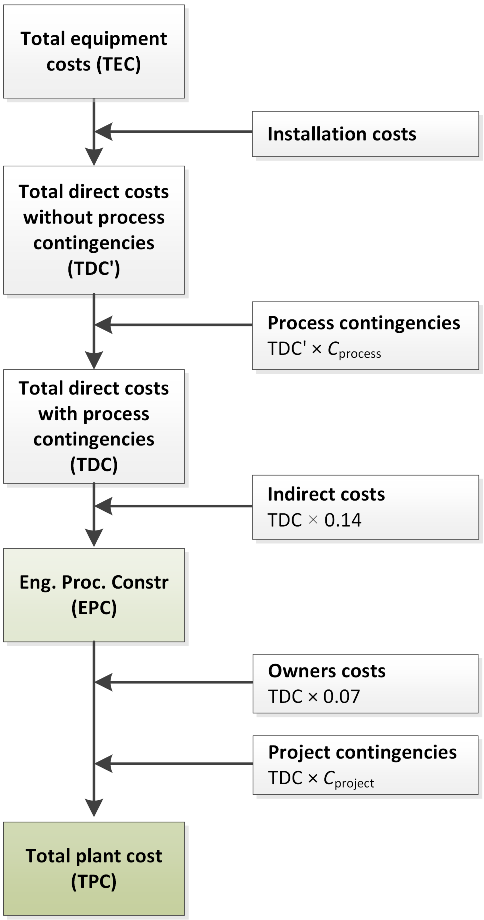

Figure 2.1. Bottom-Up approach for estimation of total plant costs.

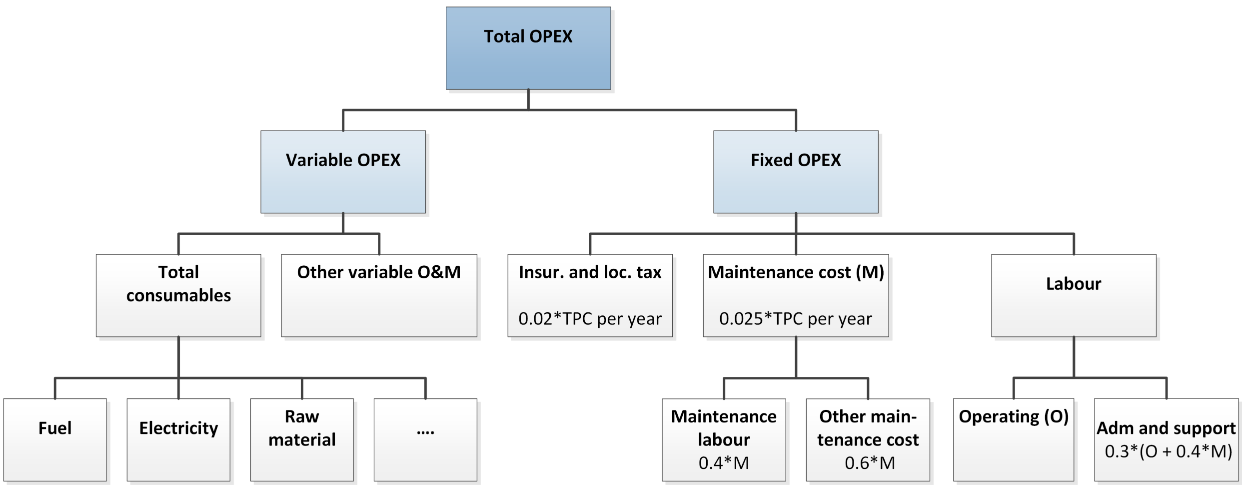

Figure 2.2. Estimation of operating costs.

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Table 2.4. Main assumptions for the economic evaluations.

General	
Cost basis	€2014
Capacity factor [%]	91.3
Tax rate [%]	0
Operational life [years]	25
Construction time, cement kiln [years]	2
Construction time – capture plant [years]	3
Inflation rate [%]	0
Discounted cash flow rate [%]	8
CAPEX	
Indirect costs [% TDC]	14
Owners costs [% TDC]	7
Project contingencies [% TDC]	15
OPEX	
Raw meal price [€/t_{clk}]	5
Coal price [€/GJ_{LHV}]	3
Natural gas price [€/GJ_{LHV}]	6
Price of electricity [€/MWh]	58.1
Cost of the steam produced from a natural gas boiler [€/MWh]	25.3
Cost of the steam produced from the cement plant waste heat [€/MWh]	8.5
Carbon tax [€/t_{CO2}]	0
Cooling water cost [€/m^3]	0.39
Process water cost [€/m^3]	6.65
Ammonia solution price for SNCR [€/t]	130
MEA solvent [€/t]	1450
Ammonia solvent [€/t]	406
Sulfuric acid [€/t]	46
Sodium hydroxide for flue gas desulfurization [€/t]	370
Membrane material replacement [€/m^2]	7.87
Other variable O&M [€/$t_{clinker}$]	1.09
Insurance and loc. Tax [% TPC]	2
Maintenance cost (including maintenance labor) [% TPC]	2.5
Cost of labor per person [k€/year]	60
Maintenance labor cost [% Maintenance]	40
Administrative labor cost [% O&M labor]	30

Process contingencies for the cost estimation of the technologies are based mainly on the maturity or status of the technology, but it was decided to adjust the resulting values based on the level of details of the equipment list, which vary among the technologies depending on the industry involvement in the cost estimations. The total process contingency factor is therefore estimated as a sum of the two factors $C_{process,1}$ (technology maturity) and $C_{process,2}$ (detail level of equipment list). An overview of the process contingency factors assumed for the different levels of technology maturity ($C_{process,1}$) is presented in Table 2.5. For the contribution from the detail level of the equipment list ($C_{process,2}$), the level of detail that can be provided by an industry partner developing a technology is the baseline (zero added process contingency). The estimated process contingencies for all CEMCAP technologies are listed in Table 2.6.

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Table 2.5. Links between technology status and process contingencies as defined in the CEMCAP project.

Technology status	Process contingency [% of TDC']
New concept with limited data	40+
Concept with bench-scale data	30-70
Small pilot plant data	20-35
Full-sized modules have been operated	5-20
Process is used commercially	0-10

Table 2.6. Process contingency factors for CEMCAP core technologies.

CO_2 capture technology	Technology status	Process contingency – maturity, $C_{process,1}$ [% TDC']	Process contingency – detail level of equipment list, $C_{process,2}$ [% of TDC']
MEA	Full-sized modules have been operated	15	3
Oxyfuel	Small pilot plant data	30	12
Chilled ammonia process	Small pilot plant data	20	0
Membrane-assisted CO_2 liquefaction	Concept with bench-scale data	40	12
Calcium looping – tail-end	Small pilot plant data	20	12
Calcium looping – integrated EF	Concept with bench-scale data	60	12

Several subsystems which are not included in the core process can be regarded to have a different technology status than the core processes. These technologies are listed in Table 2.7 together with their technology status and the associated technology maturity related process contingency, $C_{process,1}$. The process contingency associated with the level of detail of the equipment list, $C_{process,2}$, for these technologies is assumed the same as for the core technology (*cf.* Table 2.6), except for the ASU and the cooling system where the cost is estimated directly for the complete unit [CIN, 2018].

Table 2.7. Subsystems and associated processes contingencies.

Subsystem	Relevant CEMCAP technology	Technology status	Process contingency – maturity, $C_{process,1}$ [% TDC']	Process contingency – detail level of equipment list, $C_{process,2}$ [% of TDC']
Air separation units	Oxyfuel, CaL tail-end, CaL integrated EF	Commercially used process	5	0
Refrigeration systems	All technologies	Commercially used process	5	Same as core process
Cooling systems	All technologies	Commercially used process	5	0
CO_2 purification units	All technologies	Small pilot plant data	20	Same as core process

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Fixed operating costs as well as variable operating and maintenance (O&M) costs are estimated based on assumptions for material replacement and factor approach. Variable operating costs in terms of various consumables are mainly based on process simulations. Further details on assumptions regarding the different cost factors that make up the TDC and the OPEX can be found in the CEMCAP framework [VOL, 2018]. The economic evaluation is done considering retrofit to an existing plant for all technologies.

2.3 Key performance indicators

Key performance indicators (KPIs) are defined for comparative evaluation of the capture technologies, both with respect to CO_2 avoided and energy consumption (energy and environmental KPIs), and with respect to costs (economic KPIs).

2.3.1 Energy and environmental indicators

The *CO_2 capture ratio* (CCR) is a common KPI for CO_2 capture processes. It is defined as the CO_2 captured $\dot{m}_{\mathrm{CO2,capt}}$ divided by the CO_2 generated $\dot{m}_{\mathrm{CO2,gen}}$:

$$CCR = \frac{\dot{m}_{\mathrm{CO2,capt}}}{\dot{m}_{\mathrm{CO2,gen}}} \tag{1}$$

The CO_2 generated is both CO_2 generated in the cement kiln and CO_2 generated by fuel combustion internally in the capture process, but not CO_2 generated in the NG boilers or CO_2 generated indirectly by power consumption. The CO_2 captured is the total amount of CO_2 captured. It is not distinguished between CO_2 originating from the cement process, and CO_2 originating from additional fuel combustion.

The *CO_2 avoided from flue gas* evaluates the direct CO_2 emission reduction from the flue gas of the cement kiln. It is defined as:

$$AC_{\mathrm{fg}} = \frac{e_{\mathrm{clk,fg,ref}} - e_{\mathrm{clk,fg}}}{e_{\mathrm{clk,fg,ref}}} \tag{2}$$

where $e_{\mathrm{clk,fg,ref}}$ denotes specific CO_2 emissions with the reference kiln flue gas, and $e_{\mathrm{clk,fg}}$ denotes the specific CO_2 emissions with the flue gas of the kiln with capture. For technologies where a CO_2 purification unit (CPU) is used, the CPU vent gas should be accounted as a part of the flue gas. Emissions from steam generation (NG boilers) and indirect emissions related to power consumption/generation are not considered. This indicator differs from the CCR, because CO_2 captured from additional fuel combustion internally in the capture process is not counted as CO_2 avoided.

The *CO_2 avoided* evaluates the direct CO_2 emission reduction from the plant, taking the direct emissions related to the steam generation (NG boilers) in addition to the direct emissions with the flue gas, but not indirect emissions, into account. It is defined as:

$$AC = \frac{e_{\mathrm{clk,ref}} - e_{\mathrm{clk}}}{e_{\mathrm{clk,ref}}} \tag{3}$$

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where $e_{\mathrm{clk,ref}}$ is specific direct emissions from the reference kiln, and e_{clk} is the specific direct emission from the kiln with capture.

The *equivalent CO_2 avoided* evaluates the total equivalent CO_2 emissions avoided at the plant, taking both the direct emissions, as well as the indirect CO_2 emissions associated to electric power consumption/generation, into account. It is defined as:

$$AC_{eq} = \frac{e_{\mathrm{clk,eq,ref}} - e_{\mathrm{clk,eq}}}{e_{\mathrm{clk,eq,ref}}} \tag{4}$$

where $e_{\mathrm{clk,eq,ref}}$ is specific equivalent emissions from the reference kiln, and $e_{\mathrm{clk,eq}}$ is the specific equivalent emission from the kiln with capture.

Equivalent emissions are defined as the sum of direct e_{clk} and indirect $e_{el,clk}$ emissions:

$$e_{\mathrm{clk,eq}} = e_{\mathrm{clk}} + e_{\mathrm{el,clk}} \tag{5}$$

Indirect emissions can be calculated using the following equation:

$$e_{\mathrm{el,clk}} = e_{\mathrm{el}} \cdot P_{\mathrm{el,clk}} \tag{6}$$

where $P_{\mathrm{el,clk}}$ is the specific power consumption, which is positive when power is consumed and negative when it is generated, and e_{el} is the CO_2 emissions associated with each unit of electric power consumed. This value depends largely on the electricity mix considered.

The equivalent CO_2 avoided takes all direct and indirect emissions into account. It gives the best indication on the overall reduction in CO_2 emissions of the cement kiln when a certain capture technology is implemented and allows a fair comparison of different technologies.

The *specific primary energy consumption for CO_2 avoided*, SPECCA, evaluates the primary energy used to avoid CO_2 emissions to the atmosphere. It is defined as the difference in equivalent primary energy consumption of the cement plant with and without CO_2 capture, divided by the difference in equivalent CO_2 emissions without and with capture:

$$SPECCA = \frac{q_{\mathrm{clk,eq}} - q_{\mathrm{clk,eq,ref}}}{e_{\mathrm{clk,eq,ref}} - e_{\mathrm{clk,eq}}} \tag{7}$$

Equivalent specific primary energy consumption is the sum of direct (q_{clk}) and indirect ($q_{\mathrm{el,clk}}$) specific primary energy consumption:

$$q_{\mathrm{clk,eq}} = q_{\mathrm{clk}} + q_{\mathrm{el,clk}} \tag{8}$$

The direct specific primary energy consumption is the amount of energy (LHV), supplied in the form of fuel (coal or natural gas), that is used per ton of clinker:

$$q_{clk} = \frac{\dot{m}_{fuel} \cdot LHV_{fuel}}{\dot{m}_{clk}} \tag{9}$$

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The indirect specific primary energy consumption is the amount of energy consumed by the generation of power required per ton of clinker. It can be calculated by:

$$q_{\mathrm{el,clk}} = \frac{P_{\mathrm{el,clk}}}{\eta_{\mathrm{el}}} \tag{10}$$

where η_{el} is the electricity generation efficiency, which also depends largely on the electricity mix considered.

2.3.2 Economic indicators

The economic performance of the capture technology is evaluated by the *cost of clinker* and the *cost of CO_2 avoided*.

The *cost of clinker* (COC) is evaluated by summing the contributions of the capital cost C_{cap}, converted into a yearly constant annualized flow through a capital carrying charge factor, of the fuel cost C_{fuel}, of the raw material costs C_{RM}, of the electricity cost C_{el}, and of the other operating and maintenance cost $C_{\mathrm{O\&M}}$, all expressed per ton of clinker produced (i.e. as €/t_{clk}). In case the cement plant is characterized by a net power export, revenues for electricity export to the grid (i.e. negative costs) are considered.

$$COC = C_{\mathrm{cap}} + C_{\mathrm{fuel}} + C_{\mathrm{RM}} + C_{\mathrm{el}} + C_{\mathrm{O\&M}} \tag{11}$$

The *cost of CO_2 avoided* (CAC) is evaluated with the following equation, comparing the cost of clinker and the equivalent specific emissions of the assessed cement plant and the reference cement plant without CO_2 capture:

$$CAC = \frac{COC - COC_{\mathrm{ref}}}{e_{\mathrm{clk,eq,ref}} - e_{\mathrm{clk,eq}}} \tag{12}$$

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3 REFERENCE CEMENT PLANT

3.1 Plant description

The reference cement plant is a Best Available Technique (BAT) plant defined by the European Cement Research Academy (ECRA). It is based on a dry kiln process, consists of a five-stage cyclone preheater, calciner with tertiary duct, rotary kiln and grate cooler. It has a capacity of 2,896 ton clinker per day. This corresponds to ca. 1 Mt clinker per year, or 1.36 Mt cement per year, with a run time of >330 days per year. This is a representative size for European cement plants. The characteristics of the reference cement plant are summarized in Table 3.1.

Table 3.1. Characteristics of the reference cement plant.

Parameter	Value
Clinker production	2,896 t_{clk}/d
Clinker/cement factor	0.737
Raw meal/clinker factor	1.6
Specific CO_2 emissions	850 kg_{CO2}/t_{clk}
Specific electric power consumption	97.0 kWh/t_{cement} / 132 kWh/t_{clk}

The clinker burning line of the reference cement plant is shown in Figure 3.1. The raw material is first grinded in the raw mill, where it is also dried by hot flue gas from the preheater. The flue gas and the resulting raw meal are subsequently separated in a dust filter, and the raw meal is sent to the preheater while the gas is sent to the stack.

In the preheater the meal is heated by hot flue gas coming from the calciner and the rotary kiln. The meal and the hot gases are mixed for heat transfer and separated in cyclones arranged above one another. Thereafter, the raw meal enters the calciner, where the major part of the calcination ($CaCO_3$ -> $CaO + CO_2$) is performed. Around 2/3 of the plant's total fuel input is consumed here to achieve the right temperature (~860 °C) and drive the endothermal reaction.

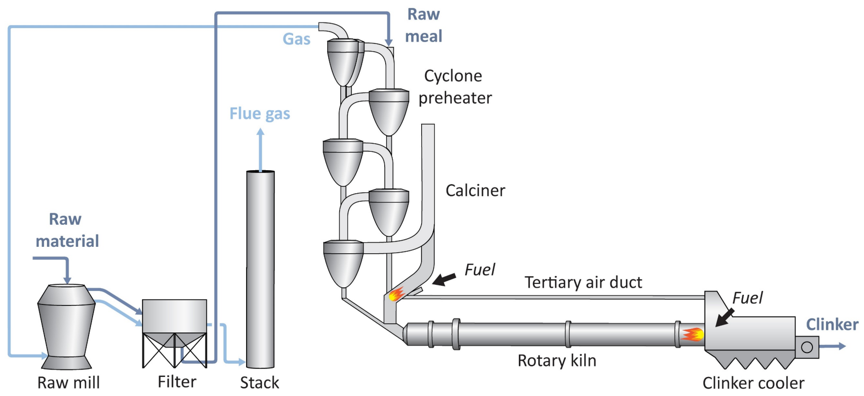

Figure 3.1. Clinker burning line in CEMCAP reference plant.

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After the calciner, the raw meal enters the rotary kiln, where formation of the clinker takes place. Around 1/3 of the plant's fuel is burnt in the main burner, which is placed in the other end of the kiln. In the rotary kiln the solid material reaches 1450 °C, and the temperature of the gas phase can reach 2,000 °C. During its way through the rotary kiln the raw material components form clinker via intermediate phases.

The hot clinker is discharged from the kiln to a clinker cooler. In the cooler ambient air is used to cool the clinker. Some of the resulting hot air is used as combustion air in the main burner (secondary air) and in the calciner (tertiary air).

A selective non-catalytic reduction (SNCR) system is installed in the kiln for control of NO_x emissions. SO_x emissions are below the limit of 400 mg/Nm^3, set by the EU directive on industrial emissions [EUR, 2010], so no system is installed for SO_x emission control.

The CEMCAP reference kiln is identical to the ECRA reference kiln [LOC, 2002; KLE, 2006], with the exception that a SNCR system is assumed to be installed in the CEMCAP kiln and not in the ECRA reference kiln.

3.1.1 Electric power consumption

The electric power consumption of the reference cement kiln is associated with fans, coal milling and handling, raw meal and cement grinding, solids handling, etc. The power consumption distributed between the different users is presented in Table 3.2.

Table 3.2. Electric power consumption associated with the reference cement plant utilities.

	Flow rate [m^3/h]	Flow rate [Nm^3/h]	Temperature [°C]	ΔP [kPa]	Power [kWh/t_{clk}]
ID fan	349 440	162 564	314	6.35	6.40
Raw mill fan	411 712	293 512	110	10.70	12.70
Filter fan	584 117	439 355	90	1.80	3.03
Cooler fans	245 081	232 323	15	2.15	1.52
Coal milling and handling	-	-	-	-	5.81
Others (raw meal and clinker grinding, solids handling, kiln drive, lightning, etc.)	-	-	-	-	102.14
Total	-	-	-	-	132

3.1.2 Flue gas conditions

The composition and flow rate of the flue gas at the stack are largely dependent on the amount of air leaking into the system. Air can typically leak into the system at the kiln inlet and outlet, in the calciner, in the preheater cyclones, and in the raw mill. The amount of air leakage in the raw mill is predominant compared to the leakage at other points in the kiln. The air that has leaked into the flue gas is commonly referred to as "false air".

The air leak typically increases with time after the yearly plant maintenance period at the plant and is decreased again with maintenance. Since the air leak in the mill is predominant, the change in air leak over the year in the reference plant is accounted for by assuming low air leak in the mill the first part of the year, and medium air leak in the mill the second part of the year.

Air leak normally also varies throughout the day, as the raw mill often is out of operation a couple of hours every day, resulting in lower air leak, which in turn gives a higher CO_2 concentration and lower flue gas flow rate [VOL, 2018]. For simplicity, this effect is not considered in the present analysis.

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Flue gas conditions during the first and second part of the year are given in Table 3.3. The NO_x and SO_x levels correspond to 500 and 200 mg/Nm^3 at 10% O_2 content respectively.

Table 3.3. Flue gas conditions at stack.

	First ½ year	Second ½ year
Air leak in mill	Low	Medium
False air at preheater outlet (Nm^3/Nm^3) [%]	5.5	5.5
False air at stack (Nm^3/Nm^3) [%]	30	43
Total flow rate [kg/h]	318,192	388,098
Temperature [°C]	130	110
Mole fraction (wet basis)		
CO_2	0.22	0.18
N_2	0.60	0.63
O_2	0.07	0.10
H_2O	0.11	0.09
NO_x (dry basis) [mg/Nm^3]	591	455
SO_x (dry basis) [mg/Nm^3]	236	182
Dust (dry basis) [mg/Nm^3]	10	10

3.1.3 Waste heat recovery

Waste heat can be recovered in the reference cement plant from the cooler exhaust air. The dust in the exhaust air is assumed to be removed with a ceramic filter prior to the heat recovery steam generator, which is assumed to have a minimum ΔT of 80 °C. The waste heat available for steam generation at various qualities is shown in Figure 3.2.

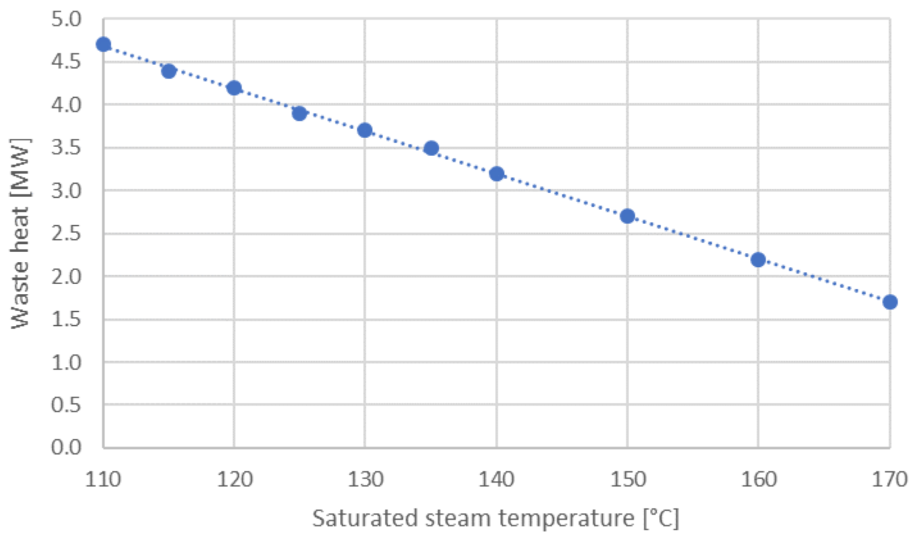

Figure 3.2. Waste heat available for steam generation in the reference plant.

3.2 Process evaluation

The core process of the reference cement kiln is simulated by VDZ with their in-house cement plant process model [LOC, 2011]. This simulation is reproduced by Politecnico di Milano (PoliMi) with their in-house process simulation tool GS [GS, 2016]. The VDZ simulation is used

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as basis for simulation of the oxyfuel process, and it is used as reference for this process and the post combustion capture processes MEA, CAP, and membrane-assisted CO_2 liquefaction. The PoliMi simulation is used as basis for the simulation of the CaL processes and used as reference for these processes. The process data presented in this document for the reference cement kiln are based on the VDZ simulation. Details on both the VDZ and PoliMi simulations can be found in CEMCAP report D4.2 "Design and performance of CEMCAP cement plant with MEA post combustion capture" [ANA, 2016], and stream data is provided in Section 1 in the Supplementary information of this report.

3.3 Economic evaluation

The TDC of the reference cement plant is based on estimations from the IEAGHG [IEA, 2013] for a BAT cement plant with the same clinker capacity as the CEMCAP reference plant and amounts to 149.8 M€$_{2014}$. This includes the added costs of a DeNOx system, based on standard selective non-catalytic reduction (SNCR) process, assumed to be installed in the reference cement plant. The SNCR system uses ammonia solution as a reduction agent and has an average reduction rate of 60%. The TDC for the SNCR system is assumed to be 1.01 M€$_{2014}$.

The utility consumption of the reference cement kiln plant is presented in Table 3.4, together with other consumables and the operating labor at the plant. The utility consumption is assumed to be constant over the year. The economic data of the reference cement kiln plant presented in Table 3.5 is estimated based on the data listed in Table 3.4 as well as the economic assumptions described in Section 2.2. The resulting cost of clinker is calculated to 62.6 €/t_{clk} and the cost of cement to 46.0 €/t_{cement}. It could be pointed out that the cost calculated in CEMCAP is somewhat lower than the 51.5 €/t_{cement} reported by the IEAGHG [IEA, 2013]. The main reasons for this difference are due to the higher capacity factor assumed in CEMCAP (91.3%, vs. 80%), causing a higher impact of CAPEX and fixed OPEX, and to the lower price of electricity assumed in CEMCAP (58.1 €/MWh vs. 80 €/MWh).

Table 3.4. Utilities, consumables and operating labor for reference cement kiln plant.

	Cement plant without CO_2 capture
Clinker production [t/h]	120.65
Coal [t/h]	13.93
Electricity [MW]	15.88
Ammonia solution for SNCR [t/h]	0.60
Operating labor - number of persons	100

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Table 3.5. Economic data for reference cement kiln plant.

	Cement plant without CO_2 capture
Variable OPEX [€/t_{clk}]	**23.8**
Raw meal	5.0
Coal	9.4
Electricity	7.7
Ammonia for SNCR	0.7
Other variable costs	1.1
Fixed OPEX [€/t_{clk}]	**18.2**
Operative, administrative and support labor	8.7
Insurance and local taxes	4.2
Maintenance cost (including maintenance labor)	5.3
Total plant costs [M€]	**203.8**
TDC with process contingencies	149.8
Indirect costs	21.0
Owner's costs and contingencies	33.0
Total OPEX [€/t_{clk}]	**42.0**
Total CAPEX [€/t_{clk}]	**20.6**
Cost of clinker [€/t_{clk}]	**62.6**
Cost of cement [€/t_{cement}]	**46.0**

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4 CAPTURE CASE SPECIFICATIONS

Several CO_2 capture cases are investigated in CEMCAP. The cases differ with respect to the amount of air leak in the raw mill, the extent of CO_2 capture, CO_2 transport option and method of steam supply. An overview of the cases is presented in Table 4.1, and a more detailed description of each case is provided subsequently.

Table 4.1. Overview of the main cases studied in CEMCAP.

Case	Air leak	CO_2 avoided from flue gas (AC_{fg})	Transport option	Steam supply
Base case	Increasing	90%	Pipeline	NG boiler and waste heat
Low air leak in mill	Constant low	90%	Pipeline	NG boiler and waste heat
Optional extent of capture	Increasing	Optional	Pipeline	NG boiler and waste heat
Ship transport	Increasing	90%	Ship	NG boiler and waste heat
Steam import	Increasing	90%	Pipeline	Steam import and waste heat

4.1 Base case

In the base case, air leakage in the raw mill of the reference cement plant is assumed to increase over the year, consequently affecting the flue gas composition, as described in Section 3.1.2. Here, low air leak in the mill is considered for the first half of the operating year and medium air leak is considered for the latter half of the year. The CO_2 capture system is designed for medium air leak conditions, for which capital costs are estimated. The system energy and mass balances are calculated for both medium and low air leak conditions and average values for consumables, i.e. steam, electricity, fuel etc., are used to calculate operating costs and energy and environmental KPIs.

Furthermore, the CO_2 avoided from the cement plant flue gases should be as close to 90% as possible. CO_2 avoided from the flue gas is by definition identical to carbon capture ratio (CCR) for the MEA, chilled ammonia and membrane-assisted CO_2 liquefaction technologies.

The captured CO_2 is compressed and prepared for transport by pipeline. The required CO_2 pressure is 110 bar and the temperature should not be higher than around 30 °C. Details on requirements for CO_2 purity and maximum impurity concentrations as well as on purified flue gas conditions can be found in the CEMCAP framework [VOL, 2018].

For technologies requiring steam in their operation, available waste heat from the cement plant is used (Section 3.1.3), with additional steam required generated by a natural gas (NG) boiler.

4.2 Low air leak in mill

In this case, increased maintenance over the year relative to the base case is assumed, resulting in a constant low air leak in the raw mill of the reference cement plant and thus constant flue gas conditions with low air leak over the whole year. The increased maintenance is accounted for by reducing the annual operating hours of the plant with 72 hours, representing an additional stoppage

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in the kiln line for the additional maintenance. This effectively reduces the capacity factor of the plant from 91.3% to 90.5%.

This case is not relevant for the oxyfuel and calcium looping technologies since these capture technologies are not influenced by air leak in the raw mill.

4.3 Optional extent of capture

In this case, the extent of capture differs from that of the base case, while all the other characteristics as listed in Table 4.1 are the same. This case is of interest for the membrane-assisted liquefaction and CAP technologies and is therefore only evaluated for these technologies and the reference technology MEA. The optimal capture rate will differ between the technologies, and the extent of capture in this case is therefore optional, although it should not be below 60%.

4.4 Ship transport

This case is identical to the base case, except for the transportation option, which in this case is by ship. The ship transport requires a different CO_2 conditioning process compared with the pipeline transport, and in this case the CO_2 is delivered in liquid form at 6.5 bar and subcooled with 3 °C (around -52 °C). The CO_2 should be at least as pure as for pipeline transport, but stricter limits apply to some components to avoid operational problems (solid formation) in the liquefaction process.

4.5 Steam import

The absorption-based CO_2 capture technologies (MEA and CAP) are highly dependent on the supply of steam to the capture process for solvent regeneration. The steam generation option will therefore have a considerable impact on the process economics and on the total amount of CO_2 avoided. In this case, the steam required in addition to the steam generated from the cement plant's waste heat is assumed to be imported from an external coal CHP plant, located near the cement plant. Except for the steam supply option, this case has the same characteristics as the base case. It should be mentioned that less than 10% of existing European cement plants are located near a coal power plant. However, in the case of a new cement plant to be built with CCS, this alternative could significantly decrease the CO_2 capture cost associated with the cement plant.

4.6 Other cases

Other cases than those previously described are considered for some of the technologies. For the oxyfuel technology, cases with different amounts of air leakage in the core process are investigated. For the calcium looping tail-end configuration, cases with different integration levels are investigated.

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5 MEA ABSORPTION

Absorption with aqueous solution of 30% monoethanolamine (MEA) is the reference technology in CEMCAP. This technology is an end-of-pipe retrofit technology. The integration with the cement plant is shown in Figure 5.1. The flue gas from the cement plant is sent to the MEA process right before it reaches the stack. The process requires a considerable amount of heat for solvent regeneration, and power is required for fans and pumps in both the MEA process, and conditioning (compression or liquefaction) of the captured CO_2. Waste heat from the cement plant is used to cover a part of the heat demand.

