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

D3.2
CEMCAP framework for comparative techno-economic analysis of CO$_2$ capture from cement plants

Revision 1

Due delivery date: 2017-01-31
Actual delivery date: 2017-05-11
Revised version delivery date: 2017-09-18

Organisation name of lead participant for this deliverable:
SINTEF-ER

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

Deliverable number:	D3.2
Deliverable title:	CEMCAP framework for comparative techno-economic analysis of CO_2 capture from cement plants
Work package:	WP3: CEMCAP Framework
Lead participant:	SINTEF-ER

Author(s)		
Name	Organisation	E-mail
Rahul Anantharaman	SINTEF ER	Rahul.Anantharaman@sintef.no
David Berstad	SINTEF ER	david.berstad@sintef.no
Giovanni Cinti	C.T.G SPA	g.cinti@itcgr.net
Edoardo De Lena	Politecnico di Milano	edoardo.delena@polimi.it
Manuele Gatti	Politecnico di Milano	manuele.gatti@polimi.it
Matteo Gazzani	ETH Zurich	matteo.gazzani@ipe.mavt.ethz.ch
Helmut Hoppe	VDZ GmbH	helmut.hoppe@vdz-online.de
Isabel Martínez	Politecnico di Milano	isabel.martinez@polimi.it
Juliana Garcia Moretz-Sohn Monteiro	TNO	juliana.monteiro@tno.nl
Matteo Romano	Politecnico di Milano	matteo.romano@polimi.it
Simon Roussanaly	SINTEF ER	simon.roussanaly@sintef.no
Erin Schols	TNO	erin.schols@tno.nl
Maurizio Spinelli	Politecnico di Milano	maurizio.spinelli@polimi.it
Sigmund Størset	SINTEF ER	Sigmund.Storset@sintef.no
Peter van Os	TNO	peter.vanos@tno.nl
Mari Voldsund	SINTEF ER	Mari.Voldsund@sintef.no

Abstract

This document provides a framework for comparative techno-economic analysis in the CEMCAP project, where four different CO_2 capture technologies are to be evaluated for application in cement plants. It defines a reference cement kiln with description of the main unit, and characteristics of raw material and flue gas. The four capture technologies (oxyfuel, chilled ammonia, membrane assisted liquefaction, and calcium looping), and a reference technology (MEA absorption) are described. Specifications are set for process units (e.g. heat exchangers, compressors etc.) and for the generation of utilities (e.g. steam, electric power etc.). Conditions like the extent of CO_2 capture and purity of the produced CO_2 are set. Furthermore, relevant economic parameters are defined, and parameters relevant for sensitivity studies are suggested.

Changes in revision 1

Clarifications are included on the definition of primary energy for renewable power production (Chapter 6.2.1) and cost of electricity in EU28 (Table 6.6). The new ISO 27913 standard on CO_2 pipeline transportation systems is commented on in relation to the CO_2 impurity specifications (Chapter 8.1.1). Temperature profiles in Fig. 3.6 are corrected. CEPCI for 2014 was updated with final 2014 value instead of a preliminary value.

TABLE OF CONTENTS

List of abbreviations

AACE:	Association for the Advancement of Cost Engineering
AC:	CO_2 avoided
ASU:	Air Separation Unit
BAT:	Best Available Technique
BREF:	BAT Reference
BUA:	Bottom-Up Approach
CAC:	Cost of Avoided CO_2
CaL:	Calcium Looping
CAP:	Chilled Ammonia Process
CCR:	CO_2 Capture Ratio
CCS:	Carbon Capture and Storage
CEPCI:	Chemical Engineering Plant Cost Index
CHP:	Combined Heat and Power
clk:	clinker
COC:	Cost of Clinker
COP:	Coefficient of Performance
CPU:	CO_2 Purification Unit
EOS:	Equation of State
EPC:	Engineering, Procurement and Construction Costs
FGD:	Flue Gas Desulphurizer
GT:	Gas Turbine
HRSG:	Heat Recovery Steam Generator
KPI:	Key Performance Indicator
LHV:	Lower Heating Value
LOI:	Loss on Ignition
MEA:	Monoethanolamine
O&M:	Operations & Maintenance
ORC:	Organic Rankine Cycle
PM:	Particulate Matter
PSA:	Pressure Swing Adsorption
R&D:	Research and Development

RDF:	Refuse Derived Fuel
SPECCA:	Specific Primary Energy Consumption for CO_2 Avoided
STP:	Standard Temperature and Pressure
TDC:	Total Direct Costs
TEC:	Total Equipment Costs
TEG:	Triethylene Glycol
TOC:	Total Organic Carbon
TOT:	Turbine Outlet Temperature
TPC:	Total Plant Costs
TRL:	Technology Readiness Level
VPSA:	Vacuum Pressure Swing Adsorption
WTE:	Waste-to-Energy

1   INTRODUCTION

This document provides a framework for techno-economic analyses in the CEMCAP project. It will ensure a consistent benchmarking of CO_2 capture technologies in the cement industry. It is based on the preliminary work of the European Benchmarking Task Force which has developed a framework for benchmarking of power plants with CO_2 capture. Due to the different scope of CEMCAP, additional information about the clinker burning process and a reference cement kiln had to be included in the present framework document. A description of the CO_2 capture technologies, which are investigated within the project, is furthermore contained in the document.

The aim of the framework document is:

- to define a reference cement kiln where different CO_2 capture technologies could be applied,
- to define boundary conditions (e.g. energy prices, investment related basis values, CO_2 quality), assumptions (e.g. by reference location) and sensitivities (e.g. economic parameters) to compare capture methods,
- to assemble a framework for simulations and evaluation of full-scale cement kilns with CO_2 capture; several capture cases/scenarios will be employed for the technology evaluation,
- to define key performance indicators (KPIs) for benchmarking; these will include (but are not limited to) energy consumption, required/produced heat, required/produced power, cost of CO_2 captured, cost of CO_2 avoided and other techno-economic parameters.

The benchmarking framework developed in CEMCAP will be disseminated in several different ways with the ECRA members as primary receiver. However, curbing CO_2 emissions from the cement industry is a global challenge, and consequently efforts will be made to spread the document globally and beyond the lifetime of CEMCAP. It is expected that the CEMCAP framework can also be a useful reference for the R&D directed towards implementation of CCS technologies in other CO_2-intensive industries.

2 GENERAL DEFINITION AND CONDITIONS

2.1 Units

In all documents SI units (Système International d'Unitès) are used [BIPM].

2.2 Reference state for gas concentrations

Gas concentrations are referred to standard temperature (273.15 K) and pressure (1,013 hPa) conditions (STP) in the dry state.

In some cases, gas concentrations are referred to a reference oxygen concentration, e.g. 10 % O_2. The following equation can be used to correct a measured pollutant concentration in a dry emitted of a conventional cement plant with a measured O_2 content to an equivalent pollutant concentration in a dry emitted gas with a specified reference amount of O_2:

$$E_B = \frac{21 - O_B}{21 - O_M} \times E_M$$

E_M:	mass concentration measured
E_B:	mass concentration, as related to reference oxygen content
O_M:	oxygen content measured
O_B:	reference oxygen content

2.3 Ambient conditions

The reference cement kiln is assumed to be situated on inland, with the following ambient conditions:

- Air temperature: 15 °C
- Air pressure: 1.013 bar
- Relative humidity: 60 %

The air composition is given in Table 2.1.

Table 2.1: Air composition

Component	Unit	Concentration	
		Dry air	**Air at 60 % relative humidity**
N_2	vol%	78.09	77.30
CO_2	vol%	0.03	0.03
H_2O	vol%	0	1.01
Ar	vol%	0.932	0.923
O_2	vol%	20.95	20.74