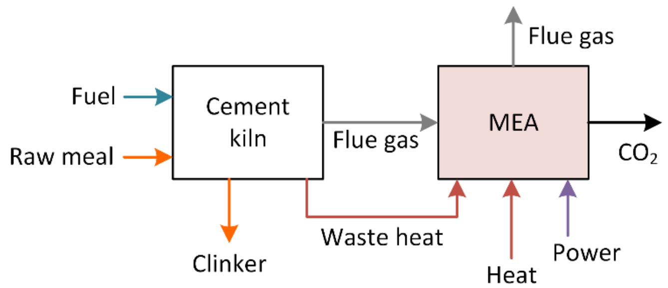

Figure 5.1. Schematic overview of the integration of MEA CO_2 capture with a cement kiln.

5.1 Process description

The concept of the MEA process is illustrated in Figure 5.2, and a detailed flowsheet of the process is shown in Figure 5.3. The CO_2 rich flue gas (stream 1) is first cooled against CO_2 lean flue gas (stream 4) after being slightly compressed in the exhaust fan. The flue gas (stream 2) is then cooled in a direct contact cooler (DCC), where it also comes into contact with a NaOH solution which is sprayed into the DCC to remove SO_x and ensure that its concentration meets the requirement of the MEA capture process. Furthermore, a major portion of water is removed from the gas in the DCC. The flue gas (stream 3) is sent to an absorber where the CO_2 is absorbed by the MEA solvent. A water wash column is placed at the top of the absorber where MEA is recovered. The CO_2 lean flue gas from the water wash column (stream 4) is then heated against the CO_2 rich flue gas before being vented to the atmosphere. A minor portion of the water is sent to the waste water treatment facility (streams 13 and 15).

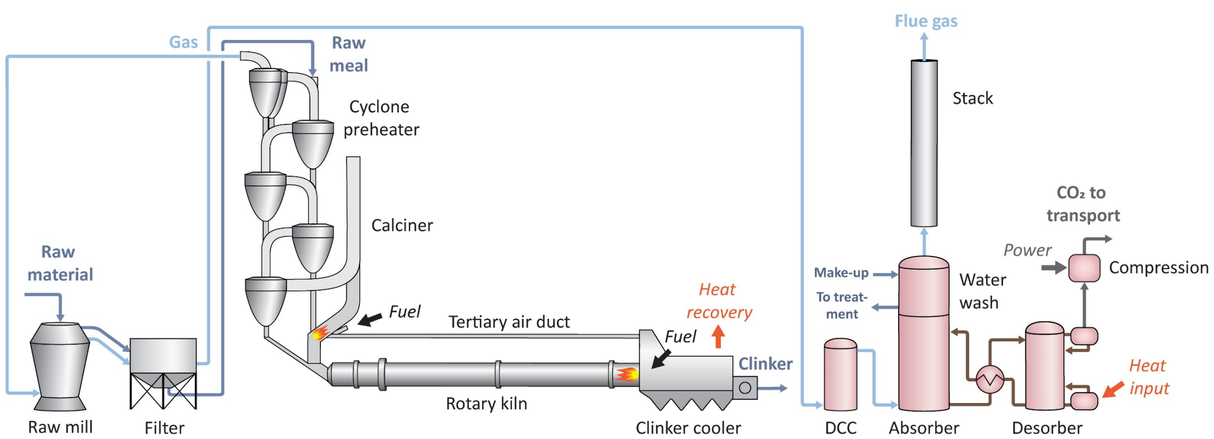

Figure 5.2. Reference cement kiln with MEA CO_2 capture.

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The CO_2 rich MEA solvent from the bottom of the absorber is pumped (stream 6) and heated against the lean MEA solvent for heat recovery before being sent to the stripper for regeneration (stream 8). The resulting CO_2 (stream 16) is compressed or liquefied in order to meet the transport specifications. The CO_2 lean solvent (stream 9) from the bottom of the stripper is sent to the absorber after being cooled in the lean-rich heat exchanger and the lean amine cooler. A thermal reclaimer unit removes heat stable salts (due to MEA degradation) from the solvent. In addition, a small portion of lean MEA passes through the amine filter to further remove the solids and degradation products. As a result, fresh MEA (stream 7) is necessary for make-up.

For modern cement plants without CO_2 capture, DeSOx units are normally not required if sulfide minerals are not presented in the raw materials [IEA, 2008]. In the reference cement kiln, there is therefore no DeSOx system installed. In order to limit MEA degradation due to SO_x, injection of a NaOH solution (stream 14) in the DCC is included in the MEA setup. DeNOx units may or may not be required in modern cement plants, depending on the NO_x emissions and local emission limits. A selective non-catalytic reduction (SNCR) system, operating with ammonia solution as a reducing agent, is assumed to be installed in the reference cement kiln. Further reduction in NO_x levels required in order to limit MEA degradation is obtained by increasing the amount of reagent in the existing system.

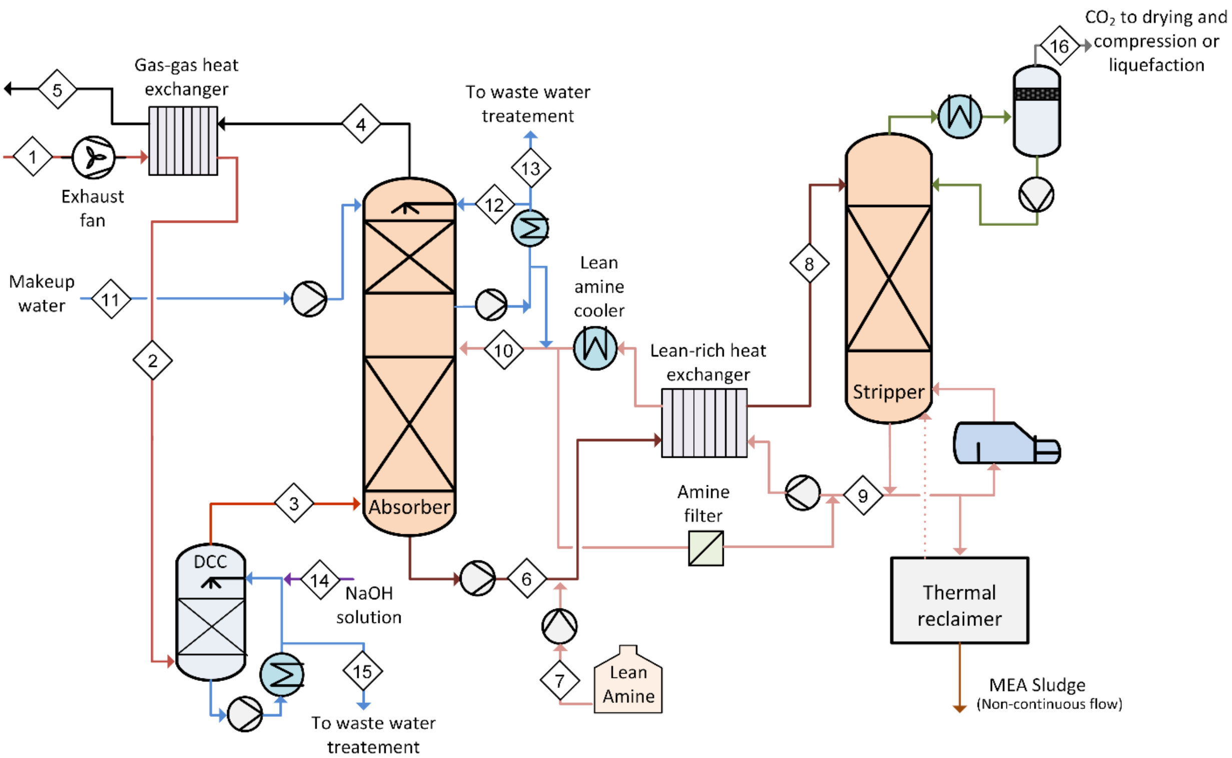

Figure 5.3. Process flowsheet of the MEA CO_2 capture process.

The process flowsheets of dehydration and compression for the pipeline transport cases and dehydration and liquefaction for the ship transport case are given in Figure 5.4 and Figure 5.5. In the case of pipeline transport, the CO_2 captured is compressed with 3 compression stages. The heat of compression is removed by the 3 coolers that are located after each compression stage. A drum is placed after each of the first two coolers to remove the water condensate. A TEG dehydration unit (Appendix A.1) is placed after the second drum to further dry the CO_2 stream, effectively removing 95% of the water from the inlet stream. The pressure of CO_2 is increased to the transport pressure by a pump that is located after the third cooler. The CO_2 is further cooled to the transportation temperature in the last cooler located after the pump.

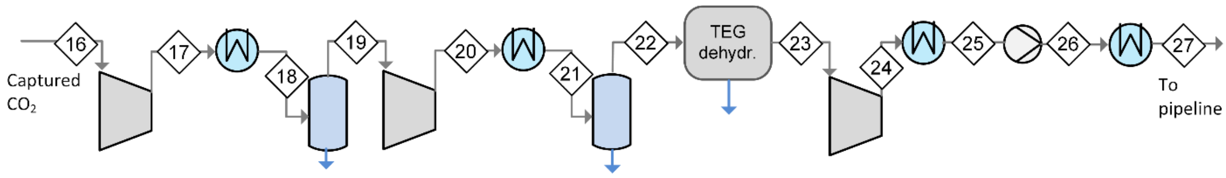

Figure 5.4. Process flowsheet of CO_2 drying and compression for pipeline transport.

In the case of ship transport, two compression stages are used. One interstage cooler (IC), one chiller and one drum are used after each compression stage to cool the CO_2 and remove the water condensate. A molecular sieve dehydration unit (Appendix A.2) is used to further limit the water content in the CO_2 stream and it practically removes all water from the inlet stream. The CO_2 is condensed in the CO_2 condenser. A valve and a subcooler are used to condition the CO_2 to the transport pressure and temperature. The vapor resulting from the valve flash (stream 32) is returned to the second-stage compressor after being heated.

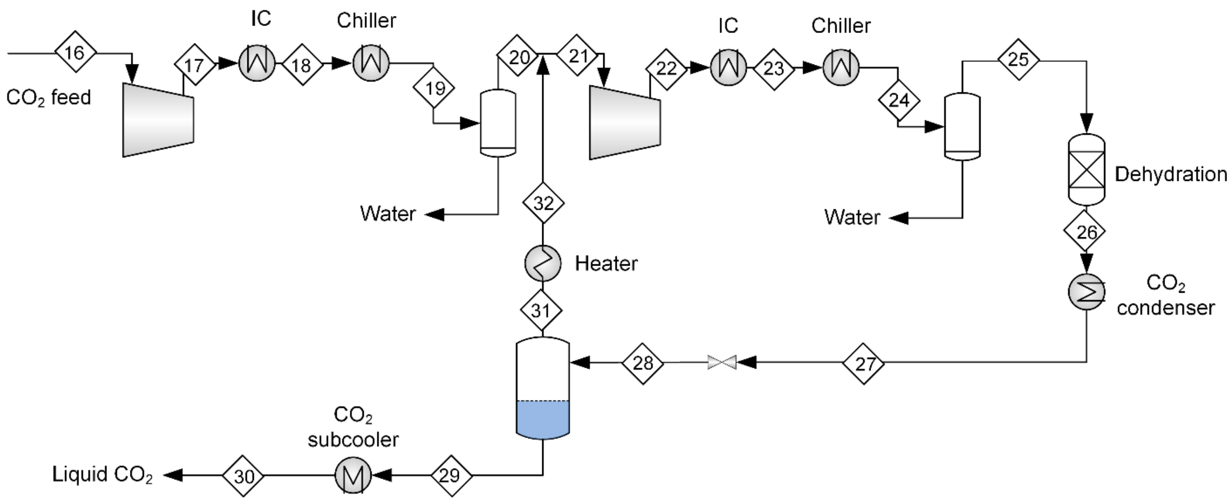

Figure 5.5. Process flowsheet of CO_2 liquefaction for ship transport.

5.2 Process evaluation

The MEA system as shown in Figure 5.3 is modelled with the process simulator Aspen HYSYS V9. The Acid Gas property package is selected for modelling the part of the process including MEA solvent. The SRK property package is used for calculating properties of the flue gas and CO_2 streams. The process design and sizing procedure of equipment is primarily based on previous work performed in the ReCap project [IEA, 2017]. The regenerator reboiler requires steam at approximately 128 °C. According to Figure 3.2, 3.7 MW is available from heat recovery at the cement plant at the required temperature level. The rest of the steam is generated in a natural gas boiler or imported from an external coal CHP plant. Thermal reclaiming of MEA solvent is not included in the process model and instead, the amount of degradation products and the energy

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requirement of the reclaimer are based on a study by IEAGHG [IEA, 2014]. The heat demand of the thermal reclaimer is taken as 10.4 kWth/kg Heat Stable Salts (HSS), where the HSS content of the treated solvent stream is assumed to be 3 %wt. The amount of solvent make-up required in the system is estimated from the solvent slip with the CO_2-lean flue gases, the high-purity CO_2 stream, the water-wash purge stream and from estimates on solvent degradation. The amount of solvent lost in the reclaimer waste stream is assumed to be 1.4 kg MEA/ton CO_2 captured [KNU, 2009].

The same system design is used for both the base case (with increasing air leak) and the low air leak case. The system is designed for 90% CO_2 avoided from the flue gas. In the case with 60% CO_2 avoided from the flue gas, a part of the flue gas bypasses the capture unit. This seems to be the most economical option for MEA [HIL, 2009]. The compression or liquefaction of CO_2 to meet the specifications for pipeline and ship transport respectively, is modelled as specified in the CEMCAP framework. For the liquefaction option the CO_2 condenser temperature is decreased compared to the process described in the framework, because of the impact of impurities on the condensation temperature of the gas.

The concentration of NO_x at the absorber inlet should be limited to 410 mg/Nm^3 for MEA-based CO_2 absorption [IEA, 2008]. The NO_x concentration in flue gases from all modes of operation in the reference cement plant, listed in Table 3.3, is higher than this limit. An increase in NO_x removal rate to reach the set limit is achieved by increasing the injection rate of ammonia solution in the SNCR system, with 1.5 times the stochiometric amount. For more details on the NO_x removal process, see Appendix B. The concentration of SO_x at the absorber inlet should be limited to 10 ppm$_v$ prior to the MEA capture process. SO_x is reduced to acceptable levels with injection of 50% NaOH solution in the DCC, where all SO_x is assumed to be removed in the process. A stochiometric amount of the NaOH solution is assumed to be sufficient to achieve the reduction. For more details on the SO_x removal process, see Appendix C.

Key simulation results for MEA CO_2 capture cases are presented in Table 5.1. Stream data are provided in Section 2 in the Supplementary information. In all cases, loading of lean MEA is 0.22 mol CO_2/mol MEA and loading of rich MEA is 0.5 mol CO_2/mol MEA.

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Table 5.1. Key process characteristics and simulation results for MEA cases.

Case	**Base case**		**Low air leak in mill**	**Optional extent of capture**		**Ship transport**		**Steam import**	
Design	Medium air leak		Low air leak	Medium air leak		Medium air leak		Medium air leak	
Operation	Low air leak	Medium air leak	Low air leak	Low air leak	Medium air leak	Low air leak	Medium air leak	Low air leak	Medium air leak
Captured and conditioned CO_2 mole fraction (wet basis)	0.999	0.998	0.999	0.999	0.998	0.999	0.998	0.999	0.998
Specific reboiler duty CO_2 desorber [MJ/kgCO2]	3.76	3.80	3.77	3.76	3.80	3.76	3.80	3.76	3.80
Reboiler temperature CO_2 desorber [°C]	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4
Steam generation from waste heat [%]	4%	4%	4%	6%	6%	4%	4%	4%	4%
Steam generation in NG boiler [%]	96%	96%	96%	94%	94%	96%	96%	0%	0%
Steam import [%]	0%	0%	0%	0%	0%	0%	0%	96%	96%
Specific power consumption [MJ/kg_{CO2}]	0.521	0.549	0.522	0.524	0.552	0.600	0.624	0.521	0.549
Cooling water demand [MJ/kg_{CO2}]	3.80	3.63	3.83	3.80	3.63	4.06	3.88	3.80	3.63
MEA make-up [kg/kg_{CO2}]	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
Process water make-up [kg/kg_{CO2}]	0.473	0.528	0.468	0.473	0.528	0.473	0.528	0.473	0.528

CO_2 capture ratio and CO_2 avoided for the MEA cases are shown in Table 5.2. SPECCA values are presented in Figure 5.6, and break-downs of the different contributions to the SPECCA values are presented in Table 5.3. The following can be noticed:

- The largest contributor to the SPECCA is the consumption of steam, both in terms of added energy consumption and associated CO_2 emissions which reduce the equivalent CO_2 avoided. The remaining contribution to the SPECCA comes from electric power consumption.

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- The base case and the low air leak case have very similar energy performances. The steam and power consumption is only slightly reduced in the low air leak case compared to the base case.
- The optional extent of capture case (60%) has SPECCA very similar to the base case. This is expected since the 60% capture case is a MEA system similar to the base case, just smaller with some of the flue gas bypassed. Since the total steam demand per clinker decreases and the waste heat per clinker is constant compared to the base case, a slightly higher share of the steam demand can be covered by waste heat recovery (6% versus 4%). This explains the small decrease in SPECCA.
- The ship transport case has a slightly lower equivalent CO_2 avoided and a higher SPECCA than the base case, due to the power consumption of the CO_2 liquefaction process necessary for ship transport, that is approximately doubled compared to the CO_2 compression required for pipeline transport.
- The case with steam import from a coal power plant has the highest equivalent CO_2 avoided, and SPECCA almost half of the base case. This is explained by the reduced equivalent primary energy consumption and CO_2 emissions of steam imported compared to steam generated in NG boilers. The equivalent primary energy consumption is approximately half for the steam imported from the coal power plant compared to the steam generated in NG boiler, because for the power plant some of the primary energy input is assigned to the power generated, and some of it is assigned to the steam. For the same reason, the equivalent CO_2 emissions is three quarters of the imported steam compared to the steam generated in a NG boiler. It should be mentioned that very few cement plants are located close to a power plant, but this option could be interesting for the planning of new plants with CCS.

Table 5.2. CO_2 capture ratio and CO_2 avoided for MEA cases.

	Base case	**Low air leak in mill**	**Optional extent of capture**	**Ship transport**	**Steam import**
CO_2 capture ratio	90%	90%	60%	90%	90%
CO_2 avoided from flue gas	90%	90%	60%	90%	90%
CO_2 avoided	70%	70%	47%	70%	90%
Equivalent CO_2 avoided	64%	64%	43%	63%	68%

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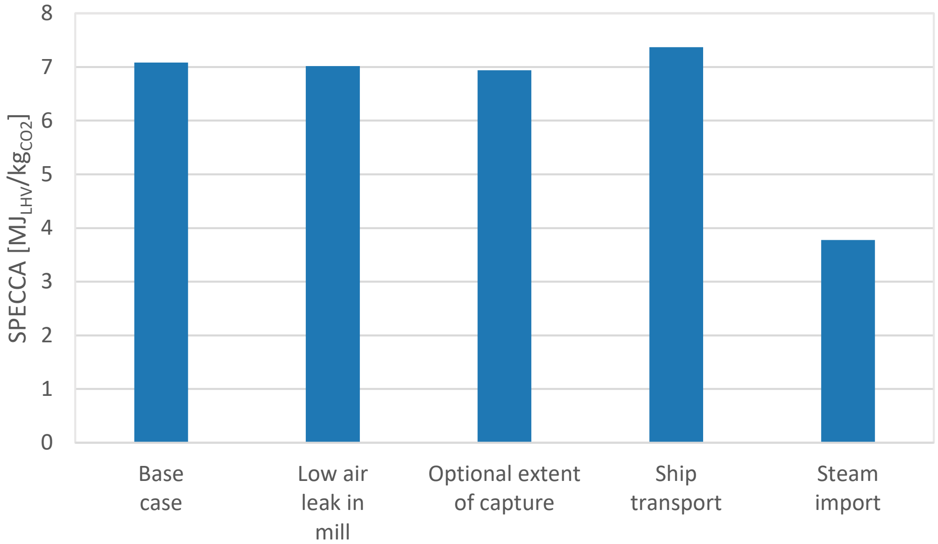

Figure 5.6. SPECCA for MEA cases.

Table 5.3. Break-down of SPECCA for MEA cases.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**3959**	**3932**	**2601**	**4088**	**2269**
Steam consumption	3073	3066	2007	3073	1382
Electric power consumption					
Fans, pumps, thermal reclaimer	188	163	129	188	188
Compression/liquefaction for transport	575	575	383	696	575
Cooling water system	123	127	82	132	123
Equivalent specific CO2 avoided [kg_{CO2}/t_{clk}]	**559**	**560**	**375**	**555**	**601**
At cement kiln stack	761	761	507	761	761
Steam consumption	-172	-172	-113	-172	-131
Electric power consumption					
Fans, pumps, thermal reclaimer	-6.3	-5.4	-4.3	-6.3	-6.3
Compression/liquefaction for transport	-19.2	-19.2	-12.8	-23.2	-19.2
Cooling water system	-4.1	-4.2	-2.7	-4.4	-4.1
SPECCA [MJ_{LHV}/kg_{CO2}]	**7.08**	**7.02**	**6.94**	**7.37**	**3.78**

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5.3 Economic evaluation

The utility consumption, other consumables and the required operating labor of the cement kiln with MEA CO_2 capture is presented in Table 5.4. Economic results in terms of both capital and operating costs are presented in Table 5.5, together with the resulting cost of clinker and CO_2 avoided cost. Detailed equipment lists with estimated equipment costs and direct costs are provided in Section 2 in Supplementary information.

The process contingency of the MEA technology is set to 18%. Full size modules of the technology have been operated in industry conditions and the contribution to the process contingency based on technological maturity ($C_{process,1}$) is therefore set to 15 %-points. The level of detail of the equipment list is relatively high, so the contribution to process contingency based on this ($C_{process,2}$) is set to 3 %-points, resulting in the overall value of 18%.

Table 5.4. Utilities, consumables and operating labor for cement kiln with MEA CO_2 capture.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Clinker production [t/h]	120.65	120.65	120.65	120.65	120.65
Coal [t/h]	13.93	13.93	13.93	13.93	13.93
Electricity [MW]	29.52	29.19	25.02	31.50	29.52
Steam – from waste heat [MW]	3.70	3.70	3.70	3.70	3.70
Steam – from NG boiler [MW]	92.67	92.49	60.53	92.67	0
Steam – import from power plant [MW]	0	0	0	0	92.67
Cooling water make-up [t/h]	208.18	214.74	138.75	222.55	208.18
MEA make-up [t/h]	0.13	0.13	0.09	0.13	0.13
Process water make-up [t/h]	45.96	42.98	30.63	45.96	45.96
NaOH solution for $DeSO_x$ [t/h]	0.09	0.09	0.05	0.09	0.09
Ammonia solution for SNCR [t/h]	0.62	0.63	0.62	0.62	0.62
Operating labor - number of persons	120	120	120	120	120

Table 5.5. Economic results for cement kiln with MEA CO_2 capture.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Variable OPEX [€/t_{clk}]	**55.2**	**54.8**	**44.5**	**56.2**	**44.5**
Raw meal	5.0	5.0	5.0	5.0	5.0
Coal	9.4	9.4	9.4	9.4	9.4
Electricity	14.2	14.1	12.1	15.2	14.2
Steam	19.8	19.4	13.0	19.8	9.1
Cooling water make-up	0.7	0.7	0.5	0.7	0.7
MEA make-up	1.6	1.6	1.1	1.6	1.6
Process water make-up	2.5	2.4	1.7	2.5	2.5
NaOH solution for $DeSO_x$	0.3	0.3	0.2	0.2	0.3
Ammonia for SNCR	0.7	0.7	0.7	0.7	0.7
Other variable costs	1.1	1.1	1.1	1.1	1.1
Fixed OPEX [€/t_{clk}]	**23.6**	**23.7**	**22.7**	**23.9**	**23.6**
Operative, administrative and support labor	10.6	10.7	10.5	10.6	10.6
Insurance and local taxes	5.8	5.8	5.4	5.9	5.8
Maintenance cost (including maintenance labor)	7.3	7.3	6.8	7.4	7.3
Total plant costs, CO_2 capture [M€]	**76.1**	**74.3**	**58.4**	**81.8**	**76.1**
TDC without process contingencies	47.6	46.5	36.5	52.0	47.6
Process contingencies	8.4	8.2	6.4	8.1	8.4
Indirect costs	7.8	7.7	6.0	8.4	7.8
Owner's costs	3.9	3.8	3.0	4.2	3.9
Project contingencies	8.4	8.2	6.4	9.0	8.4

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Total plant costs, cement kiln + CO_2 capture [M€]	**279.9**	**278.0**	**262.2**	**285.6**	**279.9**
Total OPEX [€/t_{clk}]	**78.8**	**78.6**	**67.3**	**80.1**	**68.1**
Total CAPEX [€/t_{clk}]	**28.6**	**28.7**	**26.8**	**29.2**	**28.6**
Cost of clinker [€/t_{clk}]	**107.4**	**107.3**	**94.0**	**109.3**	**96.8**
Cost of cement [€/t_{cement}]	**79.0**	**78.9**	**69.1**	**80.4**	**71.1**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**80.2**	**79.8**	**83.9**	**84.2**	**56.9**

A break-down of the cost of CO_2 avoided for all MEA cases is presented in Figure 5.7, together with the cost of clinker.

The low air leak case, where the CO_2 concentration is slightly higher compared to the base case and the flue gas flow is slightly lower, performs similarly to the base case in terms of costs. A small decrease in the variable operating costs is observed while fixed operating costs and CAPEX are slightly higher compared with the base case, due to the slightly reduced capacity factor of the plant, from 91.3 to 90.5%, because of increased yearly maintenance.

The optional extent of capture case, with 60% CO_2 avoided from flue gases, results in lower cost of clinker compared to the base case, as absolute costs are decreased (both capital and operating costs). However, the CO_2 avoidance costs increases by nearly 5%. This is mainly explained by the economy of scale of the capital cost and consequently the fixed operating cost, i.e. the capital cost does not decrease linearly with the decreased flow of flue gases treated in the CO_2 capture process. On the other hand, a small decrease in variable operating costs is observed, which is mainly a result of a larger share of the total steam requirement originating from waste heat recovery, compared to the base case.

The ship transport case shows a slightly higher cost of clinker and cost of CO_2 avoided than the base case, due to higher capital costs and power consumption of the CO_2 liquefaction process compared with the compression process for pipeline transport.

The case of steam imported from a nearby coal-fired power plant results in a 29% lower cost of CO_2 avoided compared with the base case. In this case, both the lower unit cost of steam as well as the lower CO_2 emissions related to steam production have a large impact on the CO_2 avoidance cost. A smaller decrease is observed in the cost of clinker, around 10%, as the unit cost of steam is the only cost factor that differs (*cf.* Table 5.5)

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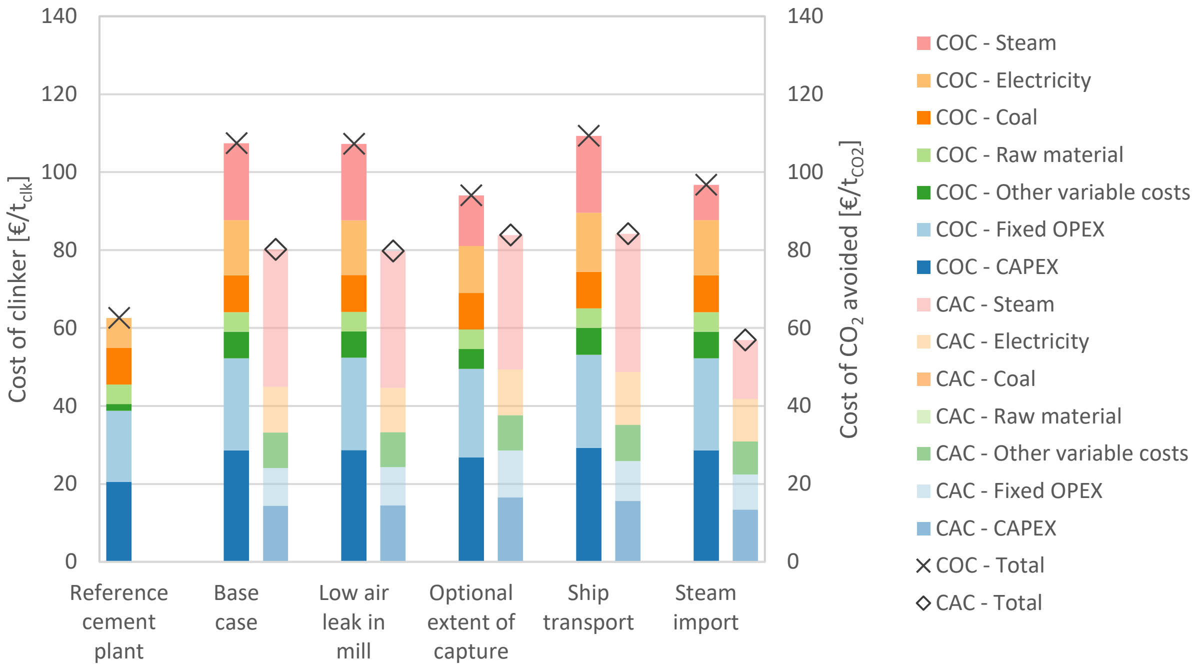

Figure 5.7. Breakdown of the cost of clinker and cost of CO_2 avoided for all MEA cases.

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6 OXYFUEL PROCESS

In the oxyfuel technology, combustion is performed with an oxidizer consisting mainly of oxygen and CO_2, to produce a CO_2 rich flue gas which allows a relatively easy purification with a CO_2 purification unit (CPU). As opposed to the MEA technology, which is an end-of-pipe technology (Figure 5.1), the cement kiln itself is modified when the oxyfuel technology is integrated into the kiln (Figure 6.1). Additional power is needed for an air separation unit (ASU) and for the CPU, but some of this power demand can be covered with an organic Rankine cycle (ORC) generating power from waste heat.

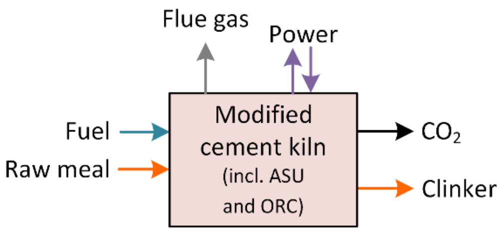

Figure 6.1. Schematic overview of the integration of the oxyfuel process with a cement kiln.

6.1 Process description

The concept of the oxyfuel process is illustrated in Figure 6.2, and more detailed process flowsheet of the process as well as the CPUs for pipeline and ship transport are given in Figure 6.3 - Figure 6.5. Several of the core process units in the oxyfuel cement kiln differ substantially from the conventional reference plant. The main units that must be modified, replaced or added include:

- clinker cooler, with cooler gas recirculation
- exhaust gas recirculation system
- gas-gas heat exchangers
- condenser
- ASU
- oxygen blower
- CPU
- rotary kiln main burner for oxyfuel combustion
- waste heat recovery system (ORC)
- particle removal units upstream of the ORC and in the cooler recycle loop

Many of the existing components in the cement plant can be used in the oxyfuel plant, such as the rotary kiln itself, the calciner and the preheating tower. Furthermore, it can be concluded from oxyfuel burner tests performed within CEMCAP [CAR, 2018] that the existing main burner can be used. The existing coal mill and handling system can also be used in the oxyfuel plant, as the total amount of fuel flow is almost unchanged compared to the reference cement plant. Existing ID, raw mill and filter fans can most likely be reused but the treated gas flows and their power consumption will be slightly changed.

Oxygen from the ASU mixed with recirculated CO_2 rich flue gas forms the so-called oxidizer gas stream. The oxidizer is first fed to the clinker cooler where it is preheated while it also cools the

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clinker. A part of the preheated air is then sent to the rotary kiln main burner and the calciner, while the other part preheats air to be sent to the raw mill, before it is recycled back to the clinker cooler. The flue gas produced from combustion in the main burner and the calciner, mixed together with the gaseous reaction products of the calcination process that takes place in the calciner, has a high concentration of CO_2.

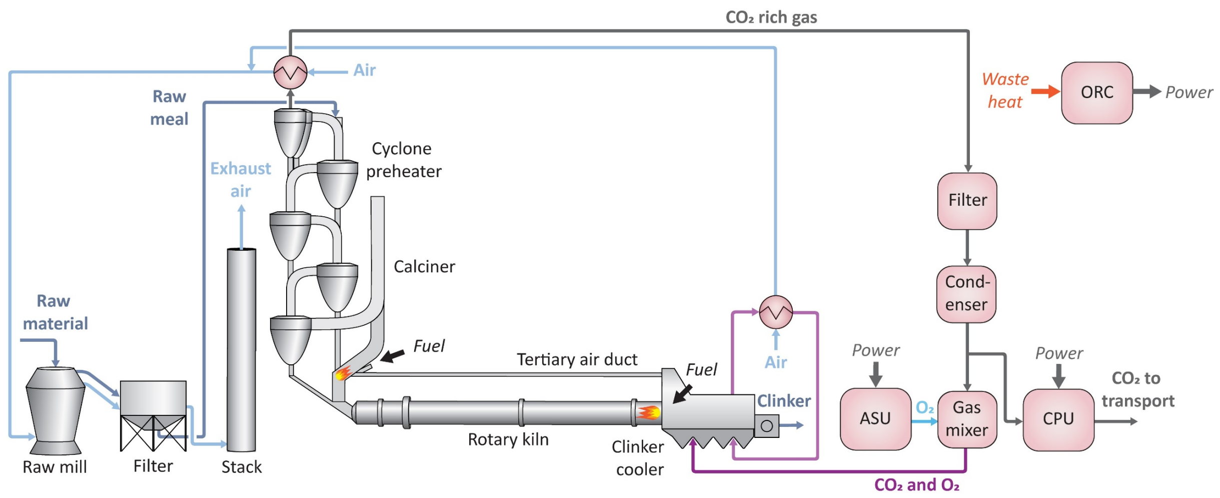

Figure 6.2. Reference cement kiln with oxyfuel CO_2 capture.

Right after the preheater exit, heat is recovered from the hot flue gases in a two-stage heat exchanger, with hot-oil as an intermediate working fluid, for drying of raw material in the raw mill and for electric power generation in an ORC. The slightly cooled flue gas is then sent to a filter for dedusting before water is removed from the flue gas in a condenser. Thereafter, a part of the flue gas is recycled and mixed with oxygen, and the rest is sent to a CPU to be conditioned before transport and storage. Excess heat from the CPU is also recovered and utilized for electricity generation in the ORC.

For pipeline transport, the CPU is designed as a single flash, self-refrigerated unit which delivers compressed CO_2 at 110 bar (see Figure 6.4). The CPU includes compression with intercooling in four stages and drying of the CO_2 stream with molecular sieves prior to cooling and liquefaction of CO_2 in a multi-stream heat exchanger. For ship transport, the CO_2 product should be at 6.5 bar and subcooled with 3 °C (around -52 °C). The CPU design differs slightly from the pipeline transport design in that a second liquid-vapor separation stage is added at the target pressure, and external refrigeration is needed to reach the low temperature required (see Figure 6.5).

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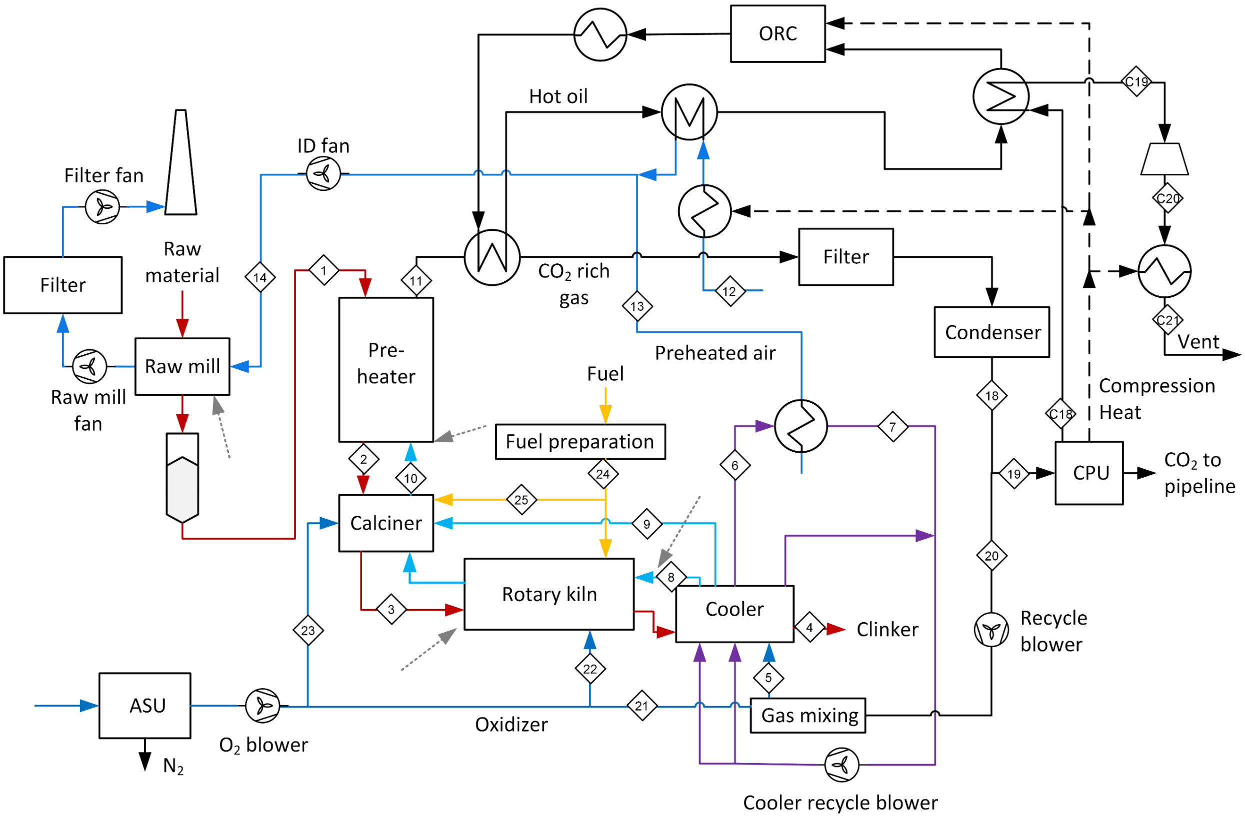

Figure 6.3. Process flowsheet of the oxyfuel cement process with hot oil heat recovery system. Grey dotted arrows indicate air leak.

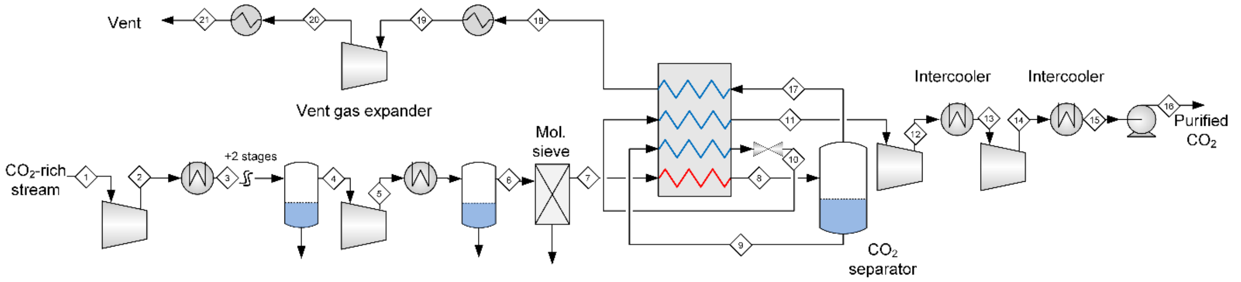

Figure 6.4. Process flowsheet of CPU for pipeline transport as applied to the oxyfuel technology.

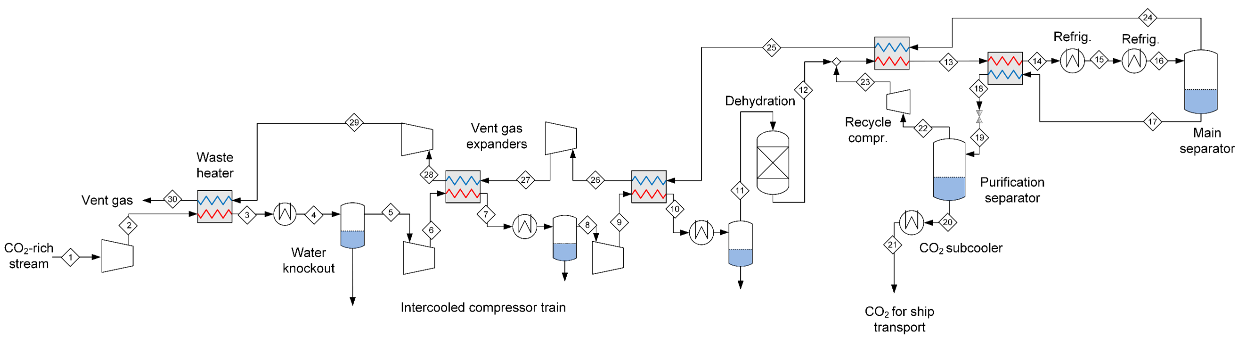

Figure 6.5. Process flowsheet of CPU for ship transport as applied to the oxyfuel technology.

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6.2 Process evaluation

The core oxyfuel cement process is simulated with VDZs cement process model, which has been developed since 2002. The model is based on a reference cement plant model [LOC, 2002a; LOC, 2011], has been further developed and applied in VDZ projects [LOC, 2002b; KLE, 2006; LOC, 2009b; LOC, 2009a], and has been adapted to the oxyfuel process as part of the work in the ECRA CCS project. This adaptation is described by a PhD-Thesis [KOR, 2013; FLE, 2014] and the ECRA CCS project reports II and III [ECR, 2009; ECR, 2012].

The VDZ process model is divided in several units with connected balance boundaries. The modelling is based on material and energy balances carried out on the material flows [LOC, 2011]. For the solids a distinction is made between the raw material components, the clinker phases, and the chemical constituents of the fuels. Comparable consideration is also given to the individual gas constituents. The model simulates the clinker burning process with the help of unit operations, which calculate e.g. the heat and mass transfer in the respective defined balance boundaries. Steady-state operating conditions of preheater, calciner, kiln, and cooler are each determined in steps using iterative calculation procedures.

Within CEMCAP the simulation of the oxyfuel process developed in the ECRA CCS project has been tuned with results from the tested performance of the oxyfuel clinker cooler, the oxyfuel burner and the oxyfuel calciner. Parameters such as flame length, primary and secondary "air" flow rates, and primary and secondary "air" O_2 concentrations are tuned based on the burner experiments. Calcination start and end temperatures and residence time are set based on the calciner experiments. Air leakage into the clinker cooler was set based on the clinker cooler experiments [JAM, 2018]. The additional air leakage added in the cooler resulted in an overall air leak of 6.3%.

The waste heat recovery system is designed and simulated using the Sequential Framework for heat exchanger network synthesis, which was developed at NTNU [ANA, 2011]. The CPU is simulated using Aspen HYSYS V9, as a single stage flash self-refrigerated unit (Figure 6.4). The power consumption of the CPU is around 0.4 MJ/kg_{CO2}. The heat required for dehydration in the CPU is assumed to be 16.6 MJ_{th}/kg_{H2O} (estimated based on [KEM, 2014]) and provided by electric heaters. The power consumption of the ASU is 226 kWh/t_{O2}. In addition to this the ASU dehydration heat demand is 58.3 kJ/kg_{O2} and assumed to be provided by electric heaters. Further details about the design and operating parameters of the waste heat recovery system and the CPU is described by Jamali et al. [JAM, 2018].

In addition to the additional power demand of the ASU and CPU, the overall power demand related to other utilities at the plant will be slightly changed compared to the reference kiln (*cf.* Table 3.2). It is assumed that the existing ID fan, raw mill fan and filter fan can be reused for the oxyfuel technology, but the power consumption decreases from 22.1 kWh/t_{clk} to 15.8 kWh/t_{clk} due to lower flow rates. Cooler fans will be replaced by O_2 blowers, a recycle blower and a cooler recycle blower with total electric power demand of 15.0 kWh/t_{clk}. The power consumption of raw meal and cement grinding, solids handling, kiln drive, lighting etc. will be unchanged compared to the reference kiln. The total power consumption of these utilities will be 138.7 kWh/t_{clk} instead of 132 kWh/t_{clk}.

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Key process characteristics of the oxyfuel base case is shown in Table 6.1. Detailed stream data are provided in Section 3 in the Supplementary information.

Table 6.1. Key process characteristics and simulation results for oxyfuel cases.

Case	Base case	Low air leak in core process	High air leak in core process	No air leak in cooler	Ship transport
False air at preheater outlet (Nm^3/ Nm^3) [%]	6.3	4.6	8.1	6.0	6.3
Clinker production [t/h]	125	125	125	125	125
Generated CO_2 flow rate [t/h]	110	110	111	110	110
Captured CO_2 flow rate [t/h]	99	99	100	99	99
Captured and conditioned CO_2 temperature [°C]	36.4	36.4	36.4	36.4	-54.9
Captured and conditioned CO_2 mole fraction (wet basis)	0.973	0.978	0.965	0.975	0.999
Coal input [MW]	109	108	110	109	109
CPU power demand [kJ/kg_{CO2}]	440	431	453	435	649
O_2 flow rate [t/h]	15.7	15.7	15.7	15.7	15.7
ASU power demand [kWh/t_{O2}]	226	226	226	226	226
ASU dehydration heat demand [kJ/kg_{O2}]	58.3	58.3	58.3	58.3	58.3
Cooling system power demand [kJ/kg_{CO2}]	0.03	0.03	0.04	0.03	0.04
Total power demand [MW]	35.1	34.8	35.2	34.9	41.1
Power generation in ORC cycle [MW]	2.9	3.0	2.8	2.9	2.9
Cooling demand [MW]	47.2	46.0	47.4	46.8	60.9

CO_2 capture ratio and CO_2 avoided for the oxyfuel cases are shown in Table 6.2. SPECCA values are presented in Figure 6.6, and a break-downs of the different contributions to the SPECCA values are presented in Table 6.3. The following can be noticed:

- Consumption of electric power in various parts of the oxyfuel process is by far the largest factor contributing to the added equivalent specific primary energy consumption. A break-down of the electricity consumption shows that the CPU unit is responsible for the largest part (nearly 60% in the base case and 65% in the ship transport case). The ASU is responsible for the second largest part (35% in the base case).
- The investigated air leak cases (4.6–8.1% false air at preheater outlet) show only small differences compared to the base case SPECCA. In the case of low air leak in the core process and in the case of no air leak in the cooler, a small reduction in the added equivalent specific coal and electricity consumption results in a slightly lower SPECCA. A small increase is observed for the case of high air leak in the core process, although this case has a slightly higher electric power generation than the base case due to higher gas flows.
- The ship transport case results in around 32% increase in added equivalent specific energy consumption. This is due to the increased electric power demand of the CPU and related to increased cooling demand compared to the base case. The increased electricity consumption also reduces the equivalent specific CO_2 avoided. Altogether the SPECCA increased with nearly 35%.

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Table 6.2. CO_2 capture ratio and CO_2 avoided for oxyfuel cases.

	Base case	Low air leak in core process	High air leak in core process	No air leak in cooler	Ship transport
CO_2 capture ratio	90%	90%	90%	90%	90%
CO_2 avoided from flue gas	90%	90%	90%	90%	90%
CO_2 avoided	90%	90%	90%	90%	90%
Equivalent CO_2 avoided	82%	82%	82%	82%	80%

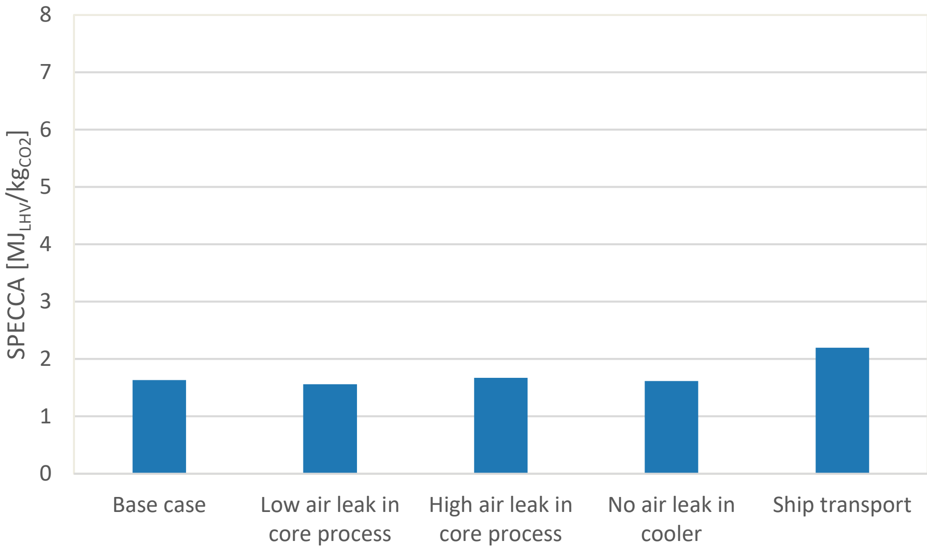

Figure 6.6. SPECCA for oxyfuel cases.

Table 6.3. Break-down of SPECCA for oxyfuel base case.

	Base case	Low air leak in core process	High air leak in core process	No air leak in cooler	Ship transport
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**1173**	**1124**	**1242**	**1163**	**1551**
Coal	4	-24	33	4	4
Electric power consumption					
Fans, milling, and handling of solids in kiln	-62	-62	-62	-62	-62
Additional fans	122	121	123	122	122
ASU	473	473	473	473	473
CPU	759	744	790	749	1119
Cooling water system	59	58	59	59	76
Electric power generation	-182	-185	-174	-181	-182
Equivalent specific CO_2 avoided [kg_{CO2}/t_{clk}]	**719**	**720**	**718**	**719**	**706**
At cement kiln stack	758	758	758	758	758
Electric power consumption					

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Fans, milling, and handling of solids in kiln	2	2	2	2	2
Additional fans	-4	-4	-4	-4	-4
ASU	-16	-16	-16	-16	-16
CPU	-25	-25	-26	-25	-37
Cooling water system	-2	-2	-2	-2	-3
Electric power generation	6	6	6	6	6
SPECCA [$MJ_{LHV}/kgCO_2$]	**1.63**	**1.56**	**1.73**	**1.62**	**2.20**

6.3 Economic evaluation

The utility consumption, other consumables and the required operating labor of the cement kiln with oxyfuel CO_2 capture are presented in Table 6.4. Economic results in terms of both capital and operating costs are presented in Table 6.5, together with the resulting cost of clinker and CO_2 avoided cost. Detailed equipment lists are provided in Section 3 in Supplementary information.

The process contingency of the technology is set to 42%. Key technologies such as oxyfuel burner, oxyfuel calciner, and oxyfuel clinker cooler have through CEMCAP been successfully tested at a pilot scale [CAR, 2018; PAN, 2018; LIN, 2017]. However, operation of an oxyfuel rotary kiln, and operation of the complete system with all subsystems together is not yet possible to test at this scale. The contribution to the process contingency based on technological maturity ($C_{process,1}$) is therefore set to 30 %-points. The level of detail of the equipment list is moderate, so the contribution to process contingency based on this ($C_{process,2}$) is set to 12 %-points, resulting in the overall value of 42%.

Table 6.4. Utilities, consumables and operating labor for oxyfuel cement kiln.

	Base case	Low air leak in core process	High air leak in core process	No air leak in cooler	Ship transport
Coal [t/h]	14.45	14.32	14.59	14.45	14.45
Electricity [MW]	35.09	34.75	35.18	34.92	41.09
Cooling water make-up [t/h]	104.54	100.98	104.06	102.75	133.64
Ammonia solution for SNCR [t/h]	0.60	0.60	0.60	0.60	0.60
Operating labor - number of persons	120	120	120	120	120

Table 6.5. Economic results for oxyfuel cement kiln.

	Base case	Low air leak in core process	High air leak in core process	No air leak in cooler	Ship transport
Variable OPEX [€/t_{clk}]	**32.8**	**32.5**	**32.9**	**32.7**	**35.7**
Raw meal	5.0	5.0	5.0	5.0	5.0
Coal	9.4	9.3	9.5	9.4	9.4
Electricity	16.3	16.2	16.4	16.2	19.1
Cooling water make-up	0.3	0.3	0.3	0.3	0.4
Ammonia for SNCR	0.7	0.7	0.7	0.7	0.7
Other variable costs	1.1	1.1	1.1	1.1	1.1
Fixed OPEX [€/t_{clk}]	**25.3**	**25.2**	**25.7**	**25.3**	**26.2**
Operative, administrative and support labor	10.4	10.4	10.4	10.4	10.4
Insurance and local taxes	6.6	6.6	6.8	6.6	7.0
Maintenance cost (including maintenance labor)	8.3	8.3	8.5	8.3	8.8

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Total plant costs, CO_2 capture [M€]	**127.8**	**126.2**	**137.0**	**127.4**	**146.3**
TDC without process contingencies	73.1	72.3	77.9	72.8	84.8
Process contingencies	20.9	20.5	22.8	20.9	22.8
Indirect costs	13.2	13.0	14.1	13.1	15.1
Owner's costs	6.6	6.5	7.1	6.6	7.5
Project contingencies	14.1	13.9	15.1	14.1	16.1
Total plant costs, cement kiln + CO_2 capture [M€]	**331.5**	**330.0**	**340.7**	**331.1**	**350.0**
Total OPEX [€/t_{clk}]	**58.1**	**57.7**	**58.6**	**58.0**	**61.8**
Total CAPEX [€/t_{clk}]	**35.0**	**34.8**	**36.0**	**35.0**	**37.0**
Cost of clinker [€/t_{clk}]	**93.1**	**92.6**	**94.6**	**92.9**	**98.8**
Cost of cement [€/t_{cement}]	**68.4**	**68.1**	**69.6**	**68.3**	**72.7**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**42.4**	**41.7**	**44.4**	**42.2**	**51.3**

Break-downs of the cost of CO_2 avoided for all oxyfuel cases are presented in Figure 6.7, together with the cost of clinker in each case. It is evident from the various air leak cases that the effect of air leak on the process economics are relatively small – <5% difference in the cost of CO_2 avoided and <2% difference in the cost of clinker compared to the base case.

The ship transport case results in the largest cost difference out of the investigated cases. The increase in cost of CO_2 avoided is due to higher capital costs and electricity consumption related to the CPU. This is mainly because the CPU for ship transport requires an external refrigeration system unlike the self-refrigerated CPU used for pipeline transport in the other cases.

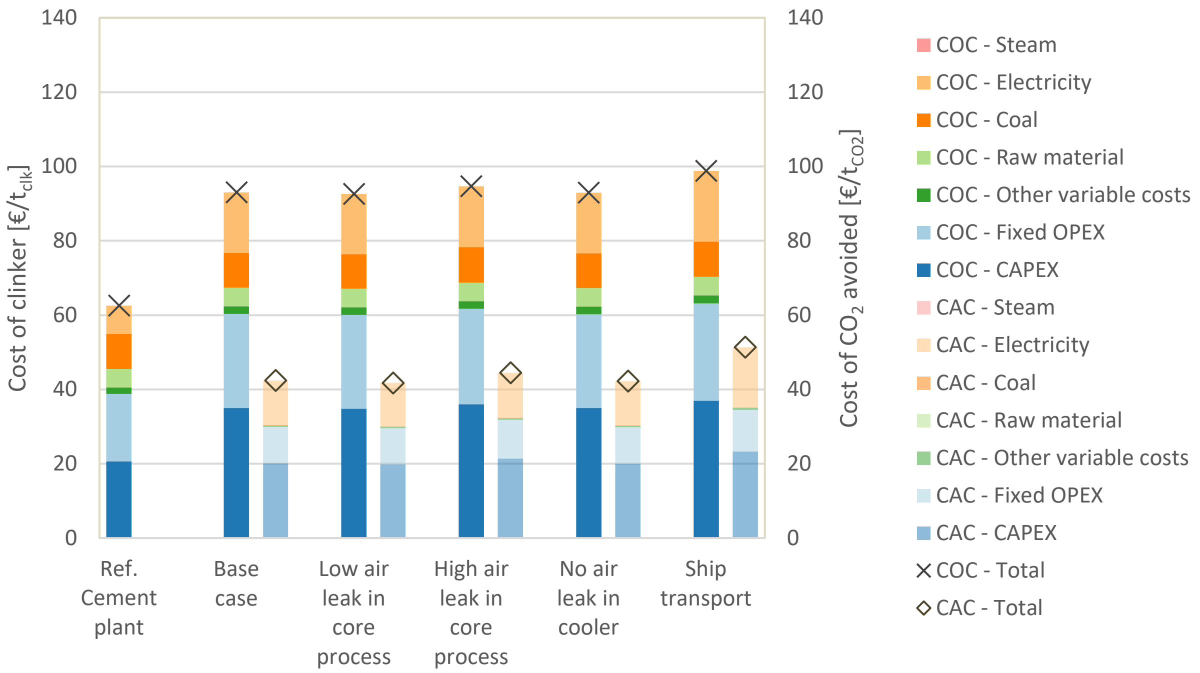

Figure 6.7. Breakdown of the cost of clinker and cost of CO_2 avoided for all oxyfuel cases.

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7 CHILLED AMMONIA PROCESS

In the chilled ammonia process (CAP), CO_2 is removed from flue gas by absorption with chilled ammonia as solvent. It is an end-of-pipe technology that can be integrated with a cement kiln as shown in Figure 7.1. Flue gas is sent to the capture process where CO_2 is captured by use of heat for solvent regeneration and power for chilling, pumping and compression. Waste heat from the cement plant can be utilized to cover a part of the heat demand.

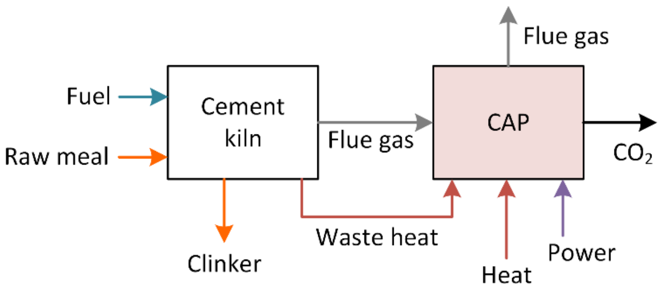

Figure 7.1. Schematic overview of the integration of CAP CO_2 capture with a cement kiln.

7.1 Process description

The concept of the CAP is illustrated in Figure 7.2, and more detailed process flowsheets of the CAP and the CO_2 conditioning are shown in Figure 7.3 - Figure 7.5. The cement plant flue gas is first cooled in a direct contact cooler to limit the water content in the flue gas entering the CO_2 absorber. In addition, SO_x is removed in the direct contact cooler by reaction with aqueous ammonia (down to below 1 ppm in order to avoid the interaction of SO_x with ammonia within the CO_2 absorber). There is no need to consider NO_x removal for this process [PER, 2018]. The cooled flue gas is sent to the CO_2 absorber at a temperature below 25°C, where CO_2 is absorbed by chilled ammonia solution. The temperature in the absorber is controlled by a CO_2-rich solvent pump-around that is chilled down to temperatures around 12-13°C. The CO_2–rich aqueous ammonia stream leaving the bottom of the CO_2 absorber is heated in the lean-rich heat exchanger and regenerated in the CO_2 desorber operating at 20 bar. NH_3 is removed from the CO_2 gas stream leaving the CO_2 desorber in a CO_2-water wash unit, and the purified CO_2 is compressed or liquefied to meet export specifications.

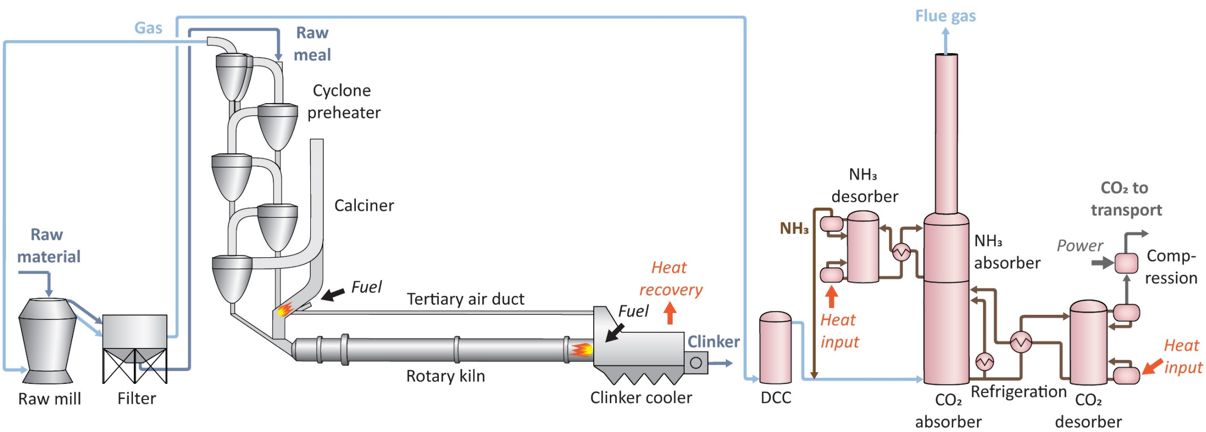

Figure 7.2. Reference cement kiln with CAP CO_2 capture.