3 REFERENCE CEMENT KILN

The reference case relies on the Best Available Technique (BAT) standard as defined in the European BREF-Document (**B**est Available Technique **Ref**erence) for the manufacture of cement [BREF, 2013]. The plant structure for the reference case, based on a dry kiln process, consists of a five stage cyclone preheater, calciner (also called precalciner) with tertiary duct, rotary kiln and grate cooler, as illustrated in Figure 3.1. This case is used as reference also in the ECRA project, and a process model has been built by VDZ (in this document referred to as the VDZ process model).
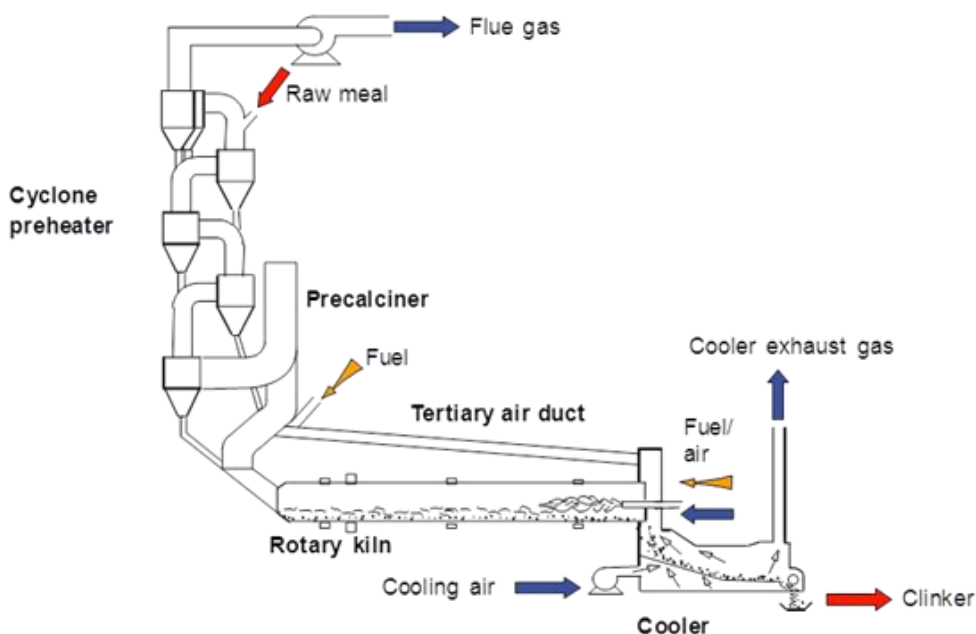
Figure 3.1: BAT cement kiln

3.1 Size

A BAT cement kiln exhibits a clinker capacity of 3,000 t/d (raw meal/clinker factor 1.6), which is a representative size for a European cement kiln (see Table 3.1). This corresponds to a yearly clinker production of 1 Mt (equivalent to a run time of >330 days per year) or a cement production of 1.36 Mt per year (clinker/cement factor 0.737).

Table 3.1: Production characteristics of a BAT cement kiln

Parameter	Value
Production capacity	1 Mt_{clk}/y (3,000 t_{clk}/d)
Cement production	1.36 Mt_{cement}/y
Clinker/cement factor	0.737
Raw meal/clinker factor	1.6
Specific CO_2 emissions	850 kg_{CO2}/t_{clk}
Specific total electricity demand	97 kWh/t_{cement}

3.2 Raw material

3.2.1 Raw meal

The raw materials for the clinker burning process are calcareous deposits, such as limestone, marl or chalk. These materials provide the source for calcium carbonate. In most cases, corrective materials (e.g. other natural raw materials or so-called secondary/alternative raw materials from other production processes) are required for an appropriate raw material mixture to provide the required amounts of other oxides like SiO_2, Fe_2O_3 and Al_2O_3. For the so-called BAT plant, an average raw material moisture (before drying) of 6 wt% was defined. The composition of a typical raw meal (that has been dried while grinding) and of a clinker is shown in Table 3.2.

Table 3.2: Composition of a typical raw meal (including correcting components) and of a clinker

Raw meal composition at preheater inlet			Clinker composition[2]		
Parameter	**Unit**	**Value**	**Parameter**	**Unit**	**Value**
SiO_2	wt%	13.80	C_3S	wt%	64.3
Al_2O_3	wt%	3.25	C_2S	wt%	14.0
TiO_2	wt%	0.06	C_3A	wt%	10.0
P_2O_5	wt%	0.04	C_4AF	wt%	8.6
Fe_2O_3	wt%	1.96	CaO	wt%	0.7
Mn_2O_3	wt%	0.05	Cl	wt%	0.015
CaO	wt%	43.22	SO_3	wt%	0.70
MgO	wt%	0.71	K_2O	wt%	0.85
CO_2	wt%	34.74	Na_2O	wt%	0.19
SO_3	wt%	0.34	LSF (lime saturation factor)		95
Sulphide[1]	wt%	0.05	Alumina ratio		1.8
K_2O	wt%	0.55	Silica ratio		2.5
Na_2O	wt%	0.12	Degree of sulphatisation		72
Cl	wt%	0.01			
TOC	wt%	0.10			
H_2O	wt%	1.00			

[1] SO_2 can be generated from the sulphides in the preheater

[2] Other trace components are not mentioned which make up the deficit up to 100 %.

The compositions of the raw meal and clinker are expressed as oxides, which is the common practice in the cement industry. Shortened forms are used to simplify the chemical formulae of the clinker phases, in which:

$$C = CaO$$
$$S = SiO_2$$
$$A = Al_2O_3$$
$$F = Fe_2O_3$$
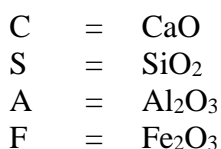
3.2.2   Fuel

In the reference case coal (residue on 90 µm sieve < 15 %) is used as fuel for the clinker burning process. At many cement plants alternative fuels (e.g. RDF or a mixture from RDF, animal meal and sewage sludge) are taken into use to replace some of the coal. Fuel characteristics are shown in Table 3.3. The characteristics of RDF are included for reference purposes.

Table 3.3: Fuel characteristics of coal and RDF

Parameter	Unit	Coal	RDF
Calorific value (LHV)	kJ/kg	27,150	22,000
Ash	wt%	16.50	9.00
Moisture	wt%	0.50	12.00
Cl	wt%	0.02	0.40
C	wt%	69.00	55.50
H	wt%	4.00	8.50
S	wt%	0.50	0.15
N	wt%	0.48	0.25
O	wt%	9.00	14.20
Ash composition			
SiO_2	wt%	41.40	17.20
Al_2O_3	wt%	27.30	55.00
TiO_2	wt%	1.40	4.60
P_2O_5	wt%	1.10	0.40
Fe_2O_3	wt%	4.00	0.80
Mn_2O_3	wt%	0.10	0.05
CaO	wt%	18.20	17.50
MgO	wt%	1.70	0.95
SO_3	wt%	4.00	0.80
K_2O	wt%	0.60	0.60
Na_2O	wt%	0.20	2.10

3.3   Preheater

Cement clinker is mainly produced in dry processes making use of cyclone preheaters. The cyclone preheater generally consists of three to six cyclone stages (depending on the raw material moisture) arranged above one another. The raw meal (kiln feed) passes through the process stages of preheating and calcining in succession from top to bottom before reaching the rotary kiln. The flow of process gases is essentially counter to the flow of the kiln feed. Each cyclone stage is made of two parts: the connecting duct which connects each cyclone to the one above, where the material and gas phase are in direct contact with each other allowing an extensive heat transfer, and the cyclone itself (Figure 3.2), where the raw meal is separated from the flue gas due to centrifugal forces.
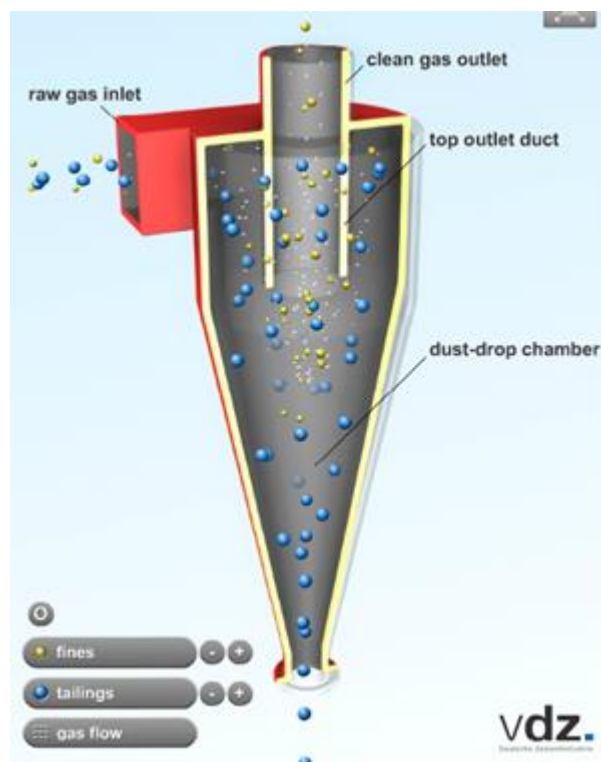
Figure 3.2: Scheme of a cyclone

Depending on the task (dust removal, low pressure drop) of the cyclone different designs are available. To ensure the function of the preheater, gas velocities of 15 to 20 m/s are necessary in the gas ducts connecting each stage of cyclone to the upper one. The preheater internal surfaces, both of connecting ducts and cyclones, are protected internally by a layer of refractory materials, in form of bricks or in special refractory concrete. Table 3.4 summarizes the reference parameters of the cyclone preheater.