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NH_3 is removed from the decarbonized flue gas in a water wash column using chilled water as absorbent in the NH_3 absorber. The recovered ammonia is stripped from the water in a dedicated desorber and re-used in the CO_2 capture section. The NH_3 content in the purified flue gas leaving the top of the NH_3 absorber is further decreased to values below 10 ppm$_v$ using sulfuric acid and heated in a direct contact heater utilizing heated water from the direct contact cooler. The purified flue gas is eventually sent to the stack at a temperature of approximately 40-50°C, at saturated conditions. In order to prevent corrosion of the stack derived from the condensation of impurities contained in the CO_2 and NH_3-depleted flue gas, a new stack built with chemically compatible materials is required. An appendix stripper working at ambient pressure recovers NH_3 and CO_2 from H_2O purge from the CO_2 desorber and the CO_2 water wash.

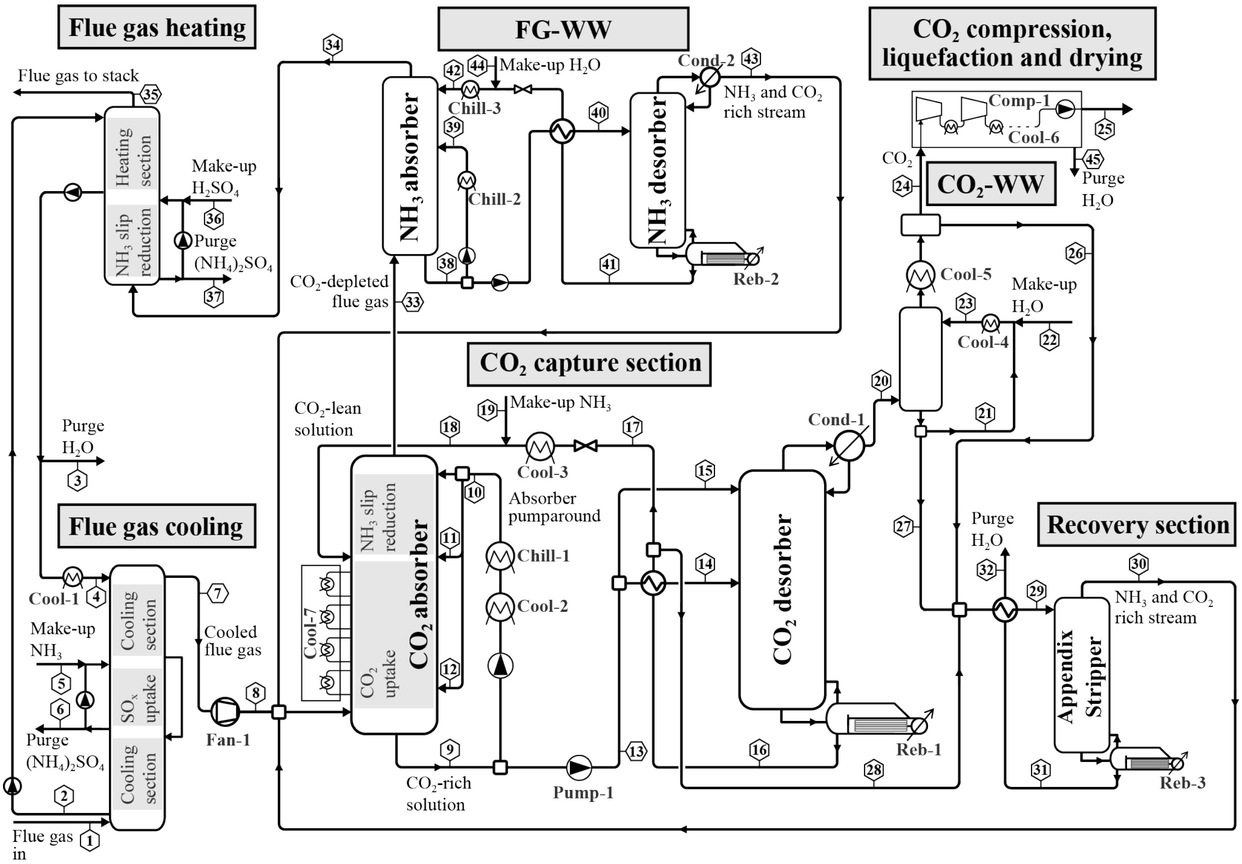

Figure 7.3. Process flowsheet of the chilled ammonia CO_2 capture process.

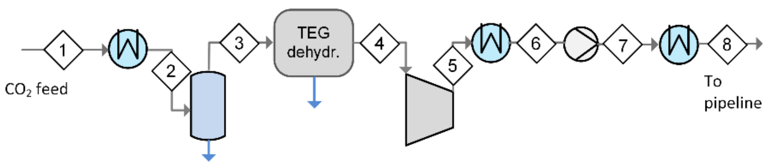

Figure 7.4. Process flowsheet of CO_2 compression for pipeline transport.

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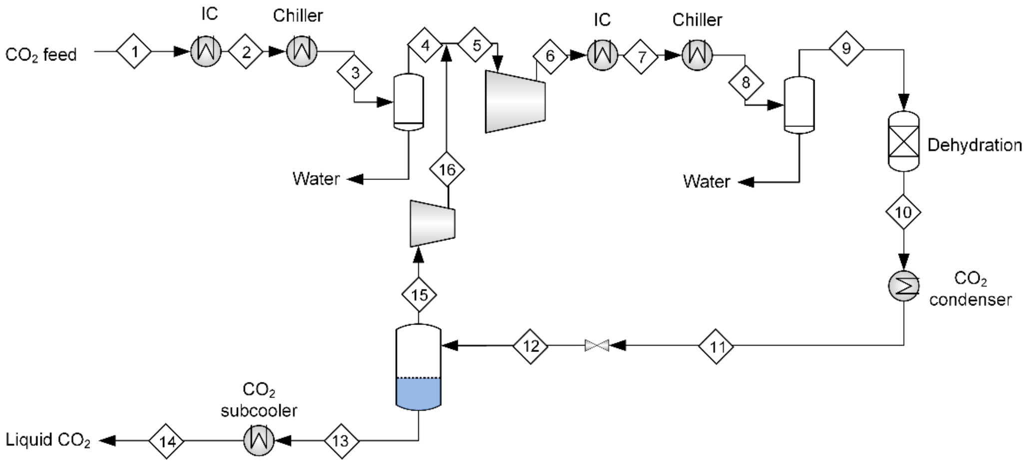

Figure 7.5. Process flowsheet of CO_2 liquefaction for ship transport.

7.2 Process evaluation

The chilled ammonia process is simulated and optimized using Aspen Plus V8.6 with the Thomsen thermodynamic model. Results from pilot plant tests of the process are used as an input for tuning of the process model. The process optimizations are done with minimization of SPECCA as objective function. A detailed description of the simulation and optimisation procedure of the core process can be found in D10.3 "CAP process optimisation and comparison with pilot plant tests" [PER, 2018].

In the base case, where the air leak in the raw mill increases from low to medium over the year, the columns are designed to handle medium air leak which is the constraining case due to higher flue gas flow rates. The system is then simulated for operation with both medium air leak and low air leak. To adapt to changing flue gas conditions, the following parameters are varied: (i) the ammonia concentration of the CO_2-lean ammonia stream, (ii) the CO_2 loading of the CO_2-lean ammonia stream, (iii) the CO_2-lean stream flowrate to inlet flue gas flowrate ratio in the CO_2 absorber, (iv) the CO_2 absorber pumparound split fraction, (v) the chilling temperature of the pumparound, (vi) the CO_2 desorber pressure and (vii) the cold-rich split fraction before the rich/lean heat exchanger. The resulting optimum sets of operating conditions for all CO_2 capture cases are presented in Table 7.1, except for the ship transport case, where the CAP operating conditions are identical to the base case.

Table 7.1. Optimum sets of operating conditions that minimize SPECCA for CAP cases. The ship transport case conditions are identical to that of the base case.

Case	**Base case**		**Low air leak in mill**	**Optional extent of capture**		**Steam import**	
Design	Medium air leak		Low air leak	Medium air leak		Medium air leak	
Operation	Low air leak	Medium air leak	Low air leak	Low air leak	Medium air leak	Low air leak	Medium air leak
CO_2-lean NH_3 concentration [mol_{NH3}/kg_{H2O}]	5	6	4	4	5	5	6

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CO_2-lean CO_2 loading [mol_{CO2}/mol_{NH3}]	0.415	0.387	0.340	0.364	0.346	0.415	0.387
CO_2-lean to inlet flue gas flowrate ratio [kg/kg]	9.0	6.5	7.0	7.0	5.0	9.0	6.5
Pumparound split fraction [%]	0.12	0.18	0.15	0.12	0.21	0.12	0.18
Pumparound temperature [°C]	12	9	12	13	12	12	9
CO_2 desorber pressure [bar]	25	25	20	20	20	10	10
Cold-rich split fraction [%]	0.025	0.035	0.04	0.035	0.04	0.0375	0.05

In the low air leak case, the columns are designed for low air leak and only operation with low air leak is simulated. For the steam import case, the optimization resulted in the same process as the base case, except for the pressure of the CO_2 desorber. The reason for this is that the pressure of the CO_2 desorber shows opposing effects on the value of the SPECCA index depending on the origin of the steam used in the reboilers of the CAP: While operating the CO_2 desorber at higher pressure decreases the SPECCA when the steam is generated in an on-site NG boiler, it increases the SPECCA when the steam is imported from a CHP plant. Increasing pressure in the CO_2 desorber, up to 20-25 bar, decreases the specific reboiler duty of the CO_2 desorber, although it leads to a higher reboiler temperature, and hence, to a higher temperature of the required steam. On the one hand, the specific energy consumed in the NG boiler, and hence the amount of natural gas burnt, does not depend on the temperature of the steam required, which makes the associated SPECCA independent of the reboiler temperature. On the other hand, when the steam is provided by a CHP plant, the amount of electricity lost in the CHP plant increases with the higher temperature and pressure of the withdrawn steam, which in turn increases the associated SPECCA value in this scenario.

In all cases, waste heat from the cement plant is used to cover all the heat required for the reboiler in the NH_3 desorber and the appendix stripper (steam at ~110 °C), and a small part of the remaining heat demand (steam at ~145 °C for the CO_2 desorber). Refrigeration of CO_2 absorber pumparound to 12-13 °C and of the water stream entering the top of the NH_3 absorber to 15 °C is required, and for estimation of power consumption and cooling demand of the refrigeration system, COPs of around 7 and 8 are assumed, in accordance with the CEMCAP framework [VOL, 2018].

Compression of the captured CO_2 for pipeline transport is simulated in accordance with the processes described in the CEMCAP framework. Since the captured CO_2 is at 19.5 bar, only one compression stage in addition to the final pumping is required in the compression process.

Key process characteristics and simulation results for the CAP cases are given in Table 7.2. These results have been obtained after optimizing the operating conditions of the process to minimize the SPECCA. Stream data are provided in Section 4 in the Supplementary information.

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Table 7.2. Key process characteristics and simulation results for CAP cases.

Case	**Base case**		**Low air leak in mill**	**Optional extent of capture**		**Ship transport**		**Steam import**	
Design	Medium air leak		Low air leak	Medium air leak		Medium air leak		Medium air leak	
Operation	Low air leak	Medium air leak	Low air leak	Low air leak	Medium air leak	Low air leak	Medium air leak	Low air leak	Medium air leak
Captured CO_2 flow rate [t/h]	91.9	92.0	91.8	86.6	88.0	91.9	92.0	91.9	91.8
Captured CO_2 mole fraction (wet basis)	99.9	99.9	99.9	99.9	99.9	1.0	1.0	99.9	99.9
CO_2 capture rate [%]	90	90	90	85	86	90	90	90	90
Specific reboiler duty CO_2 desorber [MJ/kgCO$_2$]	2.12	2.20	2.13	2.10	2.13	2.12	2.20	2.26	2.36
Reboiler temperature CO_2 desorber [°C]	155.5	157.0	158.8	156.5	156.6	155.5	157.0	134.5	136.1
Specific reboiler duty NH_3 desorber [MJ/kgCO$_2$]	0.13	0.18	0.11	0.13	0.18	0.13	0.18	0.13	0.18
Reboiler temperature NH_3 desorber [°C]	108.8	108.8	108.8	108.8	108.8	108.8	108.8	108.8	108.8
Specific reboiler duty appendix desorber [MJ/kg CO_2]	0.05	0.07	0.05	0.04	0.06	0.05	0.07	0.05	0.08
Reboiler temperature appendix desorber [°C]	115.0	115.0	115.0	115.0	115.0	115.0	115.0	115.0	115.0
Steam generation from waste heat [%]	8	7	8	9	8	8	7	8	7
Steam generation in NG boiler [%]	92	93	92	91	92	92	93	0	0
Steam import [%]	0	0	0	0	0	0	0	92	93
Specific power consumption [MJ/kg$_{CO2}$]	0.315	0.338	0.300	0.298	0.315	0.328	0.351	0.360	0.390
Cooling water demand [MJ/kg$_{CO2}$]	3.23	3.40	3.20	3.21	3.31	3.45	3.62	3.41	3.61
Process water make-up [kg/kg$_{CO2}$]	0.009	0.014	0.01	0.008	0.014	0.009	0.014	0.012	0.018
Solvent makeup [kg/kg$_{CO2}$]	0.002	0.003	0.002	0.002	0.003	0.002	0.003	0.002	0.003
H_2SO_4 consumption [g/kg$_{CO2}$]	0.6	1.0	0.4	0.5	0.8	0.6	1.0	0.6	1.0

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CO_2 capture ratio and CO_2 avoided for the chilled ammonia cases are shown in Table 7.3. SPECCA values are presented in Figure 7.6, and break-downs of the different contributions to the SPECCA values are presented in Table 7.4. The following can be noticed:

- As for MEA, the two factors contributing to the SPECCA are steam and power consumption, where steam is responsible for the largest part. Both the steam and power demands are however reduced considerably compared to MEA, leading to SPECCA values almost halved compared to MEA.
- The low air leak case shows an improved process performance compared with the base case and around 6% lower SPECCA. This is explained by the increased CO_2 concentration in the flue gases leading to lower reboiler duties in the CO_2 and NH_3 desorbers as well as lower specific electric power consumption. For the CAP, further decrease of air leak in the cement kiln would be beneficial for the technical performance.
- In the optional extent of capture case, the constraint of CO_2 avoided from the flue gas is removed and the chilled ammonia process optimized to minimize SPECCA. The optimization results show a trend towards lower SPECCA for lower capture rates. The trend is relatively flat below 90% capture rate and becomes steeper above 90%. Within the design space of the optimization, a local optimum was found at 86% capture rate. The SPECCA is decreased by 4% compared to the base case. Significantly lower capture rates were not investigated, as they would call for a re-design of the plant including an updated cost assessment.
- The case with steam import from a coal power plant has a SPECCA that is reduced by approximately 35% compared to the base case value. This considerable reduction is explained by the reduced equivalent primary energy consumption and CO_2 emissions of steam imported compared to steam generated in NG boilers. The reduction is however not as large as was seen for the MEA technology. This is mainly due to the lower relative importance of the steam demand on SPECCA of the chilled ammonia process compared to the MEA process. Still, the application of the CAP to cement plants for CO_2 capture will be highly favored by the existence of a CHP plant in vicinity of the cement plant or the building of such plant for the purpose of providing steam for the CAP plant. If steam is generated without the co-generation of electricity, primary energy consumption associated to the steam generation is higher and the SPECCA is increased.

It should be mentioned that the CAP processes evaluated here are with the standard CAP configuration. An advanced CAP configuration with promising energy performance has been developed by ETH as part of the CEMCAP project, and a patent application has been submitted.

Table 7.3. CO_2 capture ratio and CO_2 avoided for chilled ammonia cases.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
CO_2 capture ratio	90%	90%	86%	90%	90%
CO_2 avoided from flue gas	90%	90%	86%	90%	90%
CO_2 avoided	78%	78%	74%	78%	90%
Equivalent CO_2 avoided	73%	73%	69%	73%	74%

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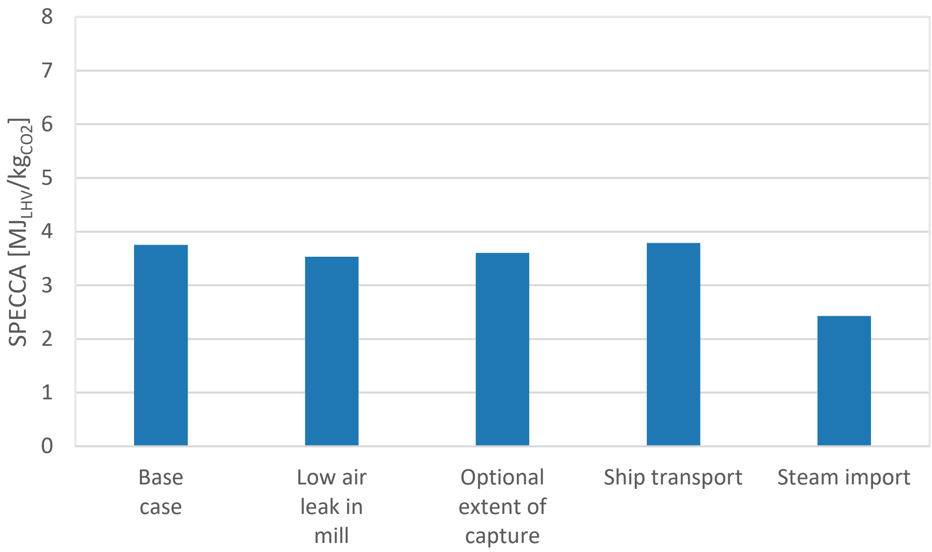

Figure 7.6. SPECCA for chilled ammonia cases.

Table 7.4. Break-down of SPECCA for chilled ammonia cases.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**2401**	**2276**	**2202**	**2422**	**1579**
Steam consumption	1859	1779	1719	1858	958
Electric power consumption					
Pumps, fans, fuel grinding and compressors	207	147	146	207	110
Refrigeration in main process	59	41	41	59	59
Compression/liquefaction for transport	167	204	194	181	336
Cooling water system	110	106	103	117	116
Equivalent specific CO_2 avoided [kg_{CO2}/t_{clk}]	**640**	**645**	**611**	**639**	**651**
At cement kiln stack	792	761	724	762	762
Steam consumption	-104	-100	-96	-104	-91
Electric power consumption					
Pumps, fans, fuel grinding and compressors	-7	-5	-5	-7	-4
Refrigeration in main process	-2	-1	-1	-2	-2
Compression/liquefaction for transport	-6	-7	-6	-6	-11
Cooling water system	-4	-4	-3	-4	-4
SPECCA [$MJ_{LHV}/kgCO_2$]	**3.75**	**3.53**	**3.60**	**3.79**	**2.43**

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7.3 Economic evaluation

The cost estimation of the CAP technology is partly carried out by GE and partly by SINTEF ER. The equipment and installation costs, resulting in total direct costs (TDC) without process contingencies, are provided by GE for the base, low air leak, optional extent of capture and the ship transport case. For the steam import case, the costs for the core CAP plant are assumed to be equal to that of the base case, while the cost for the CO_2 compression train is scaled from the base case costs against its power consumption, with a scaling exponent of 0.65. The equipment and installation costs for CAP are estimated using the same economic assumptions as for the other technologies, presented in Table 2.4. The estimates of total plant costs and all factors contributing to the total operating costs are carried out by SINTEF ER. The utility consumption, other consumables and the required operating labor of the cement kiln with CAP CO_2 capture are presented in Table 7.5. Economic results in terms of both capital and operating costs are presented in Table 7.6, together with the resulting cost of clinker and CO_2 avoided cost. Detailed equipment lists are provided in Section 4 in Supplementary information.

The process contingency of the CAP technology is set to 20%. Full size modules of the complete system are ready to be tested at industrial scale in a real cement plant environment, and the contribution to the process contingency based on technological maturity ($C_{process,1}$) is thus set to 20 %-points. The level of detail of the equipment list is high as it is provided directly from the technology developer, so the contribution to process contingency based on this ($C_{process,2}$) is set to 0 %-points.

Table 7.5. Utilities, consumables and operating labor for cement kiln with CAP.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Clinker production [t/h]	120.65	120.65	120.65	120.65	120.65
Coal [t/h]	13.93	13.93	13.93	13.93	13.93
Electricity [MW]	24.22	23.53	23.32	24.46	25.42
Steam – from waste heat [MW]	4.69	4.70	4.53	4.70	4.69
Steam – from NG boiler [MW]	56.06	53.66	51.84	56.04	0
Steam – import from power plant [MW]	0	0	0	0	56.05
Cooling water make-up [t/h]	185.68	179.32	173.45	198.37	194.04
Process water make-up [t/h]	1.06	0.91	0.96	1.06	1.36
Ammonia make-up [t/h]	0.21	0.17	0.22	0.21	0.21
Ammonia solution for SNCR [t/h]	0.60	0.60	0.60	0.60	0.60
Sulfuric acid for ammonia recovery [t/h]	0.07	0.04	0.06	0.07	0.07
Operating labor - number of persons	120	120	120	120	120

Table 7.6. Economic results for cement kiln with CAP.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport	Steam import
Variable OPEX [€/t_{clk}]	**41.3**	**40.3**	**39.9**	**41.5**	**36.3**
Raw meal	5.0	5.0	5.0	5.0	5.0
Coal	9.4	9.4	9.4	9.4	9.4
Electricity	11.7	11.3	11.2	11.8	12.3
Steam	12.1	11.6	0.6	12.1	6.4
Cooling water make-up	0.6	0.6	0.6	0.6	0.6
Process water make-up	0.1	0.1	0.1	0.1	0.1
Ammonia make-up	0.7	0.6	0.7	0.7	0.7

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Ammonia for SNCR	0.7	0.7	0.7	0.7	0.7
Sulfuric acid for ammonia recovery	0.03	0.01	0.02	0.03	0.03
Other variable costs	1.1	1.1	1.1	1.1	1.1
Fixed OPEX [€/t_{clk}]	**27.3**	**27.4**	**27.2**	**27.9**	**27.6**
Operative, administrative and support labor	10.8	10.9	10.8	10.8	10.8
Insurance and local taxes	7.3	7.4	7.3	7.6	7.5
Maintenance cost (including maintenance labor)	9.2	9.2	9.1	9.5	9.3
Total plant costs, CO_2 capture [M€]	**149.3**	**147.5**	**147.5**	**162.7**	**156.7**
TDC without process contingencies	91.5	90.4	90.4	99.7	96.1
Process contingencies	18.3	18.1	18.1	19.9	19.2
Indirect costs	15.4	15.2	15.2	16.8	16.2
Owner's costs	7.7	7.6	7.6	8.4	8.1
Project contingencies	16.5	16.3	16.3	18.0	17.3
Total plant costs, cement kiln + CO_2 capture [M€]	**353.1**	**351.3**	**351.3**	**366.5**	**360.6**
Total OPEX [€/t_{clk}]	**68.5**	**67.7**	**67.1**	**69.4**	**63.9**
Total CAPEX [€/t_{clk}]	**36.4**	**36.5**	**36.2**	**37.8**	**37.2**
Cost of clinker [€/t_{clk}]	**104.9**	**104.2**	**103.3**	**107.2**	**101.1**
Cost of cement [€/t_{cement}]	**77.1**	**76.6**	**75.9**	**78.8**	**74.3**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**66.2**	**64.5**	**66.6**	**69.8**	**59.2**

A break-down of the cost of CO_2 avoided for all CAP cases is presented in Figure 7.7, together with the cost of clinker in each case. The base, low air leak and optional extent of capture cases all show similar costs. The ship transport case differs from the base case mainly in terms of the CAPEX related costs, and consequently the fixed operating costs. With respect to the cost of CO_2 avoided, the capital costs associated with the CAP technology, and the consequent fixed OPEX are the largest contributors.

In the steam import case, the lower unit cost of steam and the lower CO_2 emissions per unit of steam consumed result in steam related CO_2 avoided cost being reduced by half. On the other hand, there are small increases in electricity and CAPEX related CO_2 avoided costs, due to the larger compression section relative to the base case.

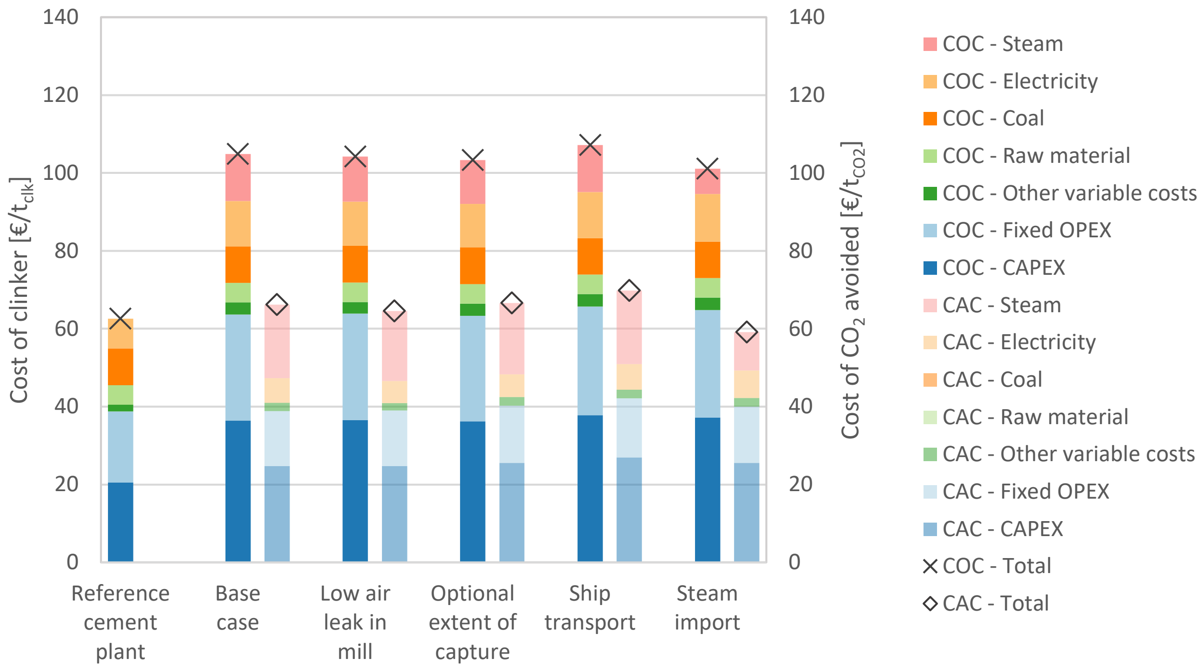

Figure 7.7. Breakdown of the cost of clinker and cost of CO_2 avoided for all CAP cases.

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8 MEMBRANE-ASSISTED CO_2 LIQUEFACTION

In the membrane-assisted CO_2 liquefaction (MAL) process, two different technologies are combined, so that each can carry out a partial separation within its favorable regime of operation. Polymeric membranes are first utilized for bulk separation of CO_2 resulting in a CO_2-rich product. This is sent to a CO_2 liquefaction process, where CO_2 is liquefied, and the more volatile impurity components are removed, resulting in a high purity CO_2 product. The technology is an end-of-pipe technology with no additional integration or feedback to the cement plant, and only power is required as input to the process (Figure 8.1).

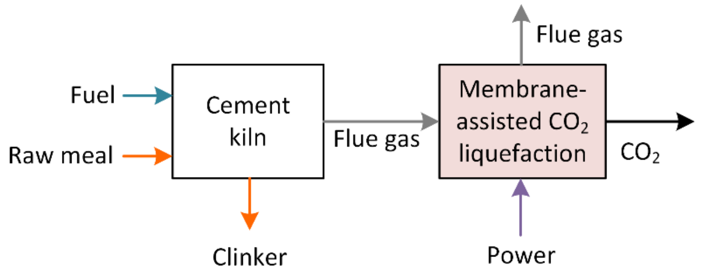

Figure 8.1. Schematic overview of the integration of membrane-assisted CO_2 liquefaction with a cement kiln.

8.1 Process description

The concept of the membrane-assisted CO_2 liquefaction process is illustrated in Figure 8.2, and more detailed process flowsheet are given in Figure 8.3. The flue gas stream from the cement plant is first cooled by internal heat exchange with CO_2-lean flue gases from the membrane separation and afterwards cooled in a DCC where water is also removed. The stream is thereafter compressed, further cooled and condensed water is removed. The flue gases are subsequently admixed with a recycle stream before entering the membrane module. The CO_2-lean membrane retentate stream is heated with the permeate stream and expanded in order to recuperate a portion of the pressure exergy in the stream, and lastly heated with the raw flue gas before exiting through the stack.

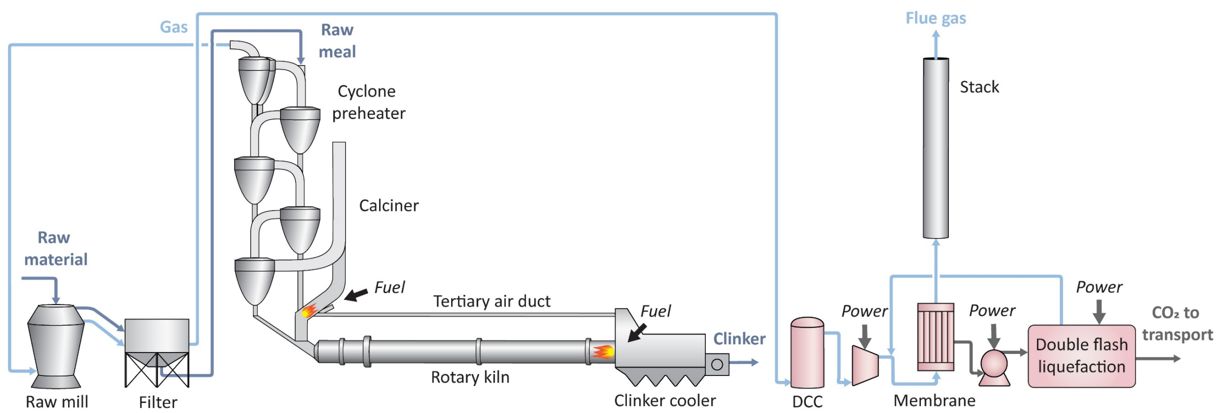

Figure 8.2. Reference cement kiln with membrane-assisted CO_2 liquefaction.

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The membrane permeate vacuum pressure is obtained by vacuum pumping. After vacuum pumping and additional water knockout, the permeate stream enters a three-stage inter-cooled compressor train with water knock-out, followed by a final molecular sieve dehydration unit and admixing with a gaseous CO_2 recycle stream. The stream enters the coldbox and is cooled to -54 °C through process-to-process heat recuperators as well as auxiliary refrigeration exchangers. The partially condensed stream is separated in a vapour–liquid separator (main separator). The gaseous separation product stream is heated against the feed stream and further against compressor intercooling waste heat. The stream is heated and expanded in two stages before recycled to the membrane unit.

To finally prepare the CO_2 for pipeline transport, the liquid CO_2 product is partly heated and throttled to 8.9 bar into a second vapour–liquid separator to maximize CO_2 purity. The liquid product is thereafter pumped to target transport pressure in two stages with intermediate heating against the permeate stream, while the flash gas is recycled.

The sensitivity to flue gas impurities like NO_x and SO_x vary largely among membrane types. The producer of the particular membrane assumed in this process claims that it has a high NO_x and SO_x tolerance. However, to be conservative, the same additional NO_x and SO_x removal as in the MEA process is assumed: the SO_x level is reduced by injection of NaOH solution in the DCC, and the NO_x level is reduced by increasing the NO_x removal in the SNCR system that is installed in the reference kiln.

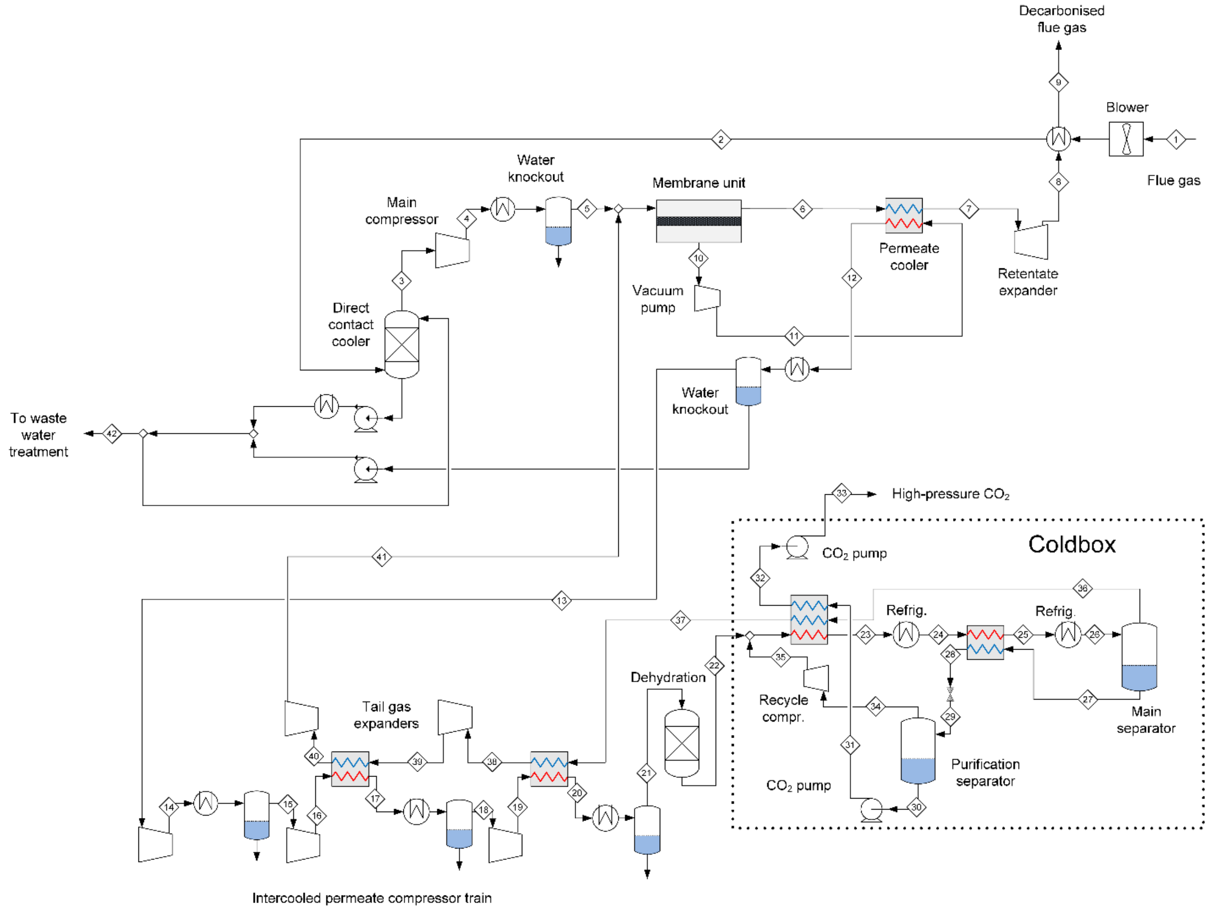

Figure 8.3. The membrane-assisted CO_2 liquefaction capture process designed for pipeline CO_2 transport.

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The process design is slightly different when the CO_2 product should be conditioned for ship transport (Figure 8.4). The second liquid-vapor separation stage takes place at the target pressure, 6.5 bar, and instead of pumping, the product is cooled to 3 °C below the boiling point.

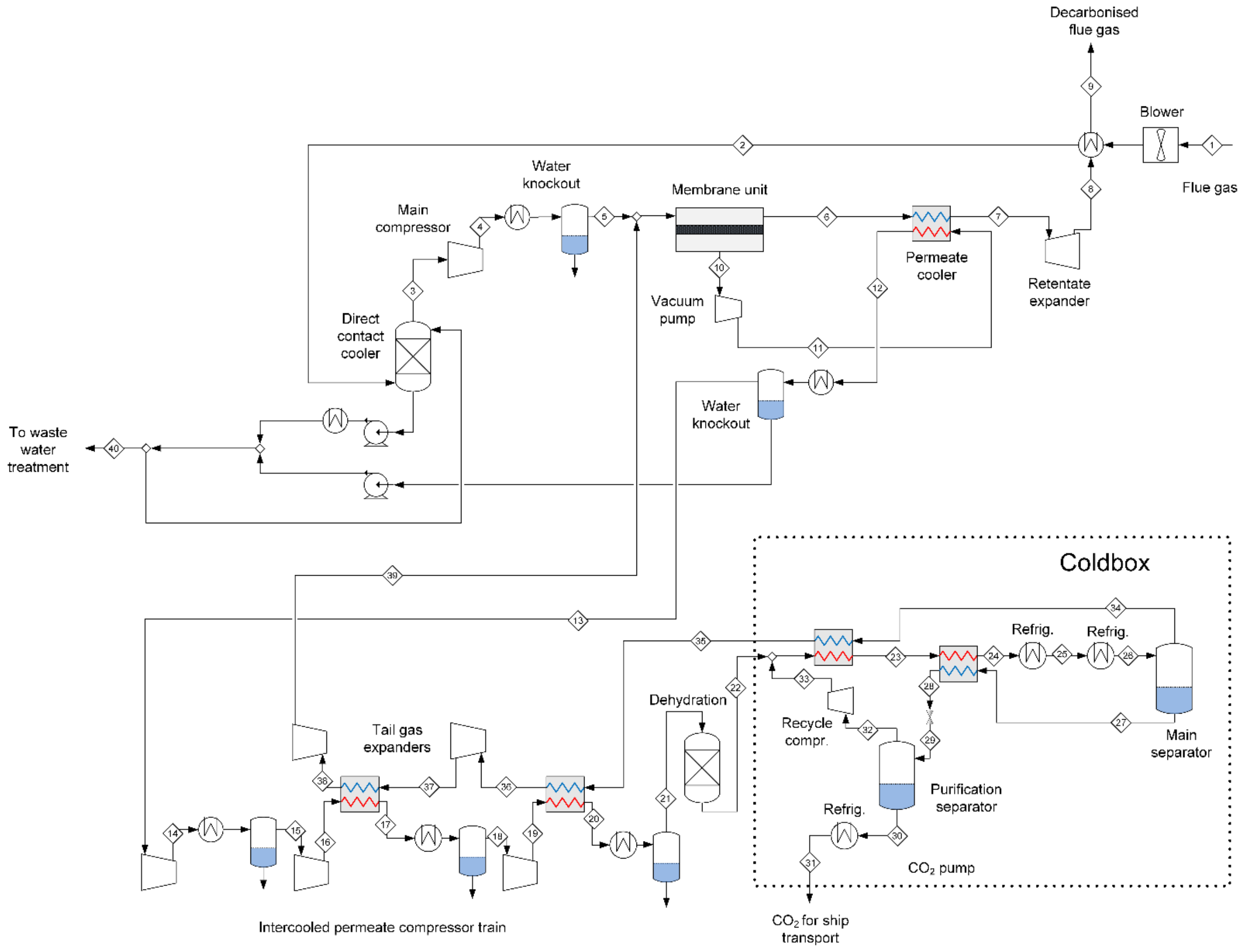

Figure 8.4. The membrane-assisted CO_2 liquefaction capture process design for ship CO_2 transport.

8.2 Process evaluation

The system is simulated in steady-state and equilibrium mode in Aspen HYSYS V9, with the Peng-Robinson equation of state. A multicomponent membrane model developed at SINTEF Energy Research is used to simulate the membrane separation. The membrane model is a 1-D distributed cross flow model that requires the following inputs:

- Feed gas mass flow, pressure and composition
- Permeate gas pressure
- Membrane permeance of all components
- Membrane area

Based on the inputs the model calculates permeate and retentate composition and mass flows. The membrane model has been integrated into the HYSYS interface.

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Membrane permeance data representative for the membrane that has been tested within CEMCAP are used in the model (Table 8.1). It should be noted that the selection of membranes available for testing was limited, and that it is likely that membranes that are more optimal for the process can be found.

Table 8.1. Membrane permeance data.

Component	Permeance [m^3(STP)/m^2sPa]	Permeance [m^3(STP)/m^2bar-h]	CO_2 selectivity over other components
CO_2	$7.5 \cdot 10^{-9}$	2.7	1
N_2	$1.5 \cdot 10^{-10}$	0.054	50
O_2	$6.0 \cdot 10^{-10}$	0.216	12.5
H_2O	$1.5 \cdot 10^{-7}$	54	0.05

Membrane area, membrane feed pressure and main separator pressure can be varied, and optimal values depend on conditions such as flue gas characteristics and targeted CO_2 capture rate, but also on whether the cost of compression (CAPEX and OPEX) and the cost of membrane (CAPEX) is more dominant. For the base case and the low air leak case membrane area of 228,000 m^2 and main separator pressure of around 30 bar are selected. Main separator pressure is adjusted to obtain the targeted CO_2 capture rate. For the case with 60% capture, the membrane area is reduced to 152,000 m^2, and the membrane feed pressure adjusted. In the ship transport case, the process is modified as shown in Figure 8.4. The same membrane as in the base case is assumed, and the membrane feed pressure is adjusted to obtain 90% capture. A more detailed description of the process design and simulation in the different cases evaluated can be found in D11.3 "Membrane-assisted CO_2 liquefaction for CO_2 capture from cement plants" [BER, 2018].

The key results from process simulations of the membrane-assisted CO_2 liquefaction cases are presented in Table 8.2. Stream data are provided in Section 5 in the Supplementary information.

Table 8.2. Key process characteristics and simulation results for membrane-assisted CO_2 liquefaction cases.

Case	Base case		Low air leak in mill	Optional extent of capture		Ship transport	
Design	Medium air leak		Low air leak	Medium air leak		Medium air leak	
Operation	Low air leak	Medium air leak	Low air leak	Low air leak	Medium air leak	Low air leak	Medium air leak
Membrane area [m^2]	228,000	228,000	228,000	152,000	152,000	228,000	228,000
Membrane feed pressure [bar]	2.23	2.50	2.23	1.42	1.58	2.07	2.51
Main separator pressure [bar]	29	32	29	29	32	29	32
CO_2 avoided from flue gas	90%	90%	90%	60%	60%	90%	90%
Captured and conditioned CO_2 mole fraction (wet basis)	99.40%	99.27%	99.40%	99.41%	99.29%	99.84%	99.81%
Specific power consumption [MJ/kg_{CO2}]	1.23	1.44	1.23	1.09	1.26	1.34	1.58
Cooling water demand [MJ/kg_{CO2}]	1.54	1.51	1.54	1.78	1.70	1.79	1.77

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CO_2 capture ratio and CO_2 avoided for the membrane-assisted CO_2 liquefaction cases are shown in Table 8.3. SPECCA is presented in Figure 8.5, and a break-down of the different contributions to the SPECCA values is presented in Table 8.4. The following can be noticed:

- In the case of low air leak, the increased concentration of CO_2 in the treated flue gases results in an 8% reduction in SPECCA compared with the base case. This is primarily due to the reduced power consumption of compressors and pumps, especially the flue gas compressor and the vacuum pump which operate with lower pressure ratios in this case and a lower volume flow rate of gases. As a result, the specific electricity consumption (in MJ/kg CO_2) decreases as well as the indirect CO_2 emissions associated with consumption of electricity from the grid.
- In the optional extent of capture case the decreased capture ratio also results in a decrease in the SPECCA of about 13% compared with the base case. Since a smaller portion of the CO_2 in the flue gases needs to pass the membrane, a significantly lower pressure is required upstream the membrane which reduces the specific electricity consumption.
- The increased electricity consumption in the ship transport case, and consequently the decrease in equivalent specific CO_2 avoided, results in a 11% increase in SPECCA compared with the base case.
  - The CO_2 separation step takes place at a lower pressure than in the base case and as a result, more gaseous CO_2 is compressed and recycled back to the bulk separation step. On the other hand, there is no need for pumping of liquid CO_2 in the ship transport case.
  - Furthermore, the increase in CO_2 flow to the bulk separation step also results in a higher cooling and refrigeration demand.

Table 8.3. CO_2 capture ratio and CO_2 avoided for membrane-assisted CO_2 liquefaction cases.

	Base case	**Low air leak in mill**	**Optional extent of capture**	**Ship transport**
CO_2 capture ratio	90%	90%	60%	90%
CO_2 avoided from flue gas	90%	90%	60%	90%
CO_2 avoided	90%	90%	60%	90%
Equivalent CO_2 avoided	78%	79%	53%	77%

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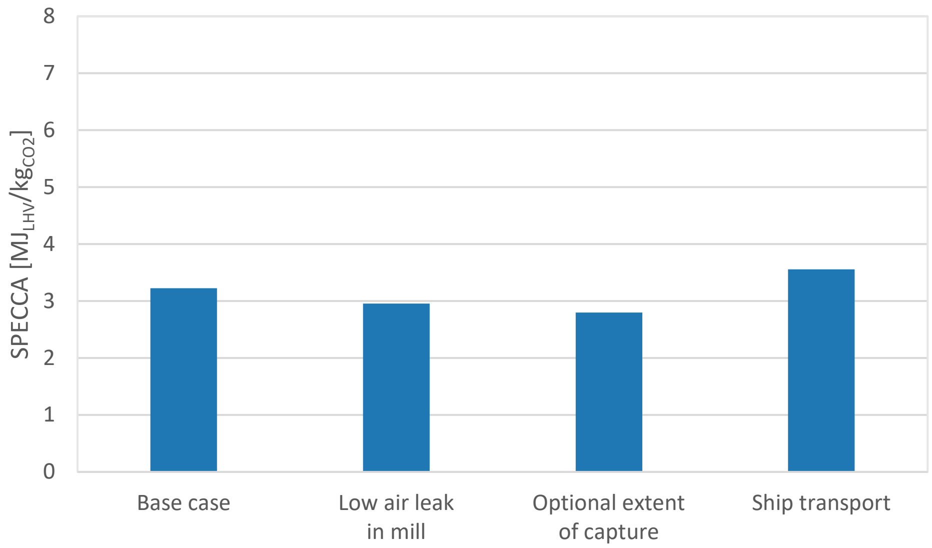

Figure 8.5. SPECCA for membrane-assisted CO_2 liquefaction cases.

Table 8.4. Break-down of SPECCA for membrane-assisted CO_2 liquefaction cases.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**2216**	**2048**	**1299**	**2419**
Electric power consumption				
Pumps, fans, compressors in main process	1905	1763	1098	1961
Refrigeration	260	261	162	399
Cooling water system	51	51	38	59
Equivalent specific CO_2 avoided [kg_{CO2}/t_{clk}]	**687**	**693**	**464**	**680**
At cement kiln stack	761	761	507	761
Electric power consumption				
Pumps, fans, compressors in main process	-63	-58	-37	-65
Refrigeration	-9	-9	-5	-13
Cooling water system	-2	-2	-1	-2
SPECCA [$MJ_{LHV}/kgCO_2$]	**3.22**	**2.96**	**2.80**	**3.56**

8.3 Economic evaluation

The utility consumption, other consumables and the required operating labor of the cement kiln with membrane-assisted CO_2 liquefaction are presented in Table 8.5. Economic results in terms of both capital and operating costs are presented in Table 8.6, together with the resulting cost of clinker and CO_2 avoided cost. Detailed equipment lists with estimated equipment costs and direct costs are provided in Section 5 in Supplementary information.

The process contingency of the MAL technology is set to 52%. Full size modules of the technology have been operated in industry conditions and the contribution to the process contingency based on technological maturity ($C_{process,1}$) is therefore set to 40 %-points. The level of detail of the

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equipment list is relatively high, so the contribution to process contingency based on this ($C_{process,2}$) is set to 12 %-points, resulting in the overall value of 52%.

Table 8.5. Utilities, consumables and operating labor for cement kiln with membrane-assisted CO_2 liquefaction.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport
Clinker production [t/h]	120.65	120.65	120.65	120.65
Coal [t/h]	13.93	13.93	13.93	13.93
Electricity [MW]	49.96	47.39	35.86	53.09
Cooling water make-up [t/h]	85.32	86.34	64.92	99.77
NaOH solution for $DeSO_x$ [t/h]	0.09	0.09	0.09	0.09
Ammonia solution for SNCR [t/h]	0.62	0.62	0.62	0.62
Membrane material replacement [m^2/year]	50160	50160	33440	50160
Operating labor - number of persons	120	120	120	120

Table 8.6. Economic results for cement kiln with membrane-assisted CO_2 liquefaction.

	Base case	Low air leak in mill	Optional extent of capture	Ship transport
Variable OPEX [€/t_{clk}]	**41.2**	**40.0**	**34.2**	**42.7**
Raw meal	5.0	5.0	5.0	5.0
Coal	9.4	9.4	9.4	9.4
Electricity	24.1	22.8	17.3	25.6
Cooling water make-up	0.3	0.3	0.2	0.3
NaOH solution for $DeSO_x$	0.3	0.3	0.3	0.3
Ammonia for SNCR	0.7	0.7	0.7	0.7
Membrane material replacement	0.4	0.4	0.3	0.4
Other variable costs	1.1	1.1	1.1	1.1
Fixed OPEX [€/t_{clk}]	**32.1**	**28.44**	**30.8**	**32.3**
Operative, administrative and support labor	11.1	10.9	11.1	11.1
Insurance and local taxes	9.3	7.8	8.7	9.4
Maintenance cost (including maintenance labor)	11.7	9.8	10.9	11.8
Total plant costs, CO_2 capture [M€]	**246.7**	**214.0**	**173.1**	**250.5**
TDC without process contingencies	127.6	111.0	88.8	130.1
Process contingencies	53.8	46.4	38.5	54.1
Indirect costs	25.4	22.0	17.8	25.8
Owner's costs	12.7	11.0	8.9	12.9
Project contingencies	27.2	23.6	19.1	27.6
Total plant costs, cement kiln + CO_2 capture [M€]	**450.4**	**417.8**	**376.9**	**454.2**
Total OPEX [€/t_{clk}]	**73.3**	**64.9**	**68.4**	**75.0**
Total CAPEX [€/t_{clk}]	**46.7**	**43.6**	**38.9**	**47.1**
Cost of clinker [€/t_{clk}]	**120.0**	**114.3**	**101.5**	**122.1**
Cost of cement [€/t_{cement}]	**88.2**	**84.1**	**74.7**	**89.8**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**83.5**	**74.7**	**84.0**	**87.6**

Break-downs of the cost of CO_2 avoided for all membrane-assisted liquefaction cases are presented in Figure 8.6, together with the cost of clinker in each case.

In the case of low air leak, the process benefits from the increased CO_2 concentration in the flue gas compared with the base case. The higher CO_2 concentration and lower flue gas flow rate results in reduced design capacity for much of the process equipment, which consequently results in lower

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CAPEX and lower fixed operating costs, despite the reduced plant capacity factor due to increased maintenance. Furthermore, the net demand of electricity decreases compared with the base case, mainly due to the reduced duty of the flue gas compressor which both processes a smaller gas flow and operates with a slightly lowered outlet pressure.

It is evident that the technology performs better in terms of energy when designed for a reduced CO_2 capture ratio. In terms of economics, the cost of clinker decreases significantly, as expected, but a slight increase in the cost of CO_2 avoided is also observed. The decreased CO_2 capture ratio results in a lower membrane area and lower pressure difference required over the membrane to achieve the desired capture ratio, i.e. lower degree of flue gas pressurisation. The reduced flue gas pressurisation results in a significantly reduced electricity requirement and resulting operating costs. However, the absolute decrease in capital costs is smaller relative to the decrease in avoided equivalent specific CO_2 emissions, resulting in higher CAPEX and fixed operating costs compared to the base case.

The ship transport case results in higher cost of clinker and CO_2 avoided than the base case. This is mainly a result of the increased electricity consumption for CO_2 liquefaction.

As shown in Figure 8.6, capital cost is the single largest cost factor in all the investigated cases. It is therefore worth mentioning that the flue gas compressor upstream of the membrane is a very expensive piece of equipment, as well as the membrane itself and the vacuum pump downstream the membrane. The costs of these components are greatly affected by different process parameters, e.g. the flue gas compressor equipment cost is largely influenced by the flow rate (a 50% reduction in flow from the base case results in nearly 60% reduction in equipment cost). However, the compressor pressure ratio also significantly affects the cost (a 50% reduction in the pressure increase over the compressor decreases the equipment cost by around 20%). Thus, there is room for process improvement with respect to the pressure required upstream the membrane. This could reduce not only capital costs (and consequently fixed operating costs which are estimated from the capital costs), but also variable costs as the compressor requires a large amount of electricity to operate. It can also be pointed out that the energy performance of the process and the capital costs of the flue gas compressor and the vacuum pump (due to pressure ratio), are highly dependent on membrane performance. In CEMCAP, the MAL process design was restricted to the specific membrane type that was tested within the project. It is likely that a different choice of membrane for the application, with different characteristics such as permeance, could result in better process performance and lower costs than observed here.

It could also be noted that due to the low maturity of the tested membrane, the associated process contingencies are relatively high compared to the other CEMCAP technologies and contribute considerably to the CAPEX and consequently the fixed operating costs. The process contingency of the whole membrane related section of the process is set based on the maturity of the membrane. This section (the upper part of the process as depicted in Figure 8.3), accounts for roughly 60% of the TDC of the whole CO_2 capture process, and its associated maturity related process contingencies account for around 16% of the TDC.

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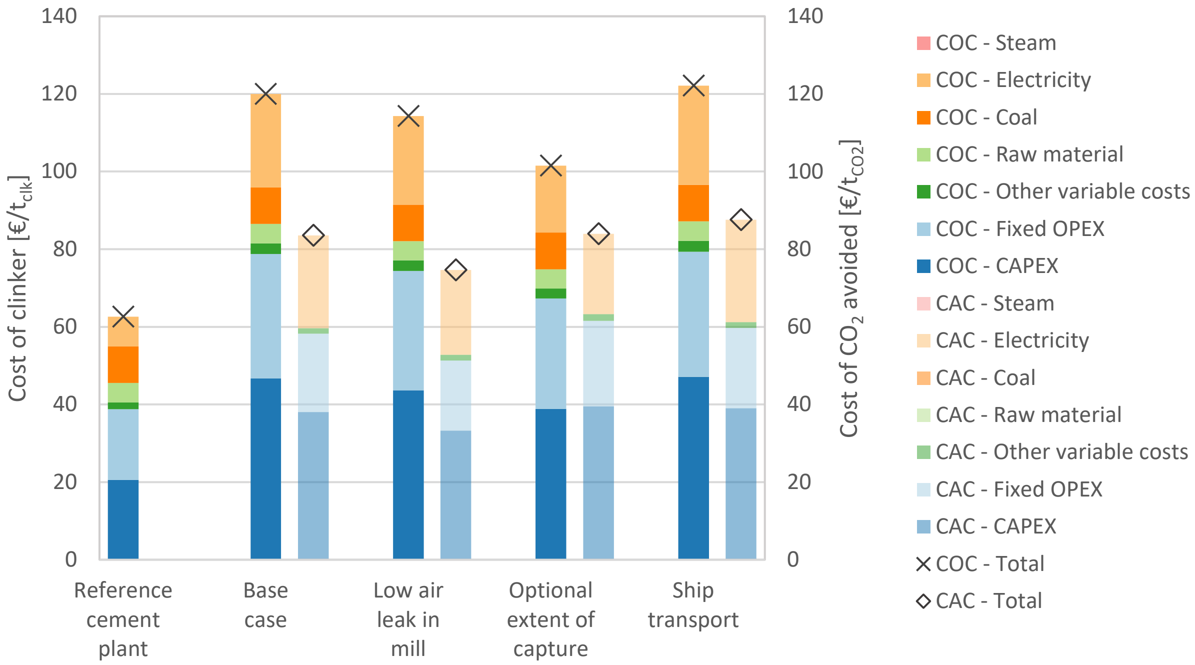

Figure 8.6. Breakdown of the cost of clinker and cost of CO_2 avoided for all membrane-assisted CO_2 liquefaction cases.

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9 CALCIUM LOOPING – TAIL-END CONFIGURATION

The Calcium-Looping (CaL) technology is based on the reversible carbonation reaction ($CaCO_3 \rightleftharpoons CaO+CO_2$), which is exploited to separate carbon dioxide from flue gas. The technology can be applied to a cement plant as a tail-end technology (Figure 9.1) or it can be integrated with the calcination process taking place in the cement kiln (see Section 10). In the tail-end configuration the flue gas from the cement kiln is sent to the CaL system for purification, and CaO-rich purge from the CaL system is sent to the cement kiln and added to the raw meal. The CaL process requires supply of limestone and coal. Oxygen is required in the process, and this is generated by an ASU. Power is required both for the ASU, the core CaL process and for CO_2 purification, but power is also generated by a steam cycle utilizing waste heat in the process.

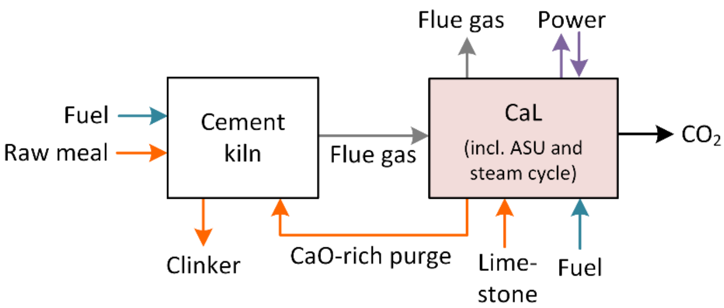

Figure 9.1. Integration of tail-end calcium looping into the cement plant.

9.1 Process description

The concept of the CaL tail-end technology is illustrated in Figure 9.2. The CaL process is carried out in two interconnected reactors: the carbonator and the calciner. In the carbonator the CaO-based CO_2-sorbent reacts with CO_2 from the flue gas and forms $CaCO_3$ which is transferred to the calciner. In the calciner oxy-combustion of coal is carried out to provide heat for the reverse calcination reaction and regenerate the sorbent while producing a CO_2-rich product stream. The carbonator can in principle be placed before or after the raw mill. When the carbonator is placed before the raw mill, air ingress in the raw mill does not dilute the gas to be treated in the carbonator, resulting in a lower flue gas flow rate with a higher CO_2 concentration compared to a post-mill configuration.

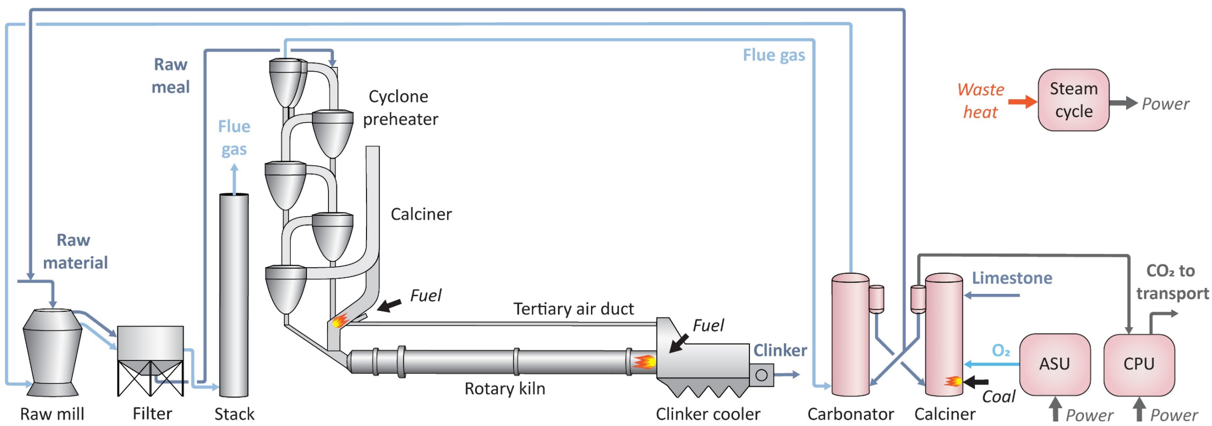

Figure 9.2. Reference cement kiln with CO_2 capture by the CaL tail-end technology.

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The CaL reactors in the tail-end configuration are circulating fluidized beds (CFB). CaO-rich solid purge from the CaL process, which is rich of CaO, is milled and fed to the clinker burning process, replacing part of the limestone in the raw meal. Milling of the CaL purge in the raw mill is needed because CFB reactors in conditions relevant for the CaL process must operate with particle sizes in the range of 100 to 300 μm to obtain a good fluidization. This particle size is too high compared to the average particle size of about 10-20 μm needed in clinker burning. The Integration Level (IL) is defined as the percentage of Ca entering the clinker burning process with the CaL purge as CaO, with respect to the total Ca fed to the cement kiln.

The high operating temperatures of the process allows exploiting the heat introduced with the fuel in the calciner for sorbent regeneration as high temperature heat, which can be used for steam production and power generation in a Rankine cycle.

The CO_2 released in the calciner has a purity of approximately 92% on dry basis because of the excess oxygen in the oxyfuel combustion calciner, of the impurities in the O_2 produced in the cryogenic ASU and of the false air ingress. Therefore, downstream CO_2 purification must be performed in a CPU.

A detailed flowsheet of the tail-end CaL system considered in this report is shown in Figure 9.3. The flowsheet of the CPU is presented in Figure 9.4 for cases with pipeline transport, and in Figure 9.5 for the case with ship transport (applies to both the tail-end and the EF integrated configurations).