Table 3.4: Preheater specification

Parameter	Unit	Value
Preheater cyclone stages	No.	5
Preheater strings	No.	1
Gas velocity	m/s	15.0
Entrance area, cyclone 1 (top of preheater)	m^2	8.25
Entrance area, cyclone 2	m^2	9.80
Entrance area, cyclone 3	m^2	10.95
Entrance area, cyclone 4	m^2	11.85
Entrance area, cyclone 5 (bottom of preheater)	m^2	11.85
Raw gas temperature (after preheater)	°C	314.0
Raw gas dust load (after preheater)	g/m^3_{STP}	61.0
Degree of calcination (meal entering the calciner)	%	18.0

3.4 Calciner

The calcination, which is the decomposition of calcium carbonate to calcium oxide and CO_2, requires around 2/3 of the burning process' energy demand. By applying a calciner the degree of decarbonisation of the raw material when entering the kiln could be increased to 90 - 95 %. Due to the required temperature level of around 900 °C especially low-calorific fuels could be used in the calciner (however, 100% coal scenario is envisaged in the reference cement kiln, see Chapter 3.2.2). Fuels can be added at different locations in the calciner and combusted flameless.

Calciners have to be fitted with a tertiary air duct (hot air from the cooler bypassing the kiln). In this way it is possible to burn a high quantity of fuel (up to 65% of the total plant demand). Moreover a variety of designs are available for different applications, such as low NO_x calciner, with pre-combustion chamber to use difficult fuels etc. In the reference case a simple in-line design with tertiary air duct was chosen, as shown in Figure 3.3.
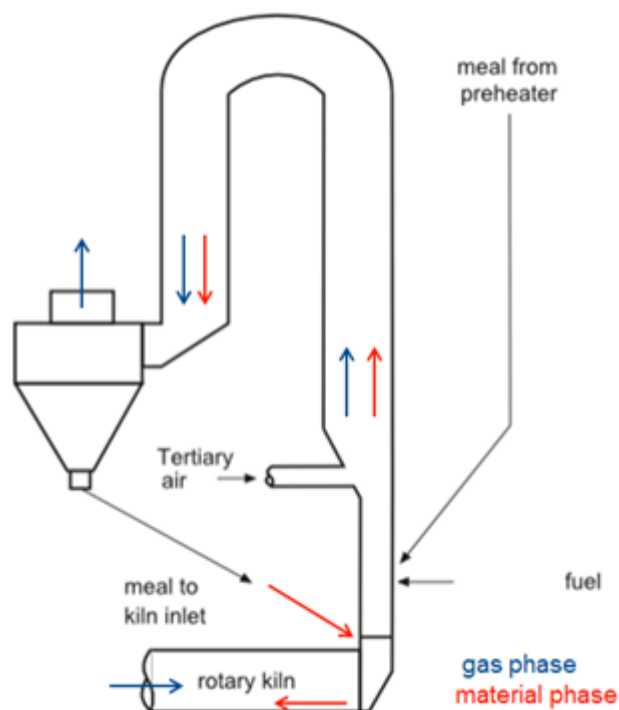
Figure 3.3: Scheme of an in-line calciner

In contrast to the other kiln units the gas and material phases are directed in co-current flow. The kiln exhaust gas entrains the hot meal from the preheater. To lift the material at least 15 - 20 m/s gas velocities are needed. In order to allow enough reaction time for calcination and the complete burnout of fuels (mainly if alternative) 3 - 6 seconds of residence time should be ensured.

The calciner is equipped with one or more burners, sometimes just fuel injectors, placed in different positions to control the temperature in the best way and prevent NO_x formation. The combustion takes place almost immediately, without the classical flame. The temperature is

stable around 900°C which is the decarbonation temperature, mainly due to the very fast heat absorption of the generated heat by the decarbonation reaction. Also the calciner is internally protected with a brick lining.

Table 3.5 summarizes the reference parameters of the calciner.

Table 3.5: Calciner specification

Parameter	Unit	Value
Calciner length	m	45.0
Calciner inner diameter	m	3.88
Gas velocity	m/s	15.0
Residence time	s	3.0
Degree of calcination (meal entering the kiln)	%	95.0
Material load (at inlet)	kg/m$^3_{STP}$	2.05
Coal mass flow	kg/h	8,700
Energy input	kJ/s	65,160
Tertiary air duct length	m	57.0
Tertiary air duct diameter	m	2.2
Gas velocity	m/s	26.0 – 27.0

3.5 Kiln

The completion of calcination, the formation of the clinker phases and the granulation of the kiln charge take place in the rotary kiln. The rotary kiln represents therefore the core of a cement work. Rotary kilns are steel tubes, placed on 2 or 3 roller stations, inclined between 3% and 4% towards the discharge end, rotating at a rate of about 1.3 to 3.5 revolutions per minute. The length of the kiln depends on production capacity and the extent of calcination of the raw meal entering the rotary kiln. Modern rotary kilns with preheater and calciner are 50 m to 80 m long and have a diameter between three and seven metres. The inside of the rotary kiln is lined with refractory bricks as the high temperatures in the kiln (gas phase up to 2,000 °C, material 1,450 °C) would otherwise destroy the tube.

Depending on the length of the kiln the gas residence time is 2 to 4 s at temperatures greater than 1,200 °C. The solid material takes 20 to 40 minutes to pass through the kiln depending on the degree of calcination and the size of the kiln. During its way through the kiln the raw material components form the mineralogical phases via intermediate phases. Figure 3.4 shows the intermediate phases along the reference rotary kiln.

Table 3.6 summarizes the reference parameters of the rotary kiln.
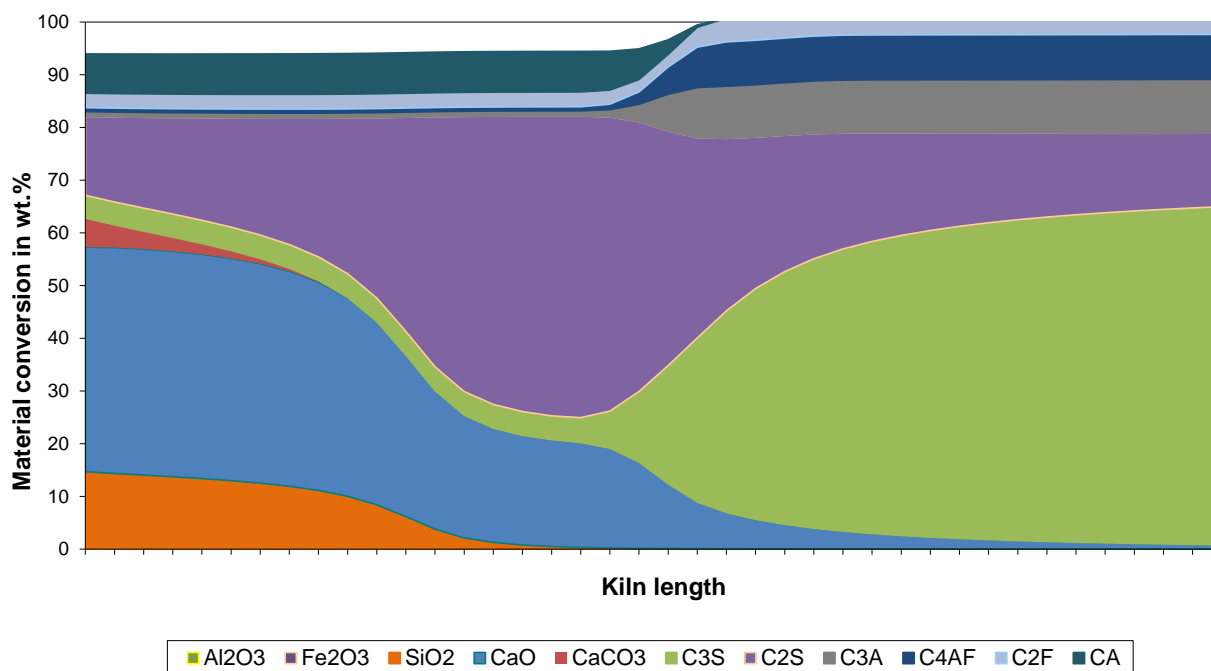
Figure 3.4: Material conversion of the main components in the reference rotary kiln (material flow from left to the right, gas flow from right to left). For the explanation of the abbreviated formulae of clinker phases see Chapter 3.2.1. The x-axis kiln length is given in arbitrary units.

Table 3.6: Kiln specification

Parameter	Unit	Value
Kiln length	m	57.0
Kiln diameter (outer)	m	4.04
L/D ratio	-	14
Pre-cooling zone	m	3.0
Thickness of the refractory	m	0.2
Volume load	t/(d*m^3)	5.2
Thermal load	MJ/(s*m^2)	3.9

3.6 Main burner

The burner placed at the rotary kiln discharge end is usually called main burner because in the past all or the major part of the fuel was burned in this part of the process. As mentioned above, in modern kilns with calciners only around 40 % of the total thermal input is generated here. It is important to note that fuels used during the clinker production process are experiencing a full utilization concerning the released thermal energy and introduced material. To achieve this, the firing systems in cement kilns are properly designed in order to handle various kinds of fuels. The general design of a kiln burner is shown in Figure 3.5.