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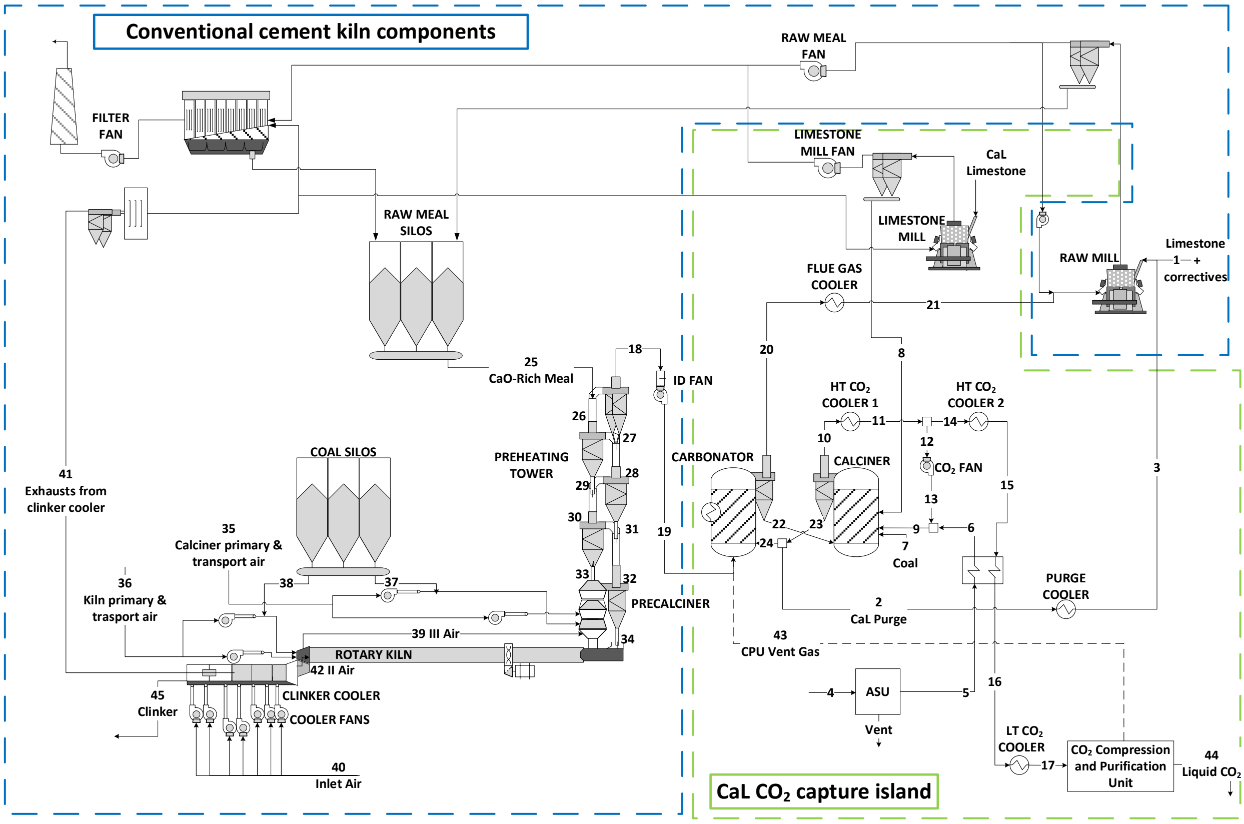

Figure 9.3. Process flowsheet of the tail-end CaL process integration for CO2 capture in a cement kiln.

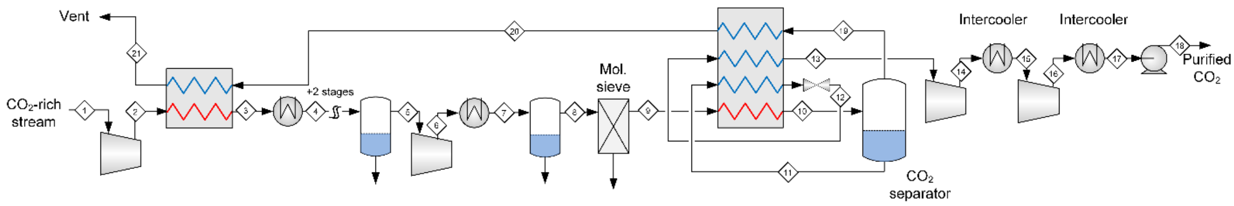

Figure 9.4. Process flowsheet of CPU for pipeline transport in CaL-Tail-End and Integrated EF applications.

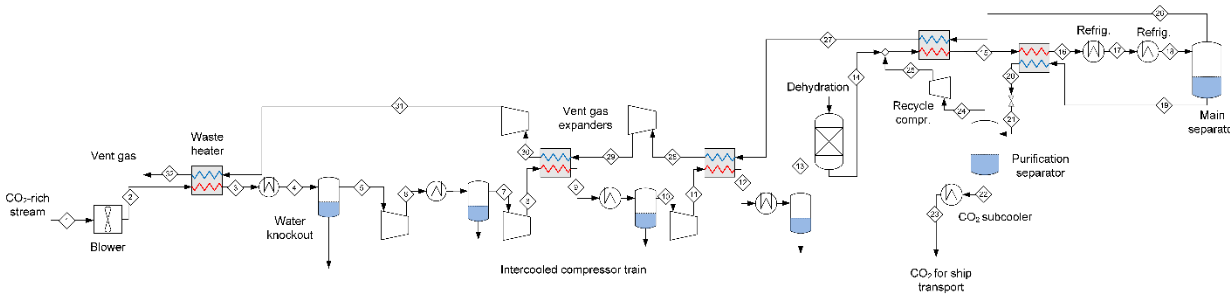

Figure 9.5. Process flowsheet of CPU for ship transport in CaL-Tail-End and Integrated EF applications.

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9.2 Process evaluation

Polimi's in-house process simulation code GS [GS, 2016] has been used to reproduce the VDZ simulation of the reference kiln [CAM, 2016], and to simulate the integration of the capture processes into this kiln. For consistency, it is the PoliMi simulation of the reference kiln that has been used as reference for the evaluation of the calcium looping processes. A detailed description of the calculation methodology and assumptions of the Tail-End CaL cement plant is given by De Lena *et al*. [DEL, 2017].

The CFB reactor model presented by Romano *et al.* [ROM, 2012] has been used to simulate the carbonator, which includes the carbonation kinetic expression proposed by Grasa *et al.* [GRA, 2008]. This reactor model allows calculating CaO conversion in the carbonator as a function of the main operating conditions in this reactor (temperature, solid inventory, solid make-up and solid circulation), taking the effects of coal ashes and sulfur species into account. The calciner is modelled assuming complete calcination of the sorbent. Based on the data in [MAR, 2013], this assumption is justified considering the high residence time of the solids in a CFB calciner and the assumed calcination temperature of 920°C. Further details about simulations of the CaL reactors are described by Spinelli *et al.* [SPI, 2018].

The heat recovery steam cycle parameters are defined according to the thermal power available, as discussed in the CEMCAP Framework [VOL, 2018]. The CPU is simulated using Aspen HYSYS V9, as a single stage flash self-refrigerated unit (Figure 9.4), slightly modified compared to the CPU of the oxyfuel process. The power consumption of the CPU is around 0.4 MJ/kg_{CO2}. The heat required for dehydration in the CPU is assumed to be 16.6 MJ_{th}/kg_{H2O} (estimated based on [KEM, 2014]). The power consumption of the ASU is 226 kWh/t_{O2}, and the dehydration heat demand is 58.3 kJ/kg_{O2}. The heat required for dehydration in these two units are provided by some of the steam generated in the process. Key process characteristics of the calcium looping cases are summarized in Table 9.1. Stream data are provided in Section 6 in the Supplementary information.

Table 9.1. Key process characteristics and simulation results for CaL tail-end cases.

	Base case		IL 20%		Ship transport (IL 50%)	
IL, %	50		20		50	
F_0/F_{CO2}	0.60		0.16		0.60	
F_{Ca}/F_{CO2}	2.94		4.78		2.94	
Clinker production, t/h	117.7		118.0		117.7	
Direct fuel consumption (q), MW_{LHV} – MJ_{LHV}/kg_{clk}	232.2	7.10	285.9	8.72	232.2	7.10
Fuel consumption in the rotary kiln, MW_{LHV} – MJ_{LHV}/kg_{clk}	39.9	1.22	39.9	1.22	39.9	1.22
Fuel consumption in the pre-calciner, MW_{LHV} – MJ_{LHV}/kg_{clk}	27.6	0.85	50.8	1.55	27.6	0.85
Fuel consumption in the CaL calciner, MW_{LHV} - MJ_{LHV}/kg_{clk}	164.6	5.04	195.2	5.95	164.6	5.04
Oxygen Input, t/day – kg_{O2}/kg_{clk}	1236.2	0.44	1453.9	0.51	1236.2	0.44
CO_2 capture efficiency of the carbonator, %	90.0		88.8		90.0	
CO_2 generated from fuel combustion, kg/s – kg_{CO2}/t_{clk}	21.6	661.6	26.6	812.6	21.6	661.6
CO_2 generated from raw meal calcination, kg/s – kg_{CO2}/t_{clk}	18.7	572.9	18.4	562.4	18.7	572.9
Captured CO_2 to storage, kg/s – kg_{CO2}/t_{clk}	37.8	1155.5	41.2	1256.6	37.5	1147.3
Total CO_2 capture efficiency, %	93.6		91.4		93.1	
Direct CO_2 emission from kiln flue gas, kg/s – kg_{CO2}/t_{clk}	1.6	48.0	2.6	79.5	1.6	48.0

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Direct CO_2 emission from the CPU, kg/s – kg_{CO2}/t_{clk}	1.0	31.1	1.3	39.0	1.3	40.4
Total direct CO_2 emission (e_{CO2}), kg/s – kg_{CO2}/t_{clk}	2.6	79.1	3.9	118.5	2.9	87.6
Direct emission reduction, %	90.9		86.3		89.9	
Power balance						
Gross steam turbine electricity production, MW_e –kWh_e/t_{clk}	41.58	353.2	66.16	560.6	41.58	353.2
Steam cycle pumps, MW_e – kWh_e/t_{clk}	-0.80	-6.8	-1.28	-10.9	-0.80	-6.8
Auxiliaries for steam cycle heat rejection, MW_e –kWh_e/t_{clk}	-0.78	-6.6	-1.11	-9.4	-0.78	-6.6
ASU consumption, MW_e – kWh_e/t_{clk}	-11.64	-98.9	-13.69	-116.0	-11.64	-98.9
Fans in CaL system, MW_e – kWh_e/t_{clk}	-2.70	-22.9	-2.97	-25.2	-2.70	-22.9
CO_2 compression, MW_e – kWh_e/t_{clk}	-16.11	-136.9	-17.60	-149.0	-29.76	-252.8
Power consumption for cooling system, MW_e – kWh_e/t_{clk}	-0.67	-5.7	-0.74	-6.3	-1.08	-9.2
Auxiliaries for calciner fuel grinding, MW_e – kWh_e/t_{clk}	-0.18	-1.5	-0.22	-1.8	-0.18	-1.5
Cement plant auxiliaries, MW_e – $kWhe/t_{clk}$	-15.49	-131.6	-15.53	-131.6	-15.49	-131.6
Net electricity production, MW_e – kWh_e/t_{clk}	-6.79	-57.7	13.01	110.3	-20.85	-177.1
Net electricity production, kWh_e/t_{cem}	-42.5		81.3		-130.5	

CO_2 capture ratio and CO_2 avoided for the calcium looping tail-end cases are shown in Table 9.2. SPECCA is presented in Figure 9.6, and a break-down of the different contributions to the SPECCA values is presented in Table 9.3. The following can be noticed:

- The largest contributor to SPECCA is the increase in coal consumption associated with the CaL system.
- The ASU and CPU require a large amount of electricity. On the other hand, a significant amount of electricity is produced in the waste heat recovery steam cycle in all cases
- In the base case and the IL 20% case, the CaL system produces enough electricity to cover the CaL system´s own demand as well as a part of the cement kiln´s demand. In fact, the CaL system in the IL 20% case is a net producer and exporter of electricity (*cf.* Table 9.1).
    - Although the IL 20% case is a net producer of electricity, the SPECCA in this case is about 9% higher than in the base case, mainly due to a significantly larger coal consumption.
- The ship transport case results in an increase in SPECCA of around 35% higher compared with the base case. This is primarily due to the CPU for ship transport being more energy intensive than the CPU for pipeline transport.

Table 9.2. CO_2 capture and CO_2 avoided for CaL tail-end cases.

	Base case / IL 50%	**IL 20%**	**Ship transport**
CO_2 capture ratio	94%	91%	93%
CO_2 avoided from flue gas	91%	86%	90%
CO_2 avoided	91%	86%	90%
Equivalent CO_2 avoided	90%	90%	85%

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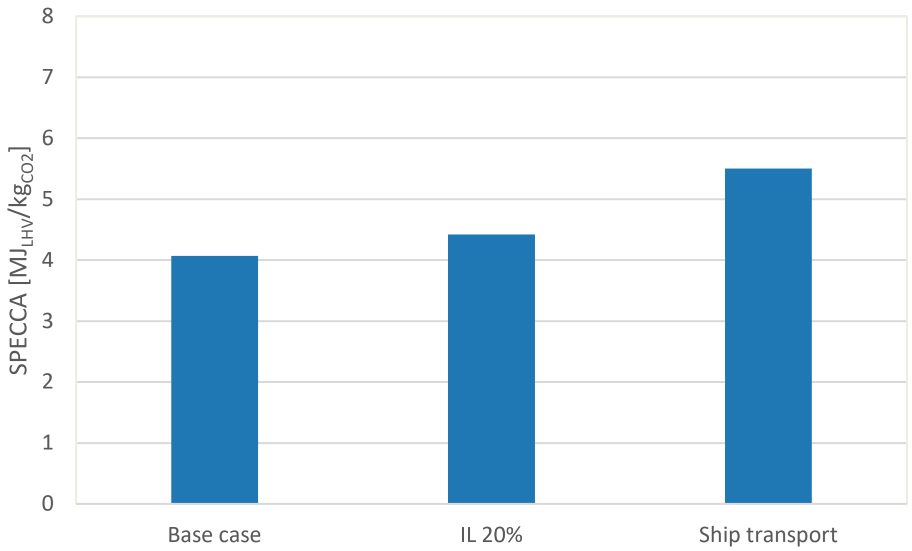

Figure 9.6. SPECCA for CaL tail-end cases.

Table 9.3. Break down of SPECCA for CaL tail-end cases.

	Base case / IL 50%	IL 20%	Ship transport
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**3280**	**3583**	**4217**
Coal consumption	3859	5480	3859
Electric power consumption			
Fans and fuel grinding	192	212	192
ASU	776	910	776
CPU	1074	1169	1983
Cooling water system	44	49	72
Electric power generation	-2665	-4238	-2665
Equivalent specific CO2 avoided [kg_{CO2}/t_{clk}]	**806**	**811**	**767**
At cement kiln stack	787	747	779
Electric power consumption			
Fans and fuel grinding	-6	-7	-6
ASU	-26	-30	-26
CPU	-36	-39	-66
Cooling water system	-1	-2	-2
Electric power generation	89	142	89
SPECCA [MJ_{LHV}/kgCO2]	**4.07**	**4.42**	**5.50**

9.3 Economic evaluation

The clinker production and the main utilities and consumables are reported in Table 9.4. The results of the cost evaluation are presented in Table 9.5. Detailed equipment lists with estimated equipment costs and direct costs are provided in Section 6 in Supplementary information.

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With the exception of the ASU and the steam cycle, whose process contingencies are indicated in Table 2.7, the process contingency of the CaL – Tail-End configuration is set to 32%. Full size modules of the complete system are ready to be operated in industry conditions and the contribution to the process contingency based on technological maturity ($C_{process,1}$) is therefore set to 20 %-points. The level of detail of the equipment list is moderate, so the contribution to process contingency based on this ($C_{process,2}$) is set to 12 %-points.

Table 9.4. Utilities, consumables and operating labor for cement kiln with CaL CO_2 capture.

	Base case	IL 20%	Ship transport
Clinker production [t/h]	117.7	118.0	117.7
CO_2 avoided from flue gas [%]	90.9	86.3	89.9
CO_2 captured [t/h]	136.0	148.3	135.1
Coal [t/h]	30.8	37.9	30.8
Electricity consumption [MW_e]	6.79	-13.01	20.85
Ammonia solution for SNCR [t/h]	0.59	0.59	0.59
Cooling water [MW_{th}]	116.7	158.0	160.4
Operating labor - number of persons	120	120	120

Table 9.5. Economic results for cement kiln with CaL CO_2 capture.

	Base case	IL 20%	Ship transport
Variable OPEX [€/t_{clk}]	**32.2**	**27.5**	**39.3**
Raw meal	4.9	4.9	4.9
Coal	21.3	26.1	21.3
Electricity	3.4	-6.4	10.3
Ammonia for DeNOx	0.7	0.7	0.7
Cooling water	0.9	1.2	1.0
Other variable costs	1.1	1.1	1.1
Fixed OPEX [€/t_{clk}]	**30.6**	**32.0**	**33.3**
Operative, administrative and support labor	11.2	11.3	11.4
Insurance and local taxes	8.6	9.2	9.7
Maintenance cost (including maintenance labor)	10.8	11.5	12.2
Total plant costs, CO_2 capture [M€]	**202.3**	**231.0**	**253.7**
TDC without process contingencies	124.8	142.2	153.5
Process contingencies	23.9	27.6	33.1
Indirect costs	20.8	23.8	26.1
Owner's costs	10.4	11.9	13.1
Project contingencies	22.3	25.5	28.0
Total plant costs, cement kiln + CO_2 capture [M€]	**406.0**	**434.7**	**457.5**
Total OPEX [€/t_{clk}]	**62.8**	**73.8**	**75.5**
Total CAPEX [€/t_{clk}]	**43.0**	**46.2**	**48.6**
Cost of clinker [€/t_{clk}]	**105.8**	**105.8**	**121.1**
Cost of cement [€/t_{cement}]	**77.8**	**77.8**	**89.1**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**52.4**	**52.1**	**75.1**

A break-down of the cost of CO_2 avoided for all calcium looping – Tail-End configuration cases is presented in Figure 9.7, together with the cost of clinker in each case.

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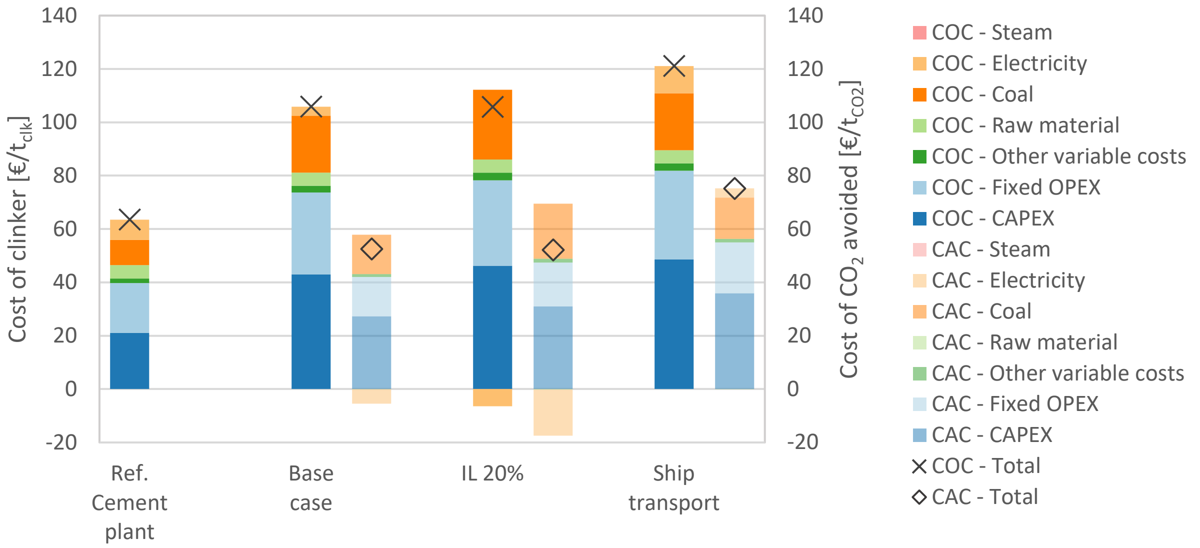

Figure 9.7. Breakdown of the cost of clinker and cost of CO_2 avoided for all CaL tail-end cases.

For both the Tail-End CaL configurations considered, a total cost of cement of about 78 €/t_{cement} has been estimated, which is about 70% higher than the reference plant cost.

The case with IL 20% is the one with the lowest variable OPEX, thanks to the net power production of the overall system. Revenues from the sale of electricity to the grid offset the higher costs of the system, mainly linked to the higher costs of the CaL reactors, the ASU and the steam cycle. An important part of the operating costs is related to fuel consumption. The increase in fuel consumption causes an increase in variable OPEX of about 16% and 35%, with respect to the case without capture system, for the IL 20% and IL 50% respectively. The high fuel consumption is partly compensated by the favorable contribution of electricity thanks to the power export (IL 20%) or lower power import (IL 50%) with respect to the reference cement kiln without capture system. The CAPEX result in a similar range for both the configurations considered, as well as the fixed OPEX. The increase in CAPEX with respect to the referent cement kiln, is mainly due to the presence of ASU and CPU, which represent about 60% (IL 20%) or 65% (IL 50%) of TPC of the CO_2-capture system. The CaL reactors represent the 24% (IL 50%) or 30% (IL 20%) of the TPC of the CO_2-capture system. A cost of CO_2 avoided of about 52 €/t_{CO2} has been estimated for both cases and the CAPEX are responsible for more than 50%.

The ship transport case results in the largest cost of clinker. The consequent increase in cost of CO_2 avoided is due to higher capital costs and electricity consumption related to the CPU.

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10 CALCIUM LOOPING – INTEGRATED ENTRAINED FLOW CONFIGURATION

The Integrated Entrained Flow (EF) CaL technology is a version of the CaL where the CO_2 capture calciner is combined with the calciner in the cement kiln, and thus an integrated part of the cement kiln (Figure 10.1). As a result, the carbonation and calcination must take place in EF reactors. As for the tail-end configuration, additional fuel is required for operation of the calciner, and oxygen is required and supplied with an ASU. Power is required both for the ASU, the core CaL process and for CO_2 purification, but power is also generated by a steam cycle utilizing waste heat in the process.

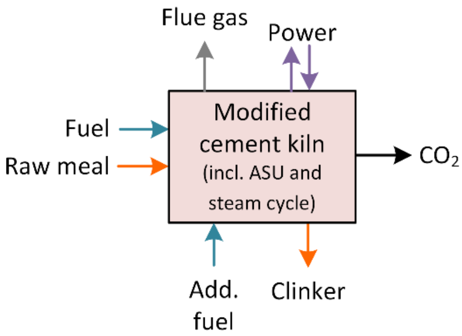

Figure 10.1. Integration of the Integrated EF CaL process to a cement kiln.

10.1 Process description

A simplified illustration of the concept of the Integrated EF CaL technology is given in Figure 10.2. The cement kiln calciner is operated under oxyfuel conditions and is combined with the calciner of the CaL process. This configuration allows maintaining air-blown combustion in the rotary kiln and capturing CO_2 generated by fuel combustion and residual calcination through the CaL, as shown in Figure 10.3.

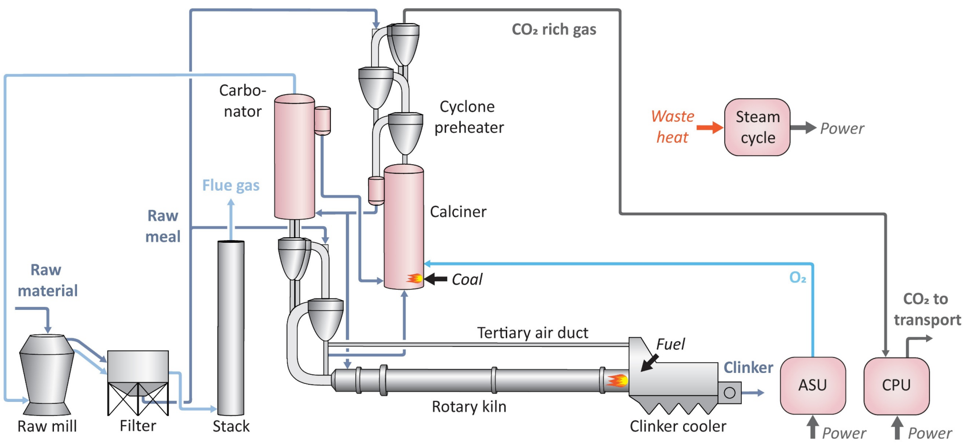

Figure 10.2. Reference cement kiln with CO_2 capture by the Integrated EF CaL

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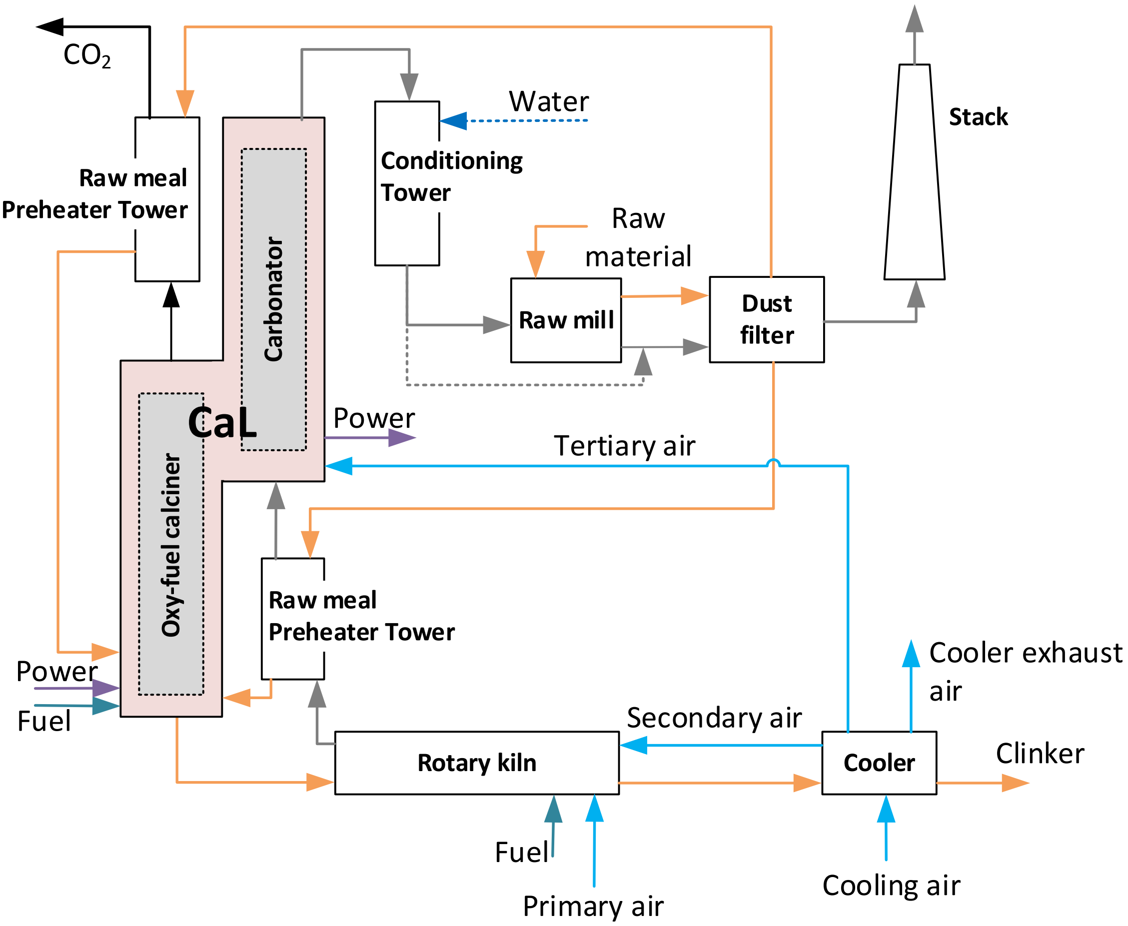

Figure 10.3. Integration of integrated EF CaL CO_2 capture into cement plant.

A portion of the calcined raw meal from the oxyfuel calciner is used as CO_2-sorbent in the carbonator, where CO_2 from the rotary kiln is separated. The sorbent purge of the CaL process represents the entire amount of material needed in the clinker burning process. In this configuration, the CaL process operates with the same material used for clinker production, with a particle size of about 10-20 µm. With such particle size, entrained flow reactors appear preferable over fluidized bed. Therefore, in the highly integrated configuration, the carbonator and calciner are two interconnected entrained flow reactors.

As in the CaL tail-end process, the high operating temperatures of the process allows exploiting the energy introduced with the fuel in the calciner for sorbent regeneration as high temperature heat which can be used for steam production and power generation in a Rankine cycle. Further, the CO_2 released in the calciner has a purity of approximately 92% on dry basis, and downstream CO_2 purification must be performed in a CPU.

A detailed flowsheet of the system is shown in Figure 10.4. For flowsheet of the CPUs for pipeline and ship transport, see Figure 9.4 and Figure 9.5.

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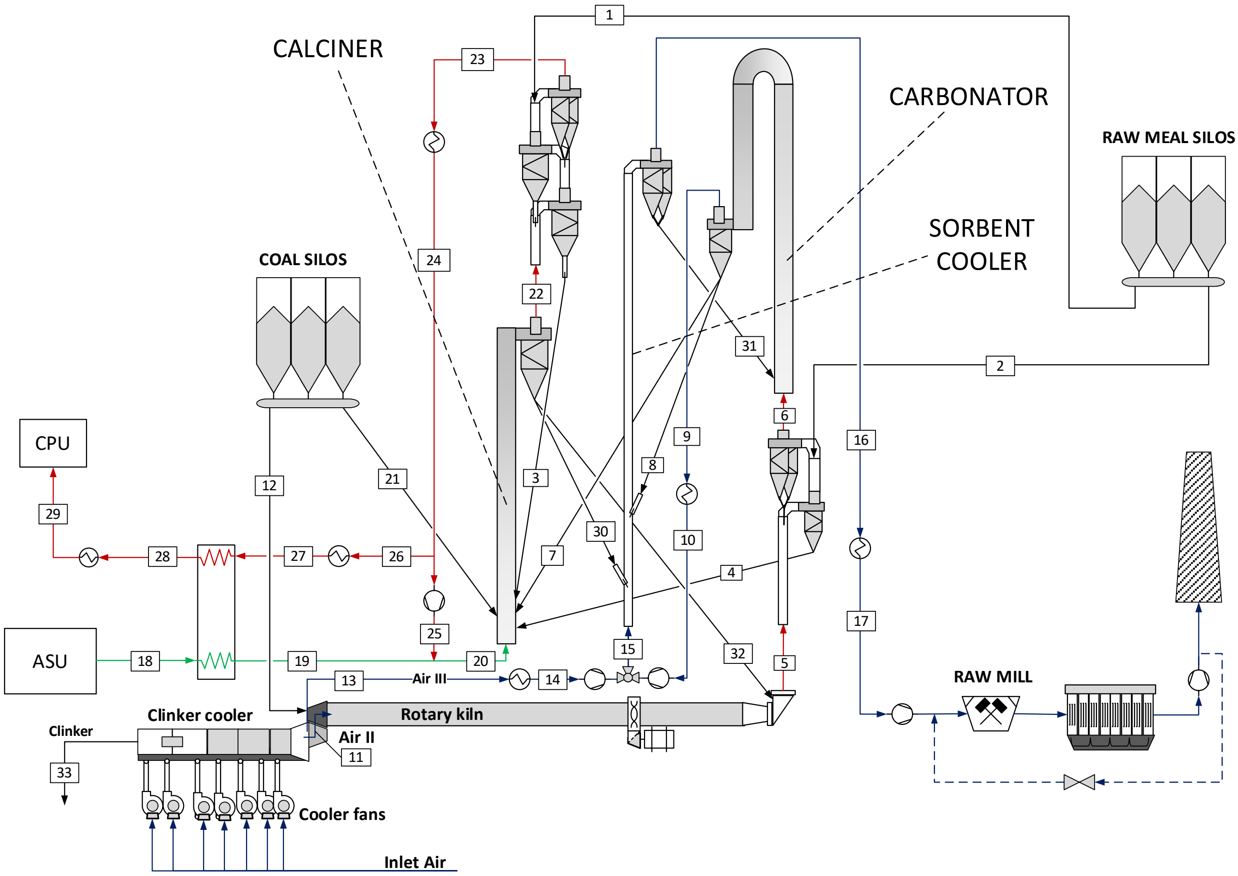

Figure 10.4: Process flowsheet of the integrated EF CaL process with entrained flow reactors

10.2 Process evaluation

As in the tail-end CaL technology, the process evaluation is based on simulations performed with Polimi's in-house process simulation code GS [GS, 2016], and the PoliMi simulation of the reference kiln has been used as reference for evaluation of the process.

A one-dimensional, steady-state model, which is described by Spinelli et al. [SPI, 2018], has been used for the calculation of the entrained flow carbonator. The model solves mass, energy and momentum balances along the axial direction for the gas and solid phases, providing cross-sectional averaged values of chemical composition, temperature and velocity. Sorbent conversion kinetics is described by the random pore model proposed by Grasa et al. [GRA, 2009]. For the calciner, an outlet temperature of 920°C has been assumed to calculate the heat input needed in that reactor to heat up and calcine the recarbonated raw meal from the carbonator.

For the heat recovery steam cycle, the CPU and the ASU the same simulation approaches and assumptions as for the CaL tail-end process are used.

Key process characteristics of the calcium looping cases are summarized in Table 10.1. Stream data are provided in Section 7 in the Supplementary information.

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Table 10.1. Key process characteristics and simulation results for CaL integrated EF cases.

	Base case		Ship transport	
F_0/F_{CO2}	3.96		3.96	
F_{Ca}/F_{CO2}	3.90		3.90	
Clinker production, t/h	117.4		117.4	
Direct fuel consumption (q), MW_{LHV} – MJ_{LHV}/kg_{clk}	177.2	5.44	177.2	5.44
Fuel consumption in the rotary kiln, MW_{LHV} – MJ_{LHV}/kg_{clk}	37.5	1.15	37.5	1.15
Fuel consumption in the CaL calciner, MW_{LHV} - MJ_{LHV}/kg_{clk}	139.8	4.29	139.8	4.29
Oxygen Input, t/day – kg_{O2}/kg_{clk}	1057.6	32.44	1057.6	32.44
CO_2 capture efficiency of the carbonator, %	82.0		82.0	
CO_2 generated from fuel combustion, kg/s – kg_{CO2}/t_{clk}	16.5	506.6	16.5	506.6
CO_2 generated from raw meal calcination, kg/s – kg_{CO2}/t_{clk}	17.9	547.5	17.9	547.5
Captured CO_2 to storage, kg/s – kg_{CO2}/t_{clk}	32.5	997.1	32.4	1011.0
Total CO_2 capture efficiency, %	94.6		94.3	
Direct CO_2 emission from kiln flue gas, kg/s – kg_{CO2}/t_{clk}	0.8	25.7	0.8	25.7
Direct CO_2 emission from the CPU, kg/s – kg_{CO2}/t_{clk}	1.0	31.3	1.2	34.9
Total direct CO_2 emission (e_{CO2}), kg/s – kg_{CO2}/t_{clk}	1.9	57.0	2.0	60.6
Direct emission reduction, %	93.4		93.0	
Power balance				
Gross steam turbine electricity production, MW_e –kWh_e/t_{clk}	21.98	187.2	21.98	187.2
Steam cycle pumps, MW_e – kWh_e/t_{clk}	-0.45	-3.8	-0.45	-3.8
Auxiliaries for steam cycle heat rejection, MW_e –kWh_e/t_{clk}	-0.47	-4.0	-0.47	-4.0
ASU consumption, MW_e – kWh_e/t_{clk}	-9.96	-84.9	-9.96	-84.9
Fans in CaL system, MW_e – kWh_e/t_{clk}	-1.21	-10.3	-1.21	-10.3
CO_2 compression, MW_e – kWh_e/t_{clk}	-14.11	-120.2	-27.27	-232.1
Power consumption for cooling system, MW_e – kWh_e/t_{clk}	-0.58	-5.0	-0.97	-8.2
Auxiliaries for calciner fuel grinding, MW_e – kWh_e/t_{clk}	-0.26	-2.2	-0.26	-2.2
Cement plant auxiliaries, MW_e – $kWhe/t_{clk}$	-15.36	-130.9	-15.36	-130.9
Net electricity production, MW_e – kWh_e/t_{clk}	-20.41	-174.1	-33.95	-289.2
Net electricity production, kWh_e/t_{cem}	-128.2		-213.1	

CO_2 capture ratio and CO_2 avoided for the calcium looping tail-end cases are shown in Table 10.2. SPECCA is presented in Figure 10.5, and a break-down of the different contributions to the SPECCA values is presented in Table 10.3. As for the CaL tail-end system, the base case and the ship transport case differ primarily due to the increased electricity requirement of the CPU for ship transport, which results in a significantly higher SPECCA in the ship transport case.

Table 10.2. CO_2 capture and CO_2 avoided for CaL integrated entrained flow cases.

	Base case	Ship transport
CO_2 capture ratio	95%	94%
CO_2 avoided from flue gas	93%	93%
CO_2 avoided	93%	93%
Equivalent CO_2 avoided	89%	85%

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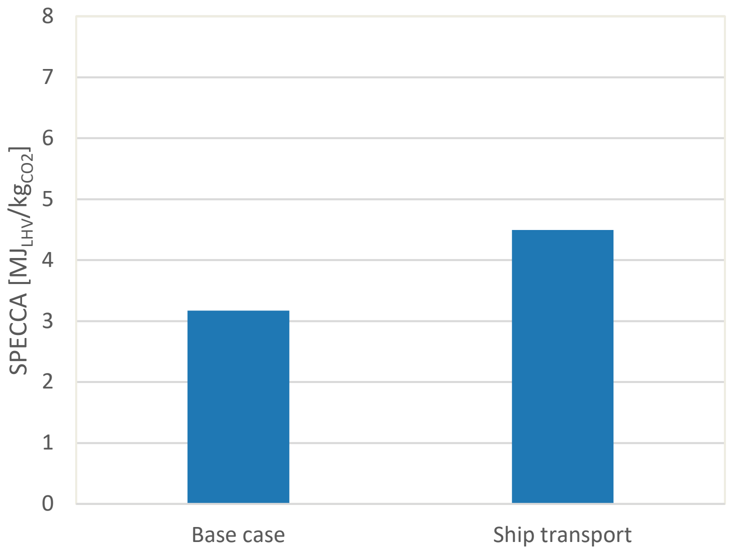

Figure 10.5. SPECCA for Cal integrated entrained flow cases.

Table 10.3. Break down of SPECCA for CaL integrated entrained flow cases.

	Base case	**Ship transport**
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**2528**	**3432**
Coal consumption	2195	2195
Electric power consumption		
Fans and fuel grinding	92	92
ASU	666	666
CPU	943	1821
Cooling water system	39	65
Electric power generation	-1408	-1408
Equivalent specific CO_2 avoided [kg_{CO2}/t_{clk}]	**797**	**764**
At cement kiln stack	808	805
Electric power consumption		
Fans and fuel grinding	-3	-3
ASU	-22	-22
CPU	-31	-61
Cooling water system	-1	-2
Electric power generation	47	47
SPECCA [$MJ_{LHV}/kgCO2$]	**3.17**	**4.49**

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10.3 Economic evaluation

The clinker production and the main utilities and consumables are reported in Table 10.4, while the results of the cost evaluation are presented in Table 10.5. Detailed equipment lists with estimated equipment costs and direct costs are provided in Section 7 in Supplementary information.

With the exception of the ASU and the steam cycle, whose process contingencies are indicated in Table 2.7, the process contingency of the technology is set to 72%. The maturity of the technology is currently low – the concept is formulated, and experiments on small scale support the concept. The process contingency based on technological maturity ($C_{process,1}$) is therefore set to 60 %-points. The level of detail of the equipment list is moderate, so the contribution to process contingency based on this ($C_{process,2}$) is set to 12 %-points.

Table 10.4. Utilities, consumables and operating labor for cement kiln with CaL CO_2 capture.

	Base case	Ship transport
Clinker production [t/h]	117.4	117.4
CO_2 avoided from flue gas [%]	93.4	93.0
CO_2 captured [t/h]	117.1	116.6
Coal [t/h]	23.5	23.5
Electricity consumption [MW_e]	20.41	33.95
Ammonia solution for SNCR [t/h]	0.59	0.59
Limestone [t/h]	87.1	77.7
Cooling water [MW_{th}]	120	120
Operating labor - number of persons	2816.9	2816.9

Table 10.5. Economic results for cement kiln with CaL CO_2 capture.

	Base case	Ship transport
Variable OPEX [€/t_{clk}]	**33.8**	**40.6**
Raw meal	5.0	5.0
Coal	16.3	16.3
Electricity	10.1	16.8
Ammonia for DeNOx	0.7	0.7
Cooling water	0.6	0.8
Other variable costs	1.1	1.1
Fixed OPEX [€/t_{clk}]	**31.6**	**34.2**
Operative, administrative and support labor	11.3	11.5
Insurance and local taxes	9.0	10.1
Maintenance cost (including maintenance labor)	11.3	12.6
Total plant costs, CO_2 capture [M€]	**219.9**	**269.9**
TDC without process contingencies	129.4	157.3
Process contingencies	32.3	41.2
Indirect costs	22.6	27.8
Owner's costs	11.3	13.9
Project contingencies	24.3	29.8
Total plant costs, cement kiln + CO_2 capture [M€]	**423.6**	**473.7**
Total OPEX [€/t_{clk}]	**65.4**	**74.8**
Total CAPEX [€/t_{clk}]	**44.9**	**50.3**
Cost of clinker [€/t_{clk}]	**110.3**	**125.1**
Cost of cement [€/t_{cement}]	**81.1**	**92.0**
CO_2 avoided cost [€/$t_{CO2,avoided}$]	**58.6**	**80.6**

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A break-down of the cost of CO_2 avoided for all calcium looping integrated entrained flow configuration cases is presented in Figure 10.6, together with the cost of clinker in each case.

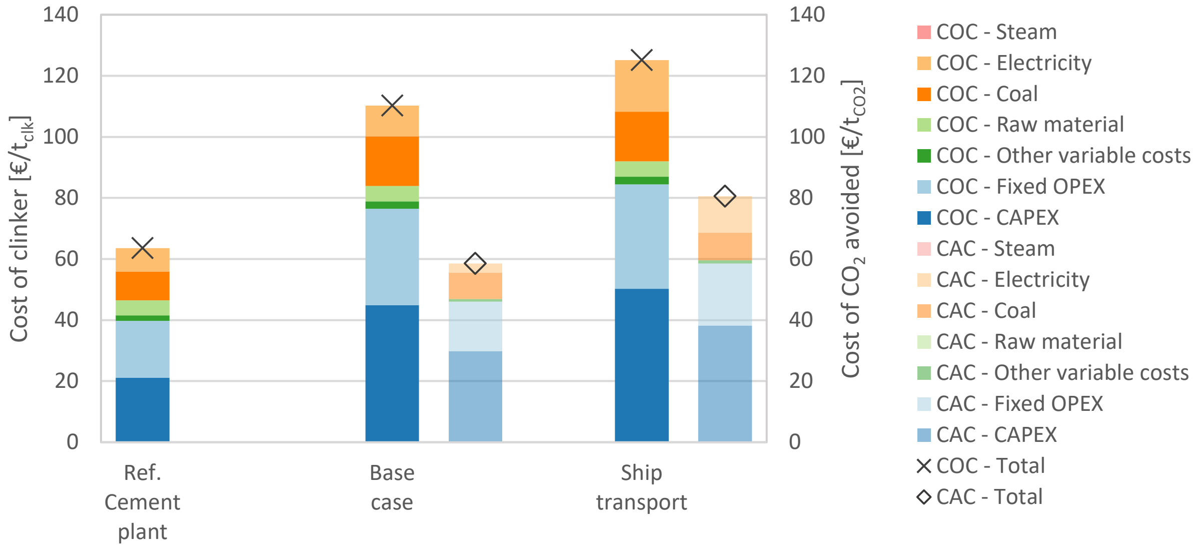

Figure 10.6. Breakdown of the cost of clinker and cost of CO_2 avoided for all CaL integrated EF cases.

For the Integrated EF CaL configuration, a total cost of cement of 81 €/t$_{cement}$ has been estimated, which is 76% higher than the reference plant cement cost.

The results show an increase in operating costs of about 42% compared to the reference cement kiln without capture system. This incremental of operating costs is mainly linked to the increase of fuel and electric power consumptions, which represent about 48% and 30% of the variable OPEX respectively. The increase in CAPEX compared to the reference cement kiln, is mainly due to the presence of ASU and CPU, which represent about 55% of TPC of the CO_2-capture system. The CaL reactors, the sorbent cooler and the supporting structure necessary to support the CaL system are instead responsible for about 18% of the overall TPC increase. A cost of CO_2 avoided of about 59 €/t$_{CO2}$ has been estimated and, as in Tail End CaL cases, the CAPEX is responsible for more than 50%.

The ship transport case results in the largest cost of clinker. The consequent increase in cost of CO_2 avoided is due to higher capital costs and electricity consumption related to the CPU.

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11 COMPARATIVE ANALYSIS

The *CO_2 capture ratio* (CCR), the *CO_2 avoided from the flue gas*, the *CO_2 avoided*, and the *equivalent CO_2 avoided* indicators evaluate the effect that the implementation of a capture technology has on the CO_2 emissions of a plant. These KPIs calculated for the base case for the reference technology MEA and each of the technologies investigated in CEMCAP are given in Table 11.1.

Table 11.1. Base case CO_2 capture ratio and CO_2 avoided.

	MEA	**Oxyfuel**	**CAP**	**MAL**	**CaL - tail-end**	**CaL – Integrated EF**
CO_2 capture ratio (*CCR*)	90%	90%	90%	90%	94%	95%
CO_2 avoided from flue gas (AC_{fg})	90%	90%	90%	90%	91%	93%
CO_2 avoided (*AC*)	70%	90%	78%	90%	91%	93%
Equivalent CO_2 avoided (AC_{eq})	64%	82%	73%	78%	90%	89%

The CCRs of the technologies range between 90% and 95%, and the CO_2 avoided from the flue gas range between 90 and 93%. For the MEA, CAP and MAL technologies, the CO_2 avoided from flue gas is by definition equal to CCR, since there is no change in internal fuel combustion generating CO_2 within the kiln or the capture process when these technologies are installed. For the oxyfuel technology the value is approximately the same as the CCR. The specific fuel consumption changes in the oxyfuel technology compared to the standard reference kiln, but the change is very small. For the calcium looping technology, the CO_2 avoided from the flue gas is lower than the CCR (e.g. 91% *versus* 94% for the tail-end configuration), because the capture of CO_2 generated by fuel combustion within the calciner is not counted as CO_2 avoided from the flue gas. The CO_2 avoided ratios range between 70% and 93%. The ratios decrease compared to the CO_2 avoided from flue gas ratio for MEA (70% *versus* 90%) and CAP (78% *versus* 86%), due to direct emissions related to steam generation. The equivalent CO_2 avoided ratios range between 64% and 90%. In this KPI indirect emissions related to power consumption or generation are taken into account, and the equivalent CO_2 avoided is reduced compared to CO_2 avoided for technologies that consume power, and it is increased for technologies that generate power. The equivalent CO_2 avoided indicator is reduced compared to the CO_2 avoided indicator for all technologies due to larger denominator. The CaL technologies end up with high equivalent CO_2 avoided indicators because there are no additional direct emissions related to these technologies, and either negative or low equivalent CO_2 emissions.

The *specific primary energy consumption for CO_2 avoided* (SPECCA) is defined as the increased equivalent primary energy consumption per equivalent CO_2 avoided, and this indicator can be used to compare the energy consumption of different technologies. The base case SPECCA for the reference technology MEA and the four technologies investigated in CEMCAP is shown in Figure 11.1. All the investigated technologies have clearly lower SPECCAs than the reference technology. The oxyfuel technology has a SPECCA of 1.63 MJ_{LHV}/kg_{CO2}, which is the lowest value among the technologies. The chilled ammonia and membrane-assisted liquefaction technologies have SPECCAs of 3.75 and 3.22 MJ_{LHV}/kg_{CO2} respectively, while the calcium looping tail-end and integrated entrained flow technologies have a SPECCA of 4.07 and 3.17 MJ_{LHV}/kg_{CO2}, respectively.

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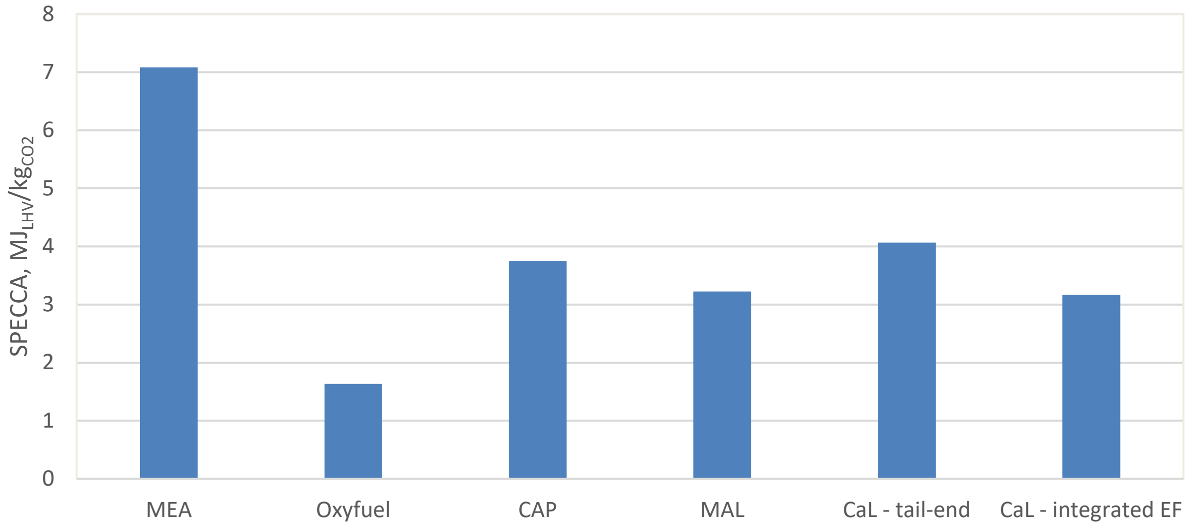

Figure 11.1. Base case specific primary energy consumption for CO_2 avoided (SPECCA).

A break-down of the SPECCA values is given in Table 11.2. More detailed break-downs were given for each technology in Sections 5–10.

The most important contributions to the SPECCA differ among the technologies:

- For the MEA technology the primary energy consumption related to the steam required in the process is responsible for the largest part of the added equivalent primary energy consumption and the reduction in equivalent CO_2 avoided.

- For the oxyfuel technology, the added equivalent primary energy consumption and reduction in equivalent CO_2 avoided are almost entirely due to the increased electric power consumption. The CPU is the largest power consumer, followed by the ASU and the fans. Electric power generation from waste heat reduces the net power consumption by almost one fifth.

- For the chilled ammonia process, the steam consumption makes up the largest part of the primary energy consumption and reduction in equivalent CO_2 avoided. The steam consumption makes up around three quarters of these values, while the electric power consumption is responsible for the rest.

- For the membrane-assisted CO_2 liquefaction process, electric power consumption is responsible for all added equivalent primary energy consumption and reduction in equivalent CO_2 avoided, where around four fifths are due to fan, pump and compressor work in the process, and the rest is mainly due to the refrigeration system.

- For both calcium looping processes, coal consumption, electric power consumption and electric power generation are important for the final SPECCA value. The considerable electric power generation is especially important for the tail-end technology as it contributes to a reduction in both added equivalent specific primary energy consumption

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and equivalent specific CO_2 avoided. This essentially means that the electricity generated covers a part of the cement plant´s demand as well as the demand of the CO_2 capture process.

Table 11.2. Break-down of base case specific primary energy consumption (SPECCA).

	MEA	**Oxyfuel**	**CAP**	**MAL**	**CaL - tail-end**	**CaL – Integrated EF**
Added equivalent specific primary energy consumption [MJ_{LHV}/t_{clk}]	**3959**	**1173**	**2401**	**2216**	**3280**	**2528**
Coal consumption	0	4	0	0	3859	2195
Steam consumption (NG boiler)	3073	0	1859	0	0	0
Electric power consumption	887	1351	542	2216	2086	1740
Electric power generation	0	-182	0	0	-2665	-1408
Equivalent specific CO_2 avoided [kg_{CO2}/t_{clk}]	**559**	**719**	**640**	**687**	**806**	**797**
At cement kiln stack	761	758	762	761	787	808
Steam consumption (NG boiler)	-172	0	-104	0	0	0
Electric power consumption	-30	-45	-18	-74	-70	-58
Electric power generation	0	6	0	0	89	47
SPECCA [$MJ_{LHV}/kgCO_2$]	**7.08**	**1.63**	**3.75**	**3.22**	**4.07**	**3.17**

The main performance indicators used to evaluate the economic performance of the cement plant with CO_2 capture are the *cost of clinker (COC)* and the *cost of CO_2 avoided (CAC)*, together with the *cost of cement*. The economic performance indicators for the base case of all the capture technologies and the reference cement plant without CO_2 capture are presented in Table 11.3.

Table 11.3. Base case economic KPIs.

	Ref. cement plant	**MEA**	**Oxyfuel**	**CAP**	**MAL**	**CaL - tail-end**	**CaL – integrated EF**
Cost of clinker [€/t_{clk}]	62.57	107.4	93.0	104.9	120.0	105.8	110.3
Cost of cement [€/t_{cement}]	46.01	79.0	68.4	77.1	88.2	77.8	81.1
Cost of CO_2 avoided [€/t_{CO2}]	N/A	80.2	42.4	66.2	83.5	52.4	58.6

Figure 11.2 and Figure 11.3 show the breakdown of the cost of clinker and CO_2 avoided cost into the main cost factors. In general, the cost of clinker and cement increases with 49-92% when the investigated CO_2 capture technologies are implemented in the cement plant under base case conditions. The oxyfuel technology shows the lowest cost of clinker compared to the other CO_2 capture technologies, both due to lower variable operating costs and lower capital costs. The absorption-based technologies MEA and CAP as well as both CaL technologies have similar costs, in the range of 105-110 €/t_{clk}. The CaL tail-end technology produces a significant amount of electricity which covers the electricity demand of the CO_2 capture process as well as a part of the cement plant´s demand. As a result, this technology shows a lower electricity cost per ton clinker than the reference cement plant. The MAL technology shows the highest cost of clinker for the base case, with capital costs being the largest individual cost factor.

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Per definition, the CO_2 avoided cost is the difference in cost of clinker between the reference cement plant and the cement plant with CO_2 capture, divided by the equivalent specific avoided CO_2 emissions. In terms of CO_2 avoided, oxyfuel has the lowest cost with the CaL technologies also having relatively low costs, especially the tail-end configuration.

- In the case of MEA, steam contributes most to the cost of CO_2 avoided, both by being responsible for a large increase in the cost of clinker compared to the reference cement plant, and as it entails a reduction in the equivalent specific CO_2 avoided due to its generation from natural gas.

- The oxyfuel technology has by far the lowest cost of CO_2 avoided. It is mainly the increase in capital costs, and associated fixed operating costs, together with increased electricity consumption that contribute to the CO_2 avoided cost. The increased electricity consumption contributes not only to an increase in the cost of clinker but also to a decrease in the equivalent specific CO_2 avoided due to associated CO_2 emissions.

- In the case of CAP, the cost of steam as well as the capital costs and fixed operating costs are the most important factors. Compared to the MEA reference technology, the cost of steam and associated emissions from steam supply are significantly lower for the CAP due to its relatively low specific heat requirement.

- For MAL, high capital costs contribute the most to the cost of CO_2 avoided. A significant share of the cost can also be attributed to the increase in electricity consumption and the consequent increase in the associated indirect CO_2 emissions, as shown in Table 11.2.

- For both calcium looping technologies, the increase in coal consumption contributes significantly to the cost of CO_2 avoided, together with the increase in capital costs. Both CaL technologies generate a significant amount of electric power, with the generation in the tail-end case covering the demand for the capture process and a part of the cement plant´s demand. As a result, the cost of electricity per ton clinker is lower in the CaL tail-end case compared with the reference cement plant. This in turn leads to negative CO_2 avoidance costs associated with electricity consumption in the CaL tail-end base case, as shown in Figure 11.3.

- For electricity-intensive technologies such as MAL, the characteristics of the electricity system and its average CO_2 emission intensity is important and influences the CO_2 avoided cost. This effect is further investigated in the power generation case study in section 11.2.1.

- Process contingencies affect the total direct cost of the technologies, and consequently the added cost factors used to estimate the final total plant cost (cf. Figure 2.1), as well as the fixed operating costs. This might be important for technologies with relatively high capital cost and relatively low technology maturity, such as MAL and the CaL integrated EF process.

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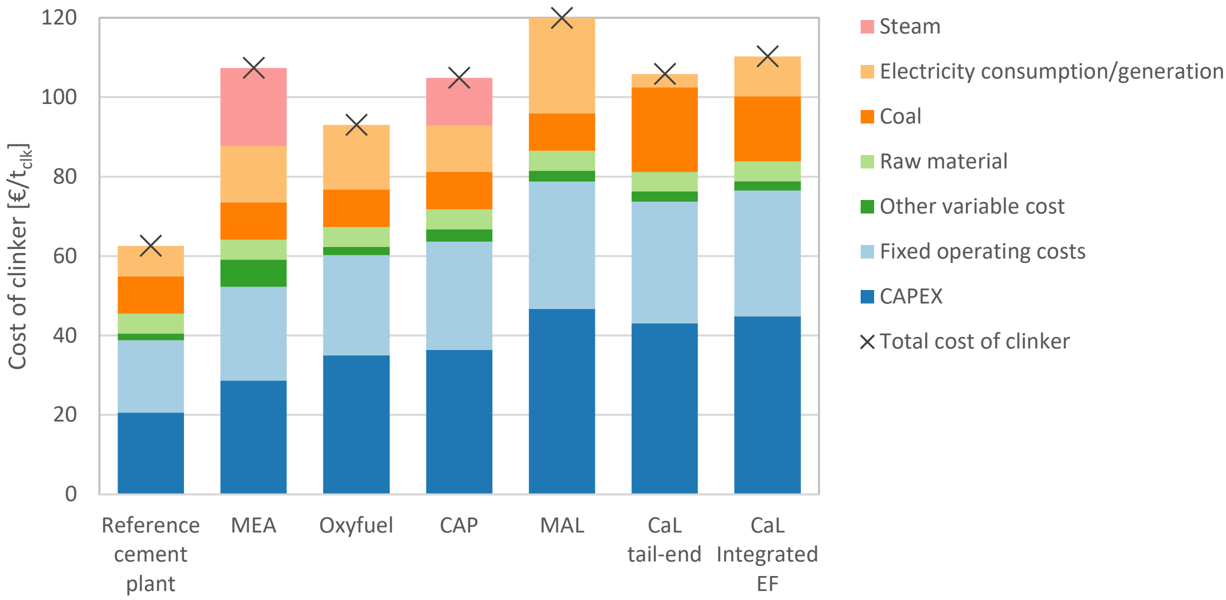

Figure 11.2. Break-down of cost of clinker for the reference cement plant and the base case of all the investigated CO_2 capture technologies.

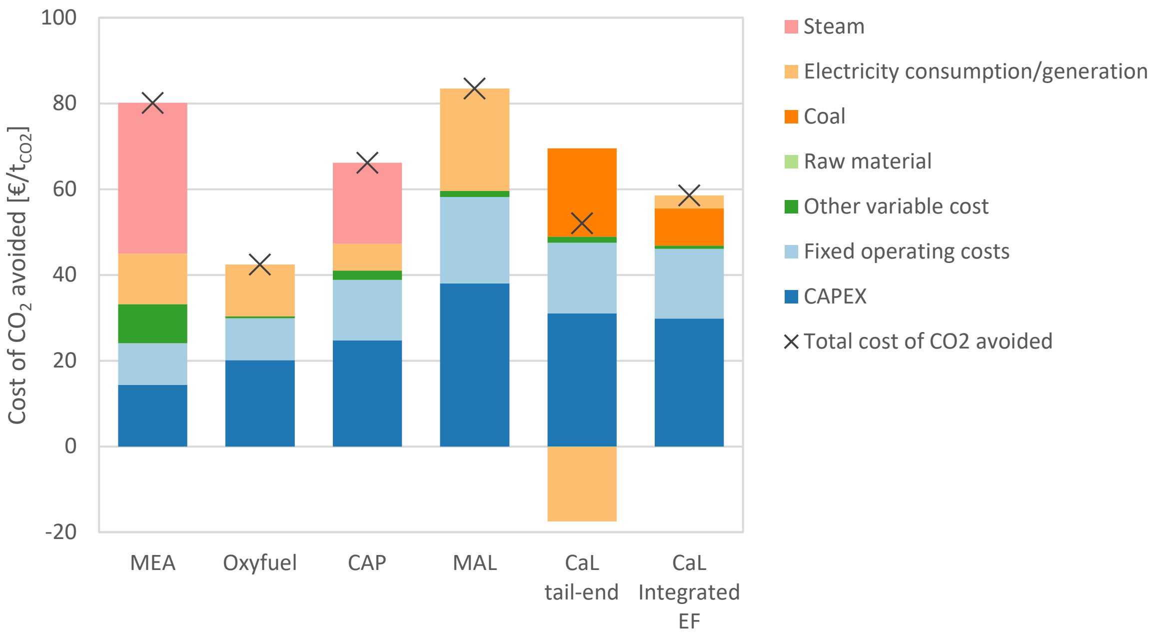

Figure 11.3. Break-down of cost of CO_2 avoided for the base case of all the investigated CO_2 capture technologies.

11.1 Comparison with capital cost estimates from other studies

In the literature, several studies of cost estimations for CO_2 capture from cement plants have been published, and recently gathered in a review by the IEAGHG [IEA, 2018]. As pointed out in the review, the level of detail of the published cost estimations, the methodology used, the

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assumptions and so forth varies considerably, making direct comparison of cost estimates somewhat difficult. In their review, the capital costs of MEA-based CO_2 capture were shown to differ significantly between several studies, although the main difference lies in whether an investment in an on-site CHP plant is included or not.

Fewer techno-economic studies have been published on other CO_2 capture technologies in the cement industry. For oxyfuel, the studies by Gerbelová et al. [GER, 2017] and IEAGHG [IEA, 2013] reported similar capital costs as estimated in CEMCAP. More specifically, Gerbelová et al. [GER, 2017] estimated the capital costs of oxyfuel to be around 2% lower in terms of €/t-clk than estimated in CEMCAP, with the ASU and CPU accounting for 68% of the capital costs (this share is around 70% in CEMCAP). The capital costs of 9.1 €/t cement reported by the IEAGHG [IEA, 2013] are also lower than the 10.6 €/t cement reported in CEMCAP. In this case, the main difference is a higher cost estimated for the ASU and CPU in CEMCAP.