1 swirl air channel 4 exit flow system, pulverized coal
2 exit flow system, swirl air 5 axial air channel
3 pulverized coal channel 6 exit flow system, axial air

Figure 3.5: General design of a cement kiln burner

A portion of the combustion air, called primary air, is used in part to transport the ground coal to the very end of the burner (the tip of the burner) and in part to control the shape of the flame (divided in axial air and swirl air). This portion of combustion air is not preheated and it is necessary to minimize its amount, both because it is better to recover as much preheated air from the cooler as possible, and because NO_x formation is favoured by a high amount of primary air. The major portion of combustion air is retrieved from the clinker cooler as preheated secondary air. This secondary air is fed from the outside of the burner through the clinker cooler and then mixed into the flame.

Flame temperatures in the range of 1,800 - 2,000 °C are required in order to sinter and granulate the kiln charge and to form clinker phases at kiln charge temperatures higher than 1,450 °C (Figure 3.6). In order to create suitable temperature profiles a compact, hot and not too long flame should be formed. The flame shape and the combustion are controlled by combining the axial momentum flow, the rotation and the divergence of the primary air. An optimum heat transfer in the kiln by radiation is shown in Figure 3.7.

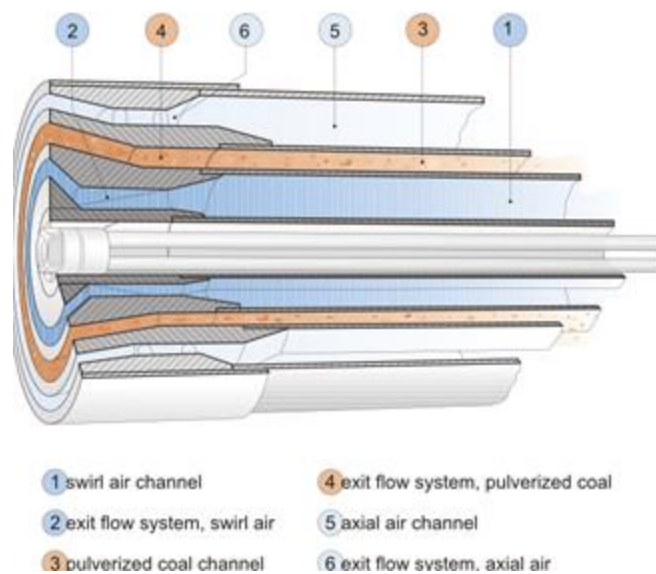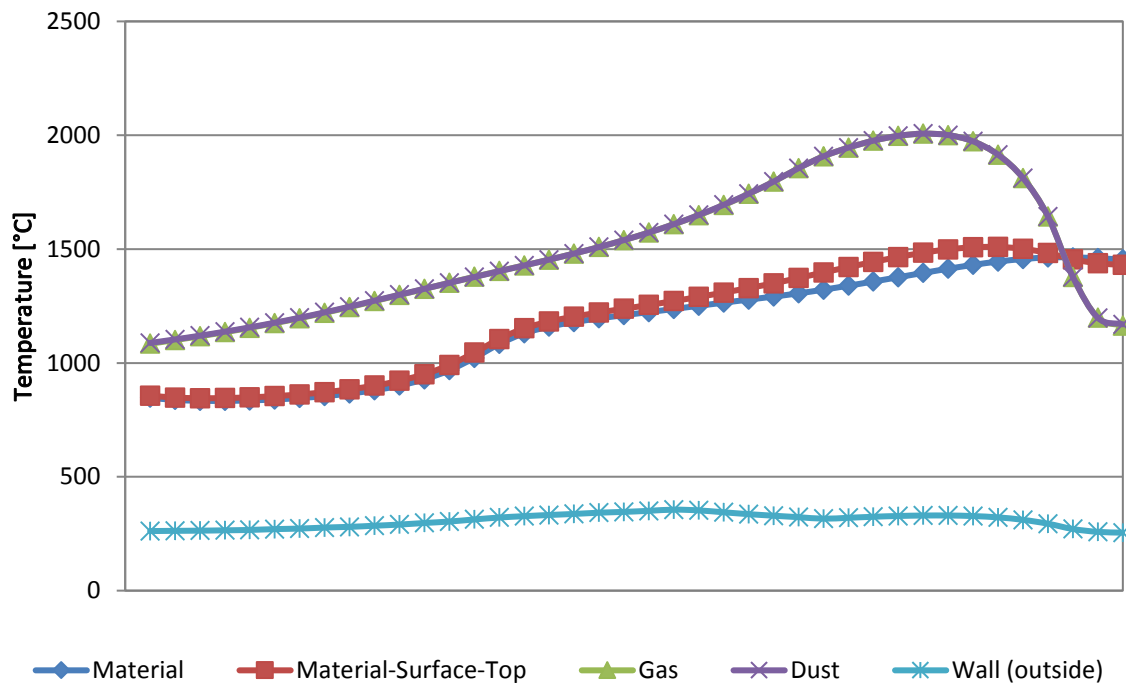Technical specifications of the burner in the reference kiln are given in Table 3.7.

Figure 3.6: Temperature profile in the reference rotary kiln (material flow from left to the right, gas flow from right to left); x-axis: kiln length in arbitrary units
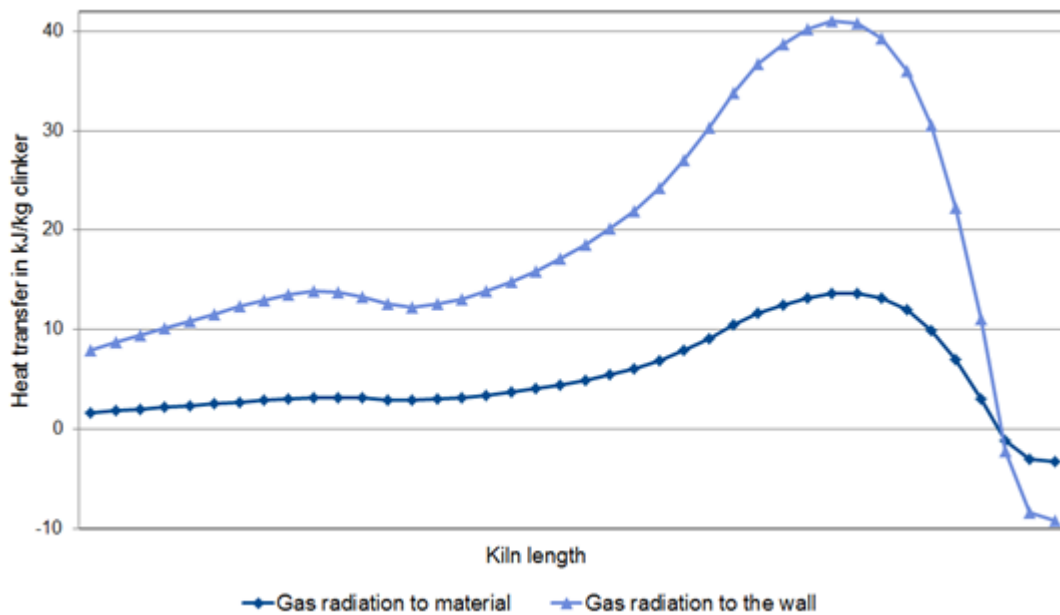

Figure 3.7: Radiation heat in the reference rotary kiln (material flow from left to the right, gas flow from right to left); x-axis: kiln length in arbitrary units

Table 3.7: Burner specification[1]

Parameter	Unit	Value	Typical Value
Primary air volume	% of combustion air	8 -15	8
Primary air temperature	°C	30 – 50[2]	50
Solid fuel injection velocity	m/s	15 – 40	30
Fuel to air ratio in transport air	kg/kg	2 – 9	5
Primary air velocity at nozzle exit	m/s	150 – 250	200
Secondary air velocity (at burner tip)	m/s	5 -10 (max. 15)	5
Coal mass flow [3]	kg/h	5,290	
Energy input	kJ/s	39,895	

[1] The values indicated are not set in the VDZ process model, because the VDZ model does not include the burner settings.

[2] Temperature increase from ambient due to the compression in the fan.

[3] Calorific value: see Table 3-3

3.7 Clinker cooler

Modern plants are mainly equipped with grate clinker coolers. Despite having very different designs (clinker transport, shape of grate plates, air feed), all grate coolers operate by the same cooling principle: once discharged from the kiln the clinker lies on a grate and cooling air flows through it from below, according to a cross flow heat exchange mechanism. The cooler generates the secondary combustion air, which flows through the kiln hood to the rotary kiln, and the tertiary combustion air, which flows through a connection located on the hot part of the cooler or in the kiln hood and then along the tertiary air duct up to the calciner. It is possible to generate different air flow rates in the different temperature zones of the cooler by using several fans underneath the grate (Figure 3.8). However, this optimised clinker cooling process means that grate coolers have an increased specific air requirement, so the excess air must be discharged as cooler exhaust air or as central exhaust air and dedusted. However, due to its temperature of up to 400 °C this excess air can be used for drying and energy recovery. Table 3.8 summarizes the reference parameters of the grate cooler.
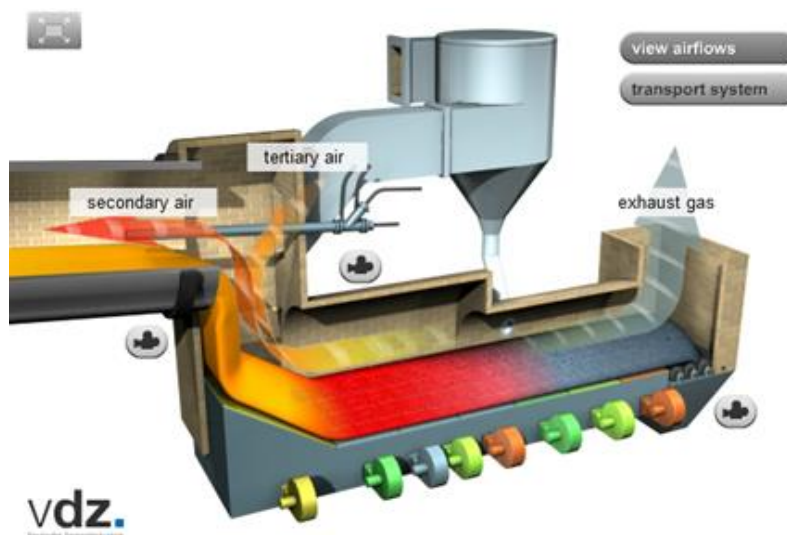

Figure 3.8: Scheme of a grate clinker cooler

Table 3.8: Cooler specification

Parameter	Unit	Value
Aerated length	m	22.0
Aerated width	m	3.5
Area load	$t/(d*m^2)$	39.0
Specific cooling air volume	m^3_{STP}/kg_{clk}	1.84
Specific exhaust air volume	m^3_{STP}/kg_{clk}	1.03
Exhaust air temperature	°C	285
Specific secondary air volume	m^3_{STP}/kg_{clk}	0.23
Secondary air temperature	°C	1,137
Specific tertiary air volume	m^3_{STP}/kg_{clk}	0.58
Tertiary air temperature	°C	1,086
Cooled clinker	kg/h	125,000
Cooled clinker temperature	°C	115

3.8 Flue gas

Flue gas is produced by burning fuel in the rotary kiln and calciner, and by calcination of the raw meal in the calciner. The flue gas conditions (e.g. flow rate and CO_2 concentration) are in addition to type of kiln and overall plant layout dependent on the amount of air leaking into the system and on the mode of operation of the raw mill. In CEMCAP we consider two main cases:

- Increasing air leak
- Constant low air leak

These cases are described in detail in Chapter 3.8.4. In the preceding chapters the impact of air leak and mode of operation is explained, and flue gas conditions for different air leak scenarios and modes of operation are defined.

3.8.1 Interconnected and direct mode

A cement kiln switches between so-called *interconnected* and *direct* mode of operation during the daily operation, and this has an impact on the resulting flue gas characteristics. The route of the flue gas through the process in *interconnected* mode is shown with grey solid arrows in Figure 3.9. After the preheater the flue gas is sent through a conditioning tower where the temperature and moisture content of the gas can be controlled by spraying water into the gas. However, in interconnected mode the amount of water spraying required in the conditioning tower is small or zero. When the water spraying is zero the conditioning tower is operated like a simple ductwork. The flue gas is thereafter sent to the raw mill where the residual thermal energy associated to the flue gas is used to dry the raw mix. After the raw mill the mix of flue gas and ground raw mix is sent to the filter where the solids are separated from the gas before the gas is sent to the stack, while the solids are sent to the top stage of the preheater via homogenization silos and raw meal silos. In this mode the kiln and the raw mill are operating in a completely interconnected way, as a single machine. A cement kiln is run in interconnected mode typically 90% of the time during a day (21-22 h/d).

The raw mill can be stopped almost every day for visual inspection and control. When the mill is out of operation the kiln is run in *direct* mode. The flue gas then bypasses the raw mill and is sent directly to the dust filter as shown with dashed arrow in Figure 3.9. This mode of operation is called direct mode because of the direct connection between the kiln and the process filter. While the mill is stopped, the flue gas coming from the preheater and passing through the conditioning tower has to be cooled down to a temperature compatible to the one tolerated by the material of the bag filter, or increased in moisture content up to the point which is necessary for the highest efficiency of the electrostatic precipitators. In the reference case we assume the use of a high temperature bag filter that can be operated at temperatures up to 210 °C. The kiln is typically run in direct mode 10% of the time during a day.
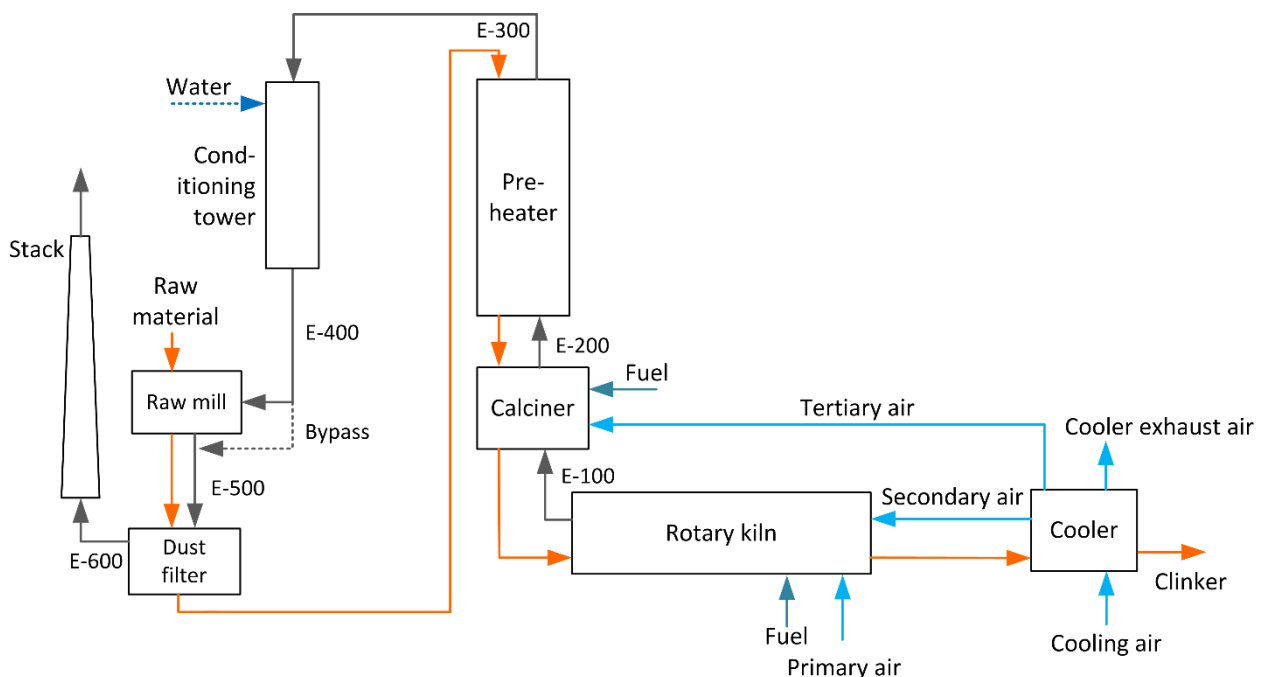
Figure 3.9: Flowsheet of cement kiln with numbered flue gas streams for interconnected operation. Dashed arrows shows streams relevant for direct operation. The lower cyclone in the preheating tower is treated as a part of the calciner.

3.8.2   Air leak

At several points along the process the flue gas is diluted by air leaking into the system – typically at the kiln inlet and outlet, in the calciner, in the preheater stages, and in the raw mill. Air that has leaked into the flue gas is referred to as "false air". The extent of air leak in the raw mill at interconnected operation is relatively high, while there is usually not much air leak after the preheater at direct operation.