For calcium looping tail-end processes, the study by Ozcan (2014) reports on capital cost, although in this study the integration is slightly different compared with CEMCAP – the treated flue gas is taken from the third preheater stage and the CO_2 depleted gas is sent to the 2nd preheater stage. The costs in this study are given directly as TCR (total capital requirement) and compared with the TPC of the tail-end base case in CEMCAP, the difference in capital cost per ton clinker ranges from -5% to 11%, depending on the integration level.

The published costs for membrane-based technologies in cement applications are difficult to compare with the results from CEMCAP as they are based on a different process concept of multi-stage membrane separation. The CAP technology and the integrated EF technology have not been investigated in cement applications to the same extent as the other technologies when it comes to techno-economics.

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12 SENSITIVITY ANALYSIS

Due to the intrinsic uncertainty and the time-place dependency of some assumptions, the sensitivities of the economic KPIs to the following parameters are investigated by performing a parameter variation in the suggested ranges:

- Coal price: +/- 50% of the reference cost
- Steam supply: +/- 50% of the reference cost
- Electricity price: +/- 50% of the reference cost
- Carbon tax: 0-100 €/t_{CO2}
- CAPEX of CO_2 capture technologies: +35/-15 % of the base case estimate

Further, a case study is performed, investigating the effect of the electricity mix on SPECCA and cost of CO_2 avoided.

12.1 Parameter variation

The sensitivity of the cost of clinker and cost of CO_2 avoided to the parameters listed above are presented in Figure 12.1 - Figure 12.5 for the base case of all CO_2 capture technologies.

The cost of coal has the largest impact on the cost of clinker and CO_2 avoided for the CaL processes, due to the significant increase in fuel consumption associated with these technologies. The cost of CO_2 avoided for the MEA, CAP and MAL technologies is unaffected by the cost of coal since these technologies do not require additional coal consumption.

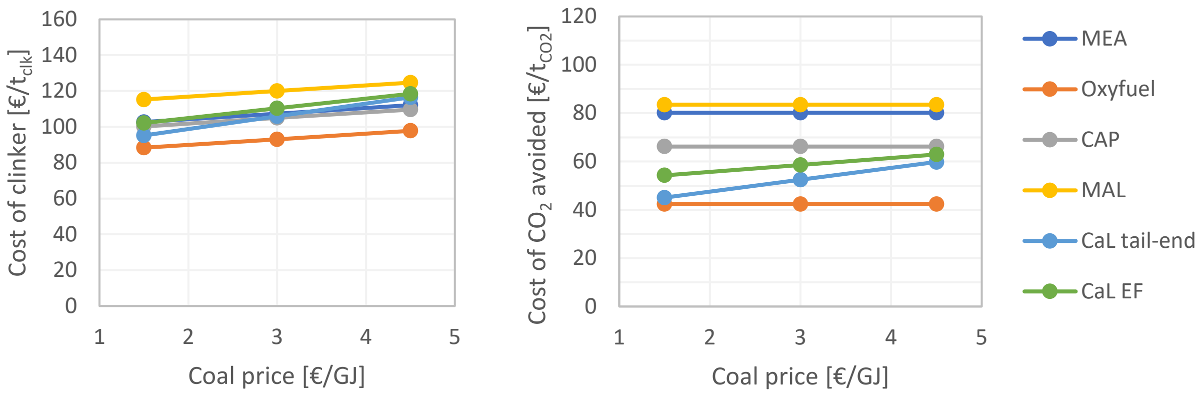

Figure 12.1. Sensitivity of the cost of clinker and CO_2 avoided to +/- 50% change in coal price.

The cost of steam naturally only affects the absorption-based MEA and CAP technologies, especially the MEA technology due to its relatively high steam requirement. At the lower end of the steam cost range, which could represent the cost of steam imported from a nearby coal-fired CHP, the cost of clinker with MEA and CAP are almost the same and just about 6% higher than with oxyfuel. A similar trend is observed for the CO_2 avoidance costs, although there is a larger difference between the lowest cost with oxyfuel and the absorption-based technologies due to the significant CO_2 emissions associated with use of steam.

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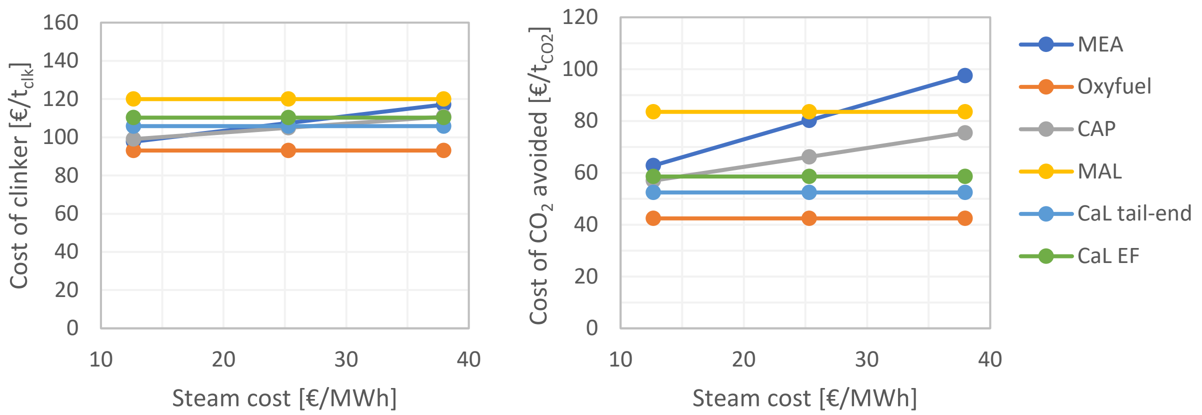

Figure 12.2. Sensitivity of the cost of clinker and CO_2 avoided to +/- 50% change in cost of steam.

Electricity intensive technologies, such as oxyfuel and MAL, are naturally the most sensitive to the price of electricity, as shown in Figure 12.3. The increase in electricity price decreases the cost of CO_2 avoided for the CaL tail-end technology, in contrast to all the other technologies. This is due to that the electricity generated in the CaL process covers a part of the cement plant´s demand and therefore the CO_2 avoidance cost associated with electricity is negative for the CaL tail-end technology.

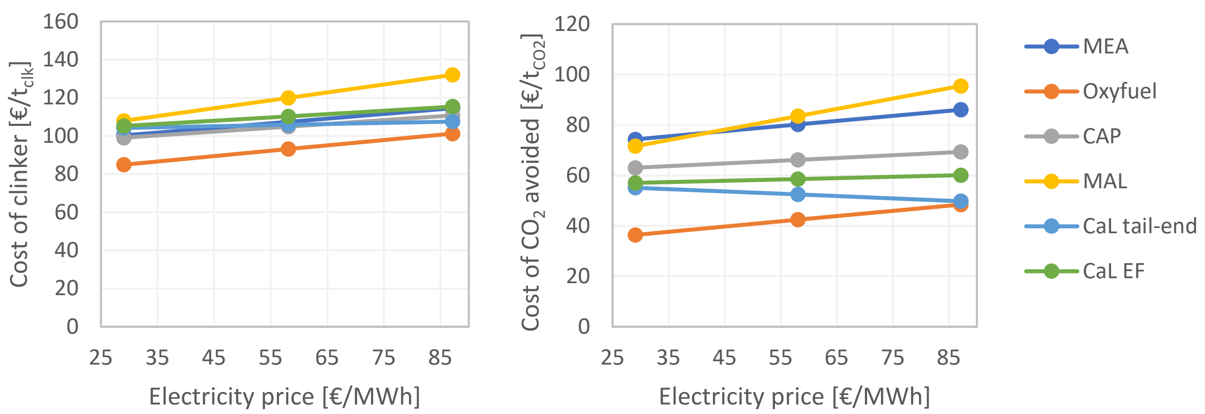

Figure 12.3. Sensitivity of the cost of clinker and CO_2 avoided to +/- 50% change in electricity price.

Should a carbon tax be implemented, the cost of clinker for the reference cement kiln increases drastically, *cf.* Figure 12.4. At a tax level of around 40 €/t CO_2, the cost of clinker with oxyfuel technology becomes lower than in the reference cement kiln, and at roughly 60 €/t CO_2 the CAP and both CaL technologies will have a lower cost of clinker compared with the cement kiln without CO_2 capture. For MEA and MAL, a carbon tax of around 75 €/t CO_2 would be required for a clinker cost lower than that of the reference cement kiln. Due to the direct CO_2 emissions from on-site steam generation for CO_2 capture with MEA and CAP, and therefore lower CO_2 avoided, these technologies are more sensitive to a carbon tax than the other CO_2 capture technologies.

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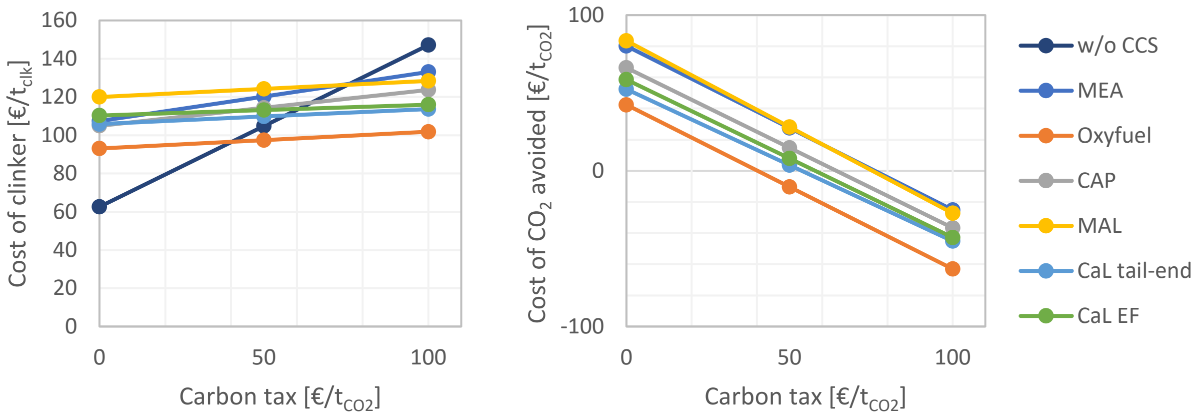

Figure 12.4. Sensitivity of the cost of clinker and CO_2 avoided to a carbon tax of 0-100 €/t CO_2. The effect on the cost of clinker for the reference cement kiln is included for comparison.

The most capital-intensive technologies, MAL and both CaL processes, are the most sensitive to a change in the capital cost estimate. The oxyfuel and CAP technologies are also significantly affected while the smallest effect is seen for the reference capture technology MEA, which has the lowest capital cost. It should be noted that the estimated fixed operating costs are also affected by a change in the capital costs.

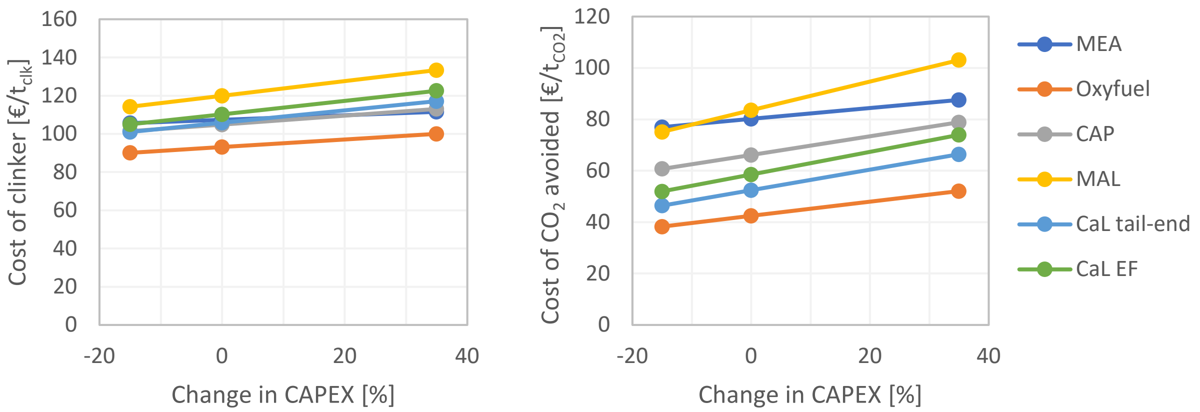

Figure 12.5. Sensitivity of the cost of clinker and CO_2 avoided to +35/-15% change in estimated CAPEX relative to the base case estimation.

12.2 Power generation case study

The characteristics of the power generation system in terms of electricity generation efficiency, η_{el}, and the specific CO_2 emissions of the electricity generation, e_{el} have an impact on the SPECCA and the cost of CO_2 avoided, as described in section 2.3. The generation efficiency and specific CO_2 emissions are directly linked to the power generation technology that is assumed to provide the electricity required by the processes. Values for η_{el} and e_{el} are defined in the CEMCAP framework based on the share of different fuel sources in EU-28 in 2014.

In order to investigate the impact of the values of η_{el} and e_{el}, SPECCA values and the cost of CO_2 avoided are calculated with several different options for power generation. The investigated cases are summarized in Table 12.1.

Table 12.1. Power generation cases with corresponding electricity generation efficiency and specific CO_2 emissions.

Power generation case	**η_{el} [%]**	**e_{el} [kg/MWh]**
CEMCAP framework (EU 2014)	45.9	262
Pulverized coal (PC) – ultra-supercritical state-of-the-art	44.2	770
Pulverized coal (PC) – sub-critical	35	973
Natural gas combined cycle (NGCC)	52.5	385
Renewables	100	0

The SPECCA for the base case of all CO_2 capture technologies, calculated for the different power generation cases are presented in Figure 12.6. For the calcium looping tail-end technology, the SPECCA increases with increasing electricity generation efficiency and decreasing specific CO_2 emissions, while the opposite is observed for all the other CO_2 capture technologies. This is because the CaL tail-end technology generates enough electricity to cover both its own demand and a part of the electricity demand of the cement plant, effectively substituting some of the electricity that was bought from the grid in the reference cement plant. With increasing generation efficiency of the power system and a decrease in the associated specific CO_2 emissions, the reduction in indirect added equivalent specific primary energy consumption becomes smaller, as well as the indirect avoided equivalent specific CO_2 emissions.

For oxyfuel and MAL technologies, where use of electricity is the largest variable cost factor, the SPECCA value is highly dependent on the characteristics of the power generation system. In the case of electricity being solely generated from renewables, the SPECCA is reduced by more than half compared to the EU 2014. On the other hand, the SPECCA of the MAL technology is almost doubled in the worst case of electricity generation from sub-critical PC plants. The SPECCA of the MEA and CAP technologies is also significantly affected by the different power generation cases.

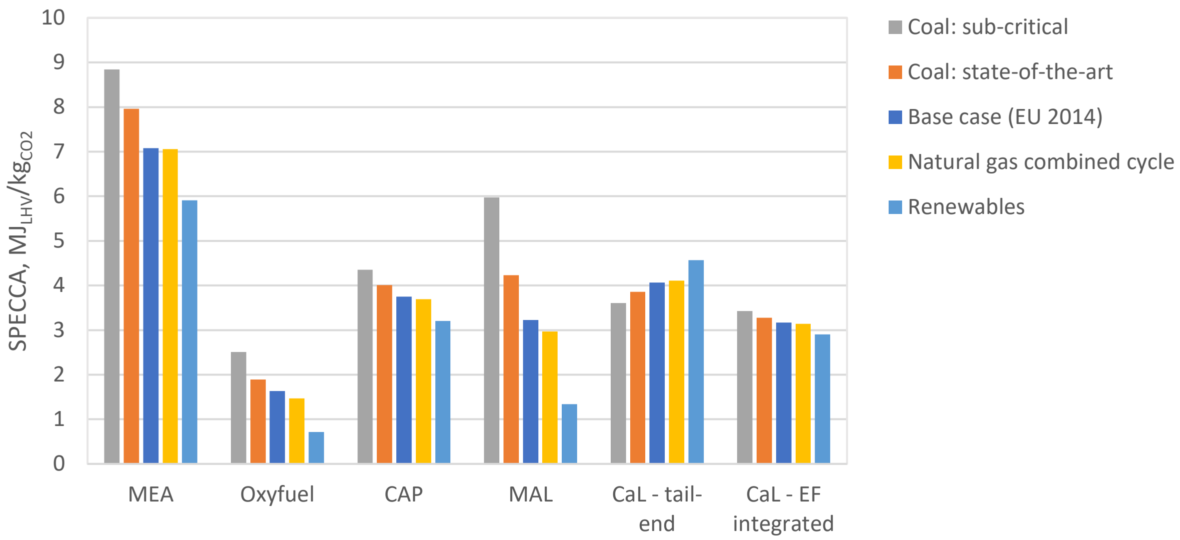

Figure 12.6. SPECCA for the base case of all CO_2 capture technologies with different cases for power generation.

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Figure 12.7 shows how the cost of CO_2 avoided for the base case of all CO_2 capture technologies is affected by the power generation system characteristics (with unchanged electricity price). Similar trends are observed for the cost of CO_2 avoided as for the SPECCA, although less prominent. Recalling that SPECCA is calculated as the added equivalent specific primary energy consumption divided by the avoided equivalent specific CO_2 emissions (as shown in section 2.3.1), both factors are affected by the power generation cases. For the cost of CO_2 avoided, which is defined as the difference in cost of clinker between the reference cement plant and the cement plant with CO_2 capture, divided by avoided equivalent CO_2 emissions, the power generation cases only affect the avoided equivalent CO_2 emissions while the difference in cost of clinker remains unchanged.

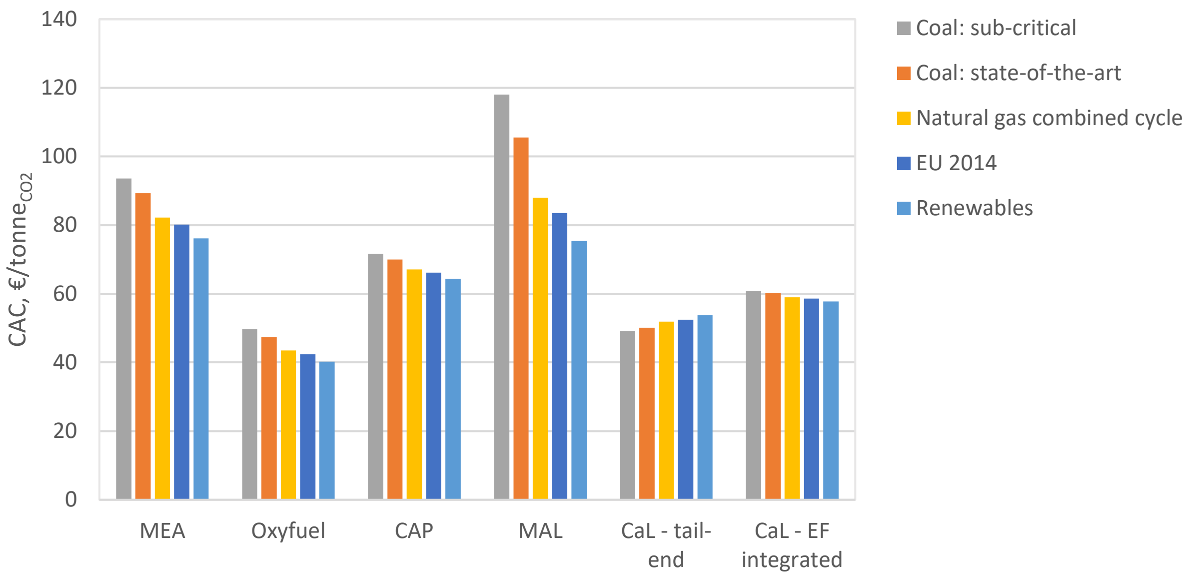

Figure 12.7. Cost of CO_2 avoided for the base case of all CO_2 capture technologies with different cases for power generation.

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13 TECHNOLOGY OUTLOOK

Through the techno-economic analysis of the technologies, several possibilities for improved performance, both in terms of SPECCA and cost can be identified.

For the solvent technologies MEA and CAP, it was observed that the steam generation contributed to the major part of the equivalent specific CO_2 emissions (which reduce the CO_2 avoided). In the base case, the equivalent specific CO_2 avoided was reduced by 23% and 13% by the emissions from the NG fired boiler (Table 5.3 and Table 7.4). To reduce these emissions, and thus decrease the SPECCA and the cost of CO_2 avoided, it can be considered to also capture CO_2 from the boiler flue gas by adding this gas stream to the cement kiln flue gas before it is sent to the capture process.

An alternative measure to reduce or omit the direct CO_2 emissions related to the steam generation is to use sustainably harvested biomass as fuel in the boilers. If CO_2 also is captured from the flue gas, this could give a positive contribution to the amount of CO_2 avoided, as net negative emissions would be related to the steam generation part of the process. The use of biomass could also be considered in the calcium looping process.

The CAP and MAL post-combustion technologies showed improved performance in terms of SPECCA and the two economic KPIs in the case of low air leak in the mill, i.e. with higher CO_2 content in the flue gases and a lower flue gas volume flow. For these technologies, measures to reduce air leak further should be considered. The major part of the air leak in the process takes place in the raw mill. In the reference cement plant, the CO_2 concentration in the flue gas is 34% at the preheater outlet and 20-25% after the raw mill (dry basis) [VOL, 2018]. It would therefore be beneficial if the direct drying of raw material in the raw mill by the flue gas could be omitted. One possibility could be to adopt a similar approach for these technologies as for oxyfuel, where heat is transferred indirectly from the flue gas to the raw mill by preheating air that is sent through the mill. This process modification will entail some increased capital costs for additional heat exchangers, and might also entail investment in a new filter, but it could potentially result in significantly improved CO_2 capture process performance. It should be mentioned that the feasibility of this option depends on the amount of heat required in the raw mill. The heat requirement depends on the moisture content of the raw meal, and this varies significantly from plant to plant.

Further, there exist cement plants where the raw meal has such a low moisture content that only part of the flue gas is needed for drying in the raw mill, and CO_2 capture could be performed from the remaining part of the flue gas, that has a high CO_2 content and a relatively low volume (the Norcem plant in Brevik is an example). The performance of the CAP and the MAL technologies would benefit from such conditions. In addition, the reduced requirement for drying of raw material results in higher amount of available waste heat compared to the CEMCAP reference cement plant, which would further benefit the CAP technology.

As mentioned in Section 8, there is room for improvement of the performance of the MAL technology be finding a membrane that is more suitable for the system. For this technology a screening of different membranes, preferably with testing in real conditions at a cement plant, and further optimization of the system with a more optimal membrane would be the next step.

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For several technologies, it was observed that process contingencies contributed heavily to the CAPEX. The process contingencies account for costs that are unknown, and the relative amount of unknown costs are assumed to be higher for technologies with lower maturity. The process contingencies are particularly important for the costs of the MAL and the CaL integrated EF technologies. For all technologies, especially MAL and CaL integrated EF, increasing the maturities by further development and demonstration of the technologies for the cement application would reduce the uncertainty of the costs, and possibly reduce the overall costs of the technologies. Such development is currently taking place for the CaL integrated EF technology in the CLEANKER project.

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14 CONCLUSION

A techno-economic evaluation of the CEMCAP technologies applied to a reference cement kiln has been carried out. Key performance indicators such as specific primary energy consumption, cost of clinker and cost of CO_2 avoided have been calculated. The calculations are based on process simulations, with input from experimental work carried out in the project. Detailed equipment lists for the CO_2 capture technologies with equipment and installations costs for each item are included in the supplementary material document to this report.

The evaluation shows that the CEMCAP technologies generally outperform the reference technology (MEA) with respect to specific primary energy consumption. In the base case, all CEMCAP technologies have specific primary energy consumption values lower than 60% of the specific primary energy consumption of the reference technology MEA-based CO_2 absorption.

The cost of clinker increases with 49-92% when the technologies are implemented in the reference cement plant under the base case conditions. The cost of CO_2 avoided lies between 42 €/tCO_2 (oxyfuel process) which is approximately halved compared to MEA, and 84 €/tCO_2 (membrane-assisted CO_2 liquefaction process), which is on the same level as MEA.

The KPI calculations rely on assumptions related to steam source, electricity mix, electricity price, fuel price, carbon tax, *etc.* A sensitivity analysis is performed, showing a strong dependency on such variables both for the specific primary energy consumption and for the economic indicators. The evaluation presented here is performed for application to a Best-Available-Technology reference cement kiln, but it should be noted that cement plants vary in general significantly from each other, for instance when it comes to CO_2 concentration in the flue gas and availability of waste heat.

In addition to the typical techno-economic parameters such as specific primary energy consumption, the cost of clinker and the cost of CO_2 avoided, it should be mentioned that there are several other aspects that are important for evaluation and practical implementation of technologies for CO_2 capture from a cement plant. Important properties are technology maturity, integration with the clinker burning process and possible effects on product quality (and therefore risk), space requirement, the need for utilities, such as power or natural gas, *etc.* Such properties are evaluated in a CEMCAP report on retrofitability [HOP, 2018].

The technologies investigated within CEMCAP are fundamentally different from each other. They provide a portfolio of technologies with different properties, suitable for application in a wide variety of conditions in cement plants. For the selection of a CO_2 capture technology for a specific plant, a plant-specific evaluation of primary energy consumption and cost should be performed. The excel model used for cost evaluations within CEMCAP is available online [DEL, 2018], and could be used for this purpose. Further, plant specific evaluation of more practical properties such as available space, capacity in local power grid, options for steam supply etc. should also be carried out.

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APPENDICES

A DEHYDRATION PROCESSES

A.1 TEG dehydration

In the MEA and CAP processes, CO_2 is compressed and dehydrated to reach the specifications for pipeline transport. The water content in the CO_2 should be maximum 300 ppm_{wt}, and it is reasonable to use a TEG process to reach this level [KEM, 2014]. In the TEG process (Figure C.1), the water vapor in the gas (CO_2) stream is absorbed by TEG in an absorber. The TEG is then regenerated in a stripper where the boiler temperature is around 170 °C. The temperature is higher than the stripper temperatures in both the MEA and the chilled ammonia capture process. According to [POE, 2017], heat to the reboiler can be by a fired heater or electrical heater. Electrical heaters are assumed to be used in the CO_2 compression process in order to avoid the need for a small amount of steam at another pressure level. The specific heat demand of the process is 2.62 kJ per kg saturated CO_2, and the electric heater efficiency is assumed to be 95%. The specific power demand of the process due to pumping of TEG is 0.045 kJ per kg saturated CO_2.

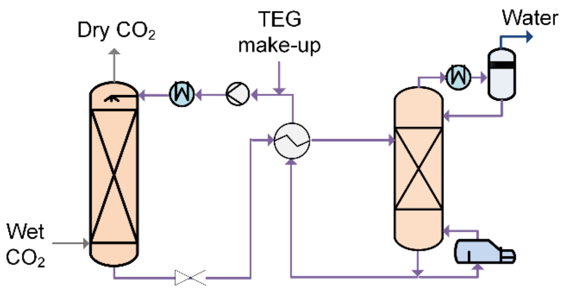

Figure C.1. TEG dehydration process.

A.2 Molecular sieve dehydration

In processes involving liquefaction of CO_2, it is necessary to reach lower water concentrations in the CO_2 in order to avoid solid formation. In CEMCAP this applies to the CO_2 liquefaction process for ship transport (relevant for MEA and CAP), the CPUs (relevant for oxyfuel and CaL) and the membrane-assisted liquefaction process. It is reasonable to use molecular sieves to reach very low water concentrations [KEM, 2014]. In order to find an approximate value for specific heat demand, the molecular sieve dehydration process described by Kemper et. al. [KEM, 2014] is assumed. In this process the wet CO_2 stream is sent through one out of two dehydration vessels, in which water is adsorbed by molecular sieves. Heat is required for reheating of a small fraction of the dry CO_2 stream which is used to regenerate the molecular sieves. The heat consumption corresponds to 16.6 MJ/kg_{H2O}.

Also in the ASUs (relevant for oxyfuel and CaL), dehydration by adsorption is required before liquefaction of the air. It is assumed that the heat requirement is 58.3 kJ/kg_{O2} [IEA, 2005].

In the CaL processes, the heat required for dehydration is assumed to be provided by steam, while for the other technologies, the heat is assumed to be provided by an electric heater (95% efficiency).

B NOX REMOVAL

According to [IEA, 2008], the concentration of NO_2 should be limited to 41 mg/Nm^3 for MEA from an economic point of view. Considering that the NO_2 is no more than 10% of the total NO_x emissions, the total NO_x should be limited to 410 mg/Nm^3 for MEA CO_2 capture applications. The same limit is assumed for the membranes in the Membrane-assisted CO_2 Liquefaction (MAL) process.

A selective non-catalytic reduction (SNCR) system is assumed to be installed in the reference cement kiln for control of NOx emissions. The SNCR process is a state-of-the-art technology in the cement industry and is applied widely in European cement kilns. The NOx reduction is achieved through injection of 25% ammonia as a reducing agent into the flue gas stream. The reduction of NOx takes place according to the following chemical reaction:

$$4\ NO \quad + \quad 4\ NH_3 \quad + \quad O_2 \quad \rightarrow \quad 4\ N_2 \quad + \quad 6\ H_2O$$

The SNCR system is assumed to operate with a long-term average reduction rate of 60% with an ammonia consumption of 5 kg/t_{clk}. The resulting NOx concentration of the cement kiln flue gases is 500 mg/Nm_3 on dry basis at 10% O_2 content, which translates to 591 mg/Nm^3 during the first part of the year (in low air leak conditions), and 455 mg/Nm_3 during the second part of the year (in medium air leak conditions). An increase in NOx removal rate to reach the suitable emission levels for the MEA and MAL technologies is achieved by increasing the injection rate of ammonia solution in the SNCR system, with 1.5 times the stochiometric amount (limited to maximum 80% removal rate).

C SOX REMOVAL

DeSO$_x$ systems are normally not installed in conventional cement plants since most SOx are absorbed in the clinker and the raw mill [IEA, 2008]. However, when the MEA CO_2 capture process is included, the MEA can react with acid gases such as SOx and NOx to form amine salts. The SOx in the flue gas should typically be limited to 10 ppmv prior to the MEA CO_2 capture process. The same limit is also assumed for the membrane in the MAL process.

The reduction of SO_x to acceptable levels is achieved through injection of 50% NaOH solution in the direct contact cooler (DCC) present in both the MEA and the MAL processes. The reduction takes place according to the following chemical reaction, where all SO_x is assumed to be in the form of SO_2:

$$2\ NaOH \quad + \quad SO_2 \quad \rightarrow \quad Na_2SO_3 \quad + \quad H_2O$$

The SO_x concentration of the cement kiln flue gases is 200 mg/Nm^3 on dry basis at 10% O_2 content, which translates to 236 mg/Nm^3 during the first part of the year (in low air leak conditions), and 182 mg/Nm^3 during the second part of the year (in medium air leak conditions). All SO_x is assumed to be removed in the process, with stochiometric amount of the NaOH solution sufficient to achieve the reduction.