The amount of air leak also changes a lot during the classical year of operation. Air leak increase over the year and is reduced during maintenance periods. Relevant variations over the year are:

- Air leakage *at seals* between rotating and non-rotating parts is typically doubled during the year. In a BAT plant the amount of false air in the flue gas at the preheater outlet can

be around 6% on dry standard volume basis right after maintenance. After the air leak is doubled, the false air content has increased to 10%.

- Air leakage *in the raw mill* can also double during the year. The amount of false air at the stack can be 30% just after maintenance. Accounting for doubled air leak both at seals and in the mill the resulting amount of false air at the stack is 46% at the end of the year.

3.8.3   Flue gas conditions

Gas flow rates, temperatures and compositions are available from the VDZ process model for the gas leaving the rotary kiln (E-100), the calciner (E-200), and the preheater (E-300). It should be noted that here the lower cyclone in the preheater tower is treated as a part of the calciner. These data are summarised in Table 3.9. The content of impurities can be back-calculated from Table 3.12.

Table 3.9: Stream data for gas streams E-100 to E-300 from the VDZ process model

	Unit	E-100	E-200	E-300
Total flow rate	kg/h	61,411	226,504	236,470
Temperature	°C	1,078	860	314
Gas composition, wet basis				
CO_2	vol%$_{STP}$	20.1	33.2	32.0
N_2	vol%$_{STP}$	70.5	59.2	58.8
O_2	vol%$_{STP}$	1.5	2.8	3.0
H_2O	vol%$_{STP}$	6.4	4.9	6.2
SO_x (expr. as SO_2)	mg/m$^3_{STP}$	2,739	499	236
Dust	g/m$^3_{STP}$			61

In this part of the plant the effect of increased air leak over the year is relatively small. By doubling the air leak, exhaust gas flow rate increases by maximum 5%, and gas compositions change with less than 2 pp. The effect of varying air leak is therefore neglected in the reference plant in these points.

When it comes to the flue gas at the stack, the variations are higher, both due to switching between interconnected and direct operation, and due to varying degrees of air leak. Ranges for gas composition and dust content of the flue gas at the stack (E-600) based on measurements from a wide range of rotary kiln systems (not assigned to a specific fuel scenario) are available from a standard document from the Association of German Engineers [VDI, 2003] and provided in Table 3.10.

Table 3.10: Measured gas phase concentrations (with nitrogen as balance) and dust content (STP, dry basis) at the stack (E-600) for a range of different rotary kiln systems [VDI, 2003].

Exhaust gas component	Unit	Range
Gas phase composition, dry basis		
CO_2	vol%$_{STP}$	14 - 35
O_2	vol%$_{STP}$	5 - 14
CO	g/m$^3_{STP}$	0.1 - 2
NO and NO_2, expressed as NO_2	g/m$^3_{STP}$	0.5 - 0.8
SO_2 and SO_3, expressed as SO_2	mg/m$^3_{STP}$	<50 - 400
Organics, expressed as total carbon	mg/m$^3_{STP}$	5 - 100
Gaseous inorganic chlorine compounds, expressed as HCl	mg/m$^3_{STP}$	<20
Gaseous inorganic fluorine compounds, expressed as HF	mg/m$^3_{STP}$	<1
Dioxins/furans	ng I-TEQ/m$^3_{STP}$	<0.1
Dust	mg/m$^3_{STP}$	<5 - 30

For the reference cement kiln, flue gas conditions for interconnected operation with medium air leak and low air leak, and for direct operation, are defined using the following assumptions:

- *Interconnected operation with medium air leak:* After the preheater outlet air leak of 139,800 kg/h is assumed in the mill which is *double* the amount in a BAT reference plant described by Müller [MÜL, 1993]. The raw material is dried from 6 to 1 wt% moisture and heated from 15 °C to 60 °C. The raw mill is in operation 90% of the day, giving a raw material flow rate 1.1 times higher here than at the preheater inlet. This corresponds to evaporation of 11,822 kg/h of water in the raw mill, and the flue gas is cooled to 110 °C.

- *Interconnected operation with low air leak:* This case is similar to the "medium air leak" case, but the air leak is 69,900 kg/h which is *the same* as in a BAT reference plant described by Müller [MÜL,1993].

- *Direct operation:* In this case the air leak after the preheater is assumed to be zero. The flue gas is cooled to 210 °C (we assume use of high temperature bag filters). This requires spraying of 9,700 kg/h water in the conditioning tower.

The cost of CO_2 capture increases as flue gas flow rate increases and CO_2 concentration decreases. It is therefore likely that there will be more maintenance over the year and focus on limiting avoidable air leak at a cement kiln where CCS is applied. In reality there are also plants with far more air leak than defined here, but it is not likely that these plants will be selected first for application of CCS.

The resulting flue gas properties at the stack (E-600) are summarised in Table 3.11. The dust content at the stack is subject to environmental regulations, and is assumed to be 10 mg/Sm$^3_{STP}$ in all cases. Impurity concentrations related to a state with 10% O_2 are assumed to be the same in all cases, and are given in Table 3.12. See Chapter 2.2 on how to relate the values in the table to the cases with O_2 content other than 10%.

The concentration of NO_x is limited to 0.5 g/m^3_{STP} by a SNCR system. The NO_x reduction is achieved by injection of ammonia (reducing agent) into the flue gas stream, according to the following chemical reaction: $4\ NO + 4\ NH_3 + O_2 \rightarrow 4\ N_2 + 6\ H_2O$.

Table 3.11: Stream data for the exhaust gas at stack (E-600) for the three cases considered in CEMCAP. Impurity contents are specified in Table 3.12.

Mode of operation	Unit	Interconnected		Direct
Air leak in mill		Medium	Low	-
Air leak in mill, flow rate	kg/h	139,806	69,903	-
Total flow rate	kg/h	388,098	318,192	246,170
Temperature	°C	110	130	210
Gas phase composition, dry basis				
$\quad CO_2$	vol%	20	25	34
$\quad N_2$	vol%	69	67	63
$\quad O_2$	vol%	11	8	3
Gas phase composition, wet basis				
$\quad CO_2$	vol%	18	22	29
$\quad N_2$	vol%	63	60	54
$\quad O_2$	vol%	10	7	3
$\quad H_2O$	vol%	9	11	13
Dust	mg/m^3_{STP}	10	10	10

Table 3.12: Impurity concentrations at 10% O_2 content.

Component	Unit	Value
CO	g/m^3	1
NO and NO_2, expressed as NO_2	g/m^3_{STP}	0.5
SO_2 and SO_3, expressed as SO_2	mg/m^3_{STP}	200
Organics, expressed as total carbon	mg/m^3_{STP}	50
Gaseous inorganic chlorine compounds, expressed as HCl	mg/m^3_{STP}	5
Gaseous inorganic fluorine compounds, expressed as HF	mg/m^3_{STP}	0.5
Dioxins/furans	ng I-TEQ/m^3_{STP}	0.05

3.8.4 Flue gas base case definition

The flue gas conditions (e.g. flow rate and CO_2 concentration) are significantly affected by the air leaking into the system. Normally the air leak at the cement kiln increases over the year, while it is decreased by maintenance. If we assume more maintenance over the year it should be possible to keep the air leak low.

We therefore consider two cases for air leak over the year:
- Increasing air leak (base case)
- Constant low air leak

In the case with increasing air leak it is considered low air leak the first ½ of the year, and medium air leak the second ½ of the year. In this case the capture system should be designed for medium air leak, which has a higher flow rate than the process with low air leak. The performance of the system should be calculated for both flue gas conditions, and average values for power consumption, steam consumption *etc.* should be used to calculate KPIs.

In the case with constant low air leak it is considered low air leak the whole year.

In both cases there will be direct operation (mill out of operation) 0-2 hours every day. The period with direct operation will most likely give a small reduction in the average heat and power consumption. This will however not be taken into account in the calculation of KPIs.

4 CAPTURE TECHNOLOGIES

In this chapter the principles of the relevant capture technologies are explained. For each technology a list of the relevant process units is given. They are either defined as case specific units, which are units where specifications must be set individually for the different technologies (e.g. absorption and desorption columns), or as general process units, where specifications are set commonly in Chapter 5 (e.g. heat exchangers, pumps and compressors).

4.1 Reference case: MEA absorption

Reactive absorption with aqueous amine solutions as solvent is considered as the most mature option for CO_2 capture. Absorption with amine is the technology used at the first full scale CO_2 capture power plant in Boundary Dam in Canada [BOU, 2013]. Absorption with aqueous solution of monoethanolamine (MEA) is established as reference technology in benchmark studies of CO_2 capture processes, and will be used as the base case also in CEMCAP.

The principle of the technology is shown in Figure 4.1. The solvent (30 wt% MEA) is circulated between an absorber and a desorber. Flue gas is sent through the absorber, where the CO_2 reacts with MEA and is dissolved in the solvent. The CO_2 rich solvent is heated and sent to the desorber. In the desorber the solvent is regenerated by further heating with a reboiler operated around 120 °C. The reaction is reversed, and CO_2 is released from the solvent. The desorber is operated at a pressure of 1-2 bars, which means that the purified CO_2 is produced at this pressure. It is sent to further conditioning for transport and storage/reuse by compression or liquefaction. The CO_2 lean solvent is recirculated back to the absorber. The process units required in the process are listed in Table 4.1.

MEA can separate CO_2 at very low partial pressures, and the CO_2 is produced at high purity. However, it requires high thermal energy for regeneration, it is corrosive, and the amine degrades over time [LIN, 2014].
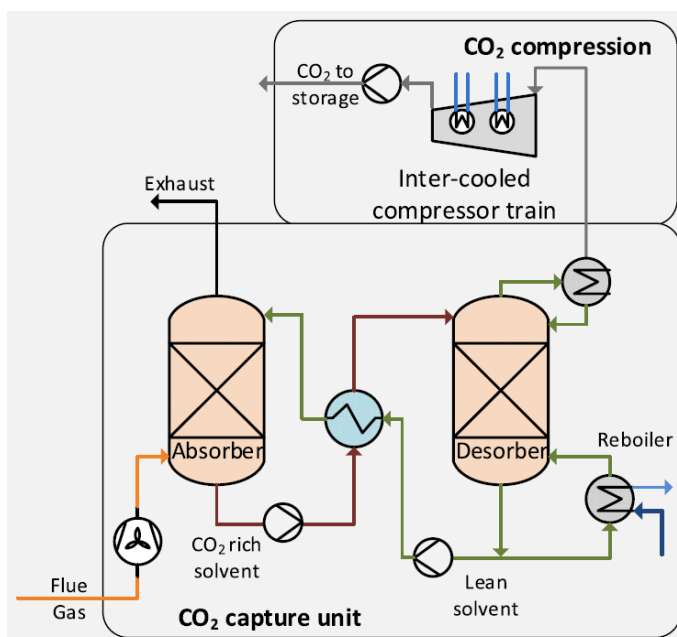

Figure 4.1: Principle of MEA absorption

Table 4.1: List of process units required for MEA absorption

Case specific units	General units
Absorber	Fan
Desorber	Liquid/liquid heat exchangers
	Solvent pump
	CO_2 pump

4.2   Oxyfuel

The oxyfuel technology relies on the combustion with pure oxygen and a recirculation of flue gas in order to enrich CO_2 to an amount which allows a relatively easy purification by liquefaction systems. For this purpose different integration systems can be chosen, the full and the partial oxyfuel technology.

Implementing the <u>full oxyfuel concept</u> almost all generated CO_2 can theoretically be captured. In this case the whole plant is operated under oxyfuel conditions. Therefore, all plant units are influenced by the changed gas atmosphere. The heat transfer, the combustion, the capacity streams of material and gas as well as the clinker formation are affected due to the different gas properties like heat capacity, emissivity or density.

Within a joint research project of the European Cement Research Academy (ECRA) a full concept of an oxyfuel cement kiln has been developed [ECR, 2009]. The principal configuration of this design uses the conventional technology as the starting point (see Figure 4.2).
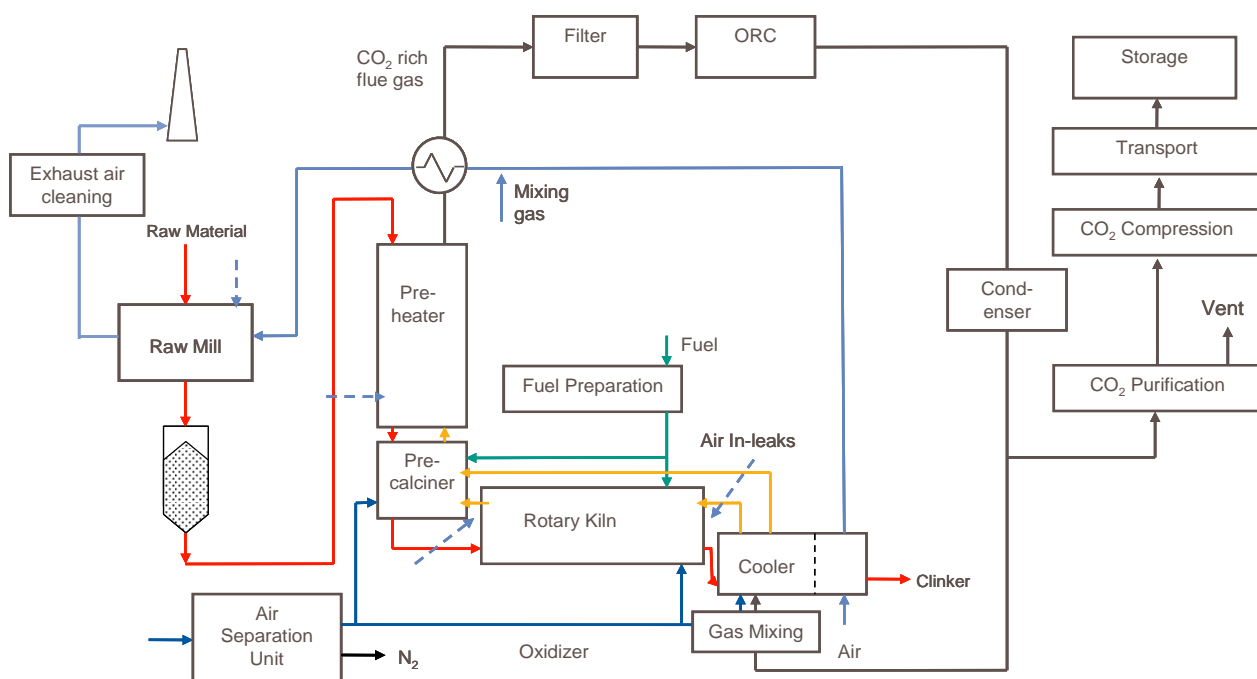

Figure 4.2: Configuration of a full oxyfuel cement kiln [ECR, 2009]

The main additional installations required for the oxyfuel kiln are:

- two stage clinker cooler (first stage operated in oxyfuel mode, the second one in air mode)
- exhaust gas recirculation system
- gas-gas heat exchanger (optionally, a gas-steam heat exchanger)
- condensing unit
- air separation unit (ASU)
- CO_2 purification (by liquefaction) unit (CPU)
- rotary kiln burner for oxy-combustion.

In Table 4.2 these units are sorted as case specific units and general units.

Clinker cooler: The first cooler stage is operated with recycled flue gas, which is needed in the burning process. As this would result in still too high clinker temperatures, a second cooler stage, which is operated with ambient air, is considered. The air leaves the cooler as exhaust air and can be used for raw material drying or fuel preparation. A major advantage of using cooler exhaust air for drying purposes is the fact that e.g. the raw mill must not be operated under air-tight conditions.

ASU: The oxygen from the ASU mixed with the recirculated CO_2 rich exhaust gas is forming the so-called oxidizer. This is provided to the precalciner and kiln firing (as primary "air") as well as to the premixing of cooling gas. For a medium-size cement kiln with a capacity of 3,000 tpd the oxygen demand is estimated to be around 30 to 35 tph. Such amounts of oxygen can for logistical reasons only be provided by an on-site oxygen supply system.

Recirculation/heat exchanger/condenser: Within the recirculation the flue gas undergoes different steps like the removal of heat, dedusting and dehydration. Part of the flue gas is discharged to the CO_2 purification unit (CPU) and the residual fraction to the cooler for another cycle. Heat from the flue gas, which is leaving the preheater, could also be used to increase the drying potential of the cooler exhaust air by a gas-gas heat exchanger. If the flue gas still contains enough energy, power can be produced, e.g. by an Organic Rankine Cycle (ORC).

The partial oxyfuel concept concentrates the oxyfuel operation only on the calciner, which is separated from the kiln units of the plant. In the case of a double line preheater tower one line could also be switched to oxyfuel operation. This concept takes advantage of the fact that most of the CO_2 emissions are generated in the calciner by a major part of the decomposition of carbonates (responsible for approx. 60 % of CO_2 from cement plants) and fuel input (ca. 60 % of total fuel input). As the other installations (kiln, cooler, raw mill) are operated conventionally, this option avoids the increased effort involved with the improvement of seals and does not have any impacts on the product quality. Due to fewer changes to the kiln plant design and reduced influence on the plants operation this concept is seen preferably for retrofitting purposes. In this layout the gas supplied to the calciner possesses lower temperatures compared to conventional operation, which also affects the overall plant energy efficiency. Higher energy input to the main burner leads to unabated CO_2 . This circumstance and losses by the CPU lead to an overall capture rate of this technology of 60 %. Therefore the capture efficiency is lower compared to full oxyfuel operation of the clinker burning process (> 85 %).
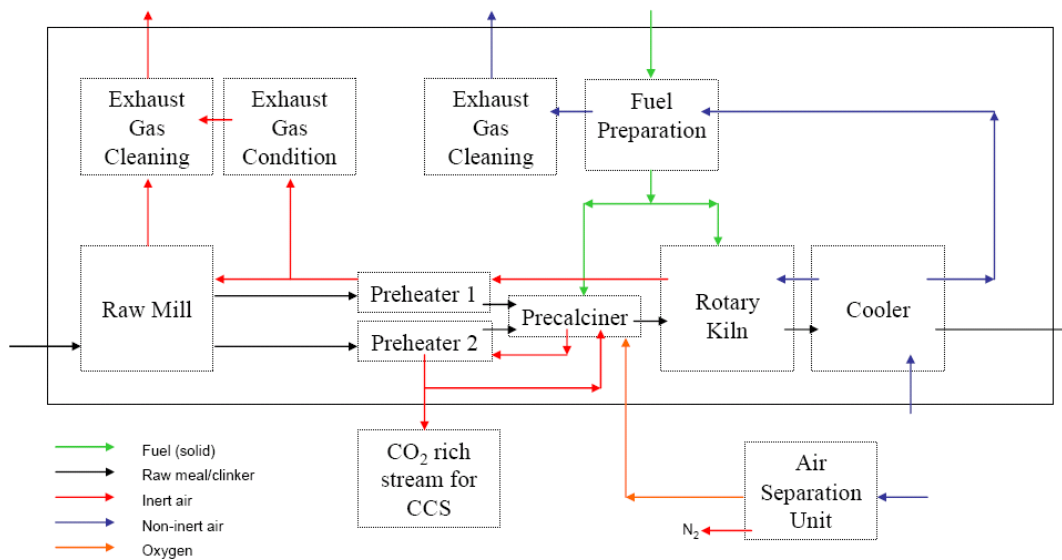
Figure 4.3: Configuration of a partial oxyfuel cement kiln [IEA, 2008]

Exemplary for the partial oxyfuel technology the configuration of [IEA-08] is explained more in detail in the following (see Figure 4.3). Oxygen from the air separation unit is supplied premixed with recycled flue gas before being provided to the calciner. Here the preheated material from both preheater strings is calcined and then supplied to the kiln for further mineralogical conversion. Gas from the calciner, which is enriched by CO_2 from the material and combustion gases, is provided to preheater string 2. Preheater string 1 is operated with combustion gases from the rotary kiln. These combustion gases can be used for the drying of raw material. After this usage it is released to the environment, while the oxyfuel string gases are captured. As tertiary air from the grate cooler is not provided to the calciner as usual, it can be used for other issues like preheating, drying or power generation.

Table 4.2: Case-specific and general units for the oxyfuel process

Case specific units	General units
Oxyfuel burner	ASU
Two-stage cooler	ORC
Heat exchanger	CPU
Exhaust gas recirculation system	
Condenser unit	

4.3 Chilled ammonia

The Chilled Ammonia Process (CAP) follows the same technology principle of MEA absorption, previously introduced in Chapter 4.1. CAP is a promising technology for post-combustion CO_2 capture and shows competitive energetic performance. The use of aqueous NH_3 as a solvent offers advantages concerning global availability, environmental footprint, cost and chemical stability in the presence of impurities such as SO_x and NO_x.

The overall plant layout, shown in Figure 4.4 is similar to most of the amine-based scrubbing processes and is subdivided in three main sections: (i) the flue gas cooling section, (ii) the CO_2 capture section, and (iii) the ammonia slip abatement section (generally called NH_3 water wash). Following Figure 4.4, the untreated flue gases exiting the Flue Gas Desulfurizer (FGD) are cooled by means of a direct contact cooler which limits the temperature increase within the CO_2-absorber. The cooled flue gases are then sent to the absorber, where the CO_2 content is reduced by the ammonia solution. The solution is regenerated in the CO_2 desorber by heating up the solution with steam in a conventional kettle reboiler at about 120 - 130 °C. High purity CO_2 leaves the column condenser with pressure up to 20 bar and is further conditioned by compression or liquefaction. The decarbonized flue gases are sent to the ammonia control section, where the NH_3 slip is reduced using chilled water. Eventually, the recovered ammonia is stripped in a dedicated desorber and is re-used in the CO_2 capture section.

CAP has been developed and demonstrated at pilot scale by Alstom [GAL, 2006], [GAL, 2008], [GAL, 2011] in a range of CO_2 concentration between 3 and 16 % [TEL, 2011], [KOZ, 2009]. The increased CO_2 concentration in the flue gas typical of cement kilns affects the operation of the entire process. The higher concentration in the flue gas can be compensated with a higher liquid to gas flow rate ratio, with a higher NH_3-concentration of the solvent, with a higher CO_2 loading of the solvent, or with a combination of these. Therefore, a significant change in the flue gas composition requires fundamental modifications and a re-optimization of the existing CAP design.
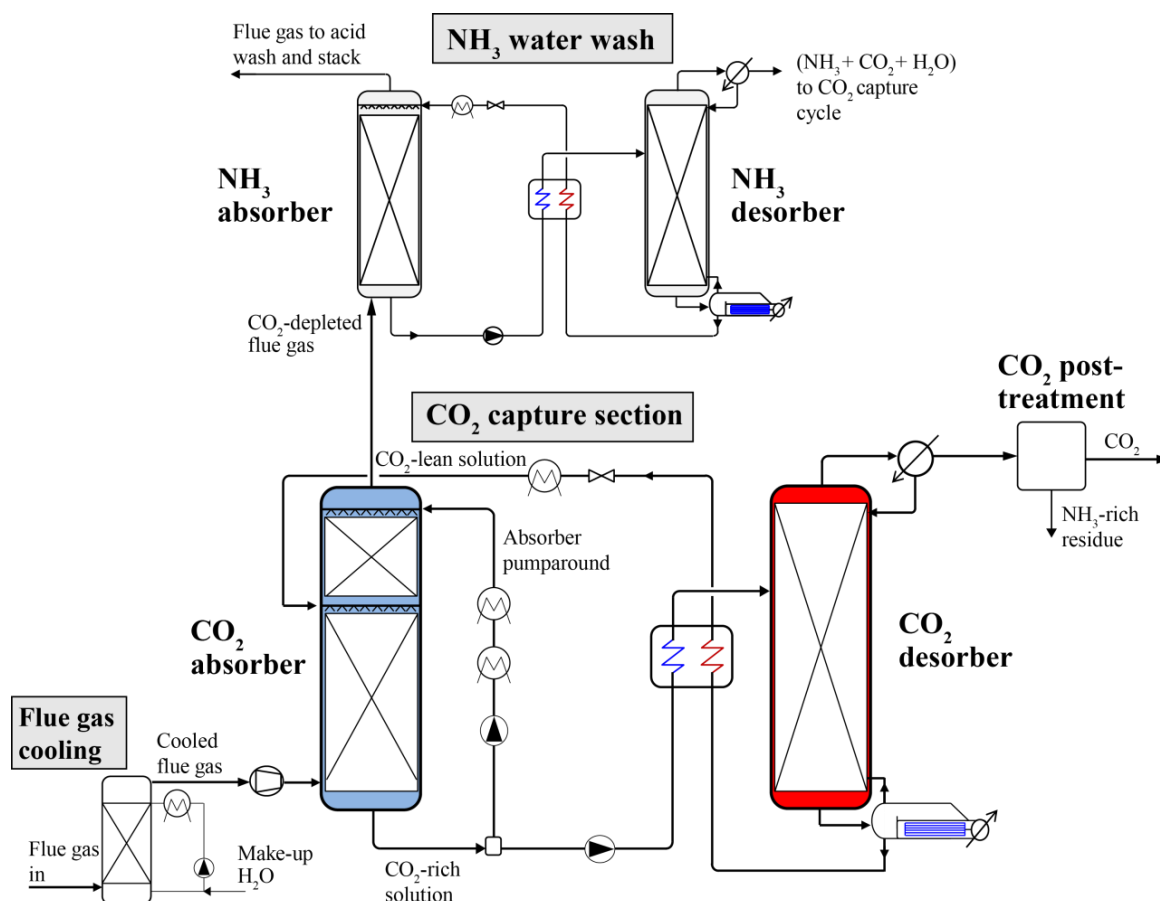
Figure 4.4: Chilled Ammonia Process (CAP) layout

The CAP plant operating conditions strongly affect the chemical behavior of the process: solid phases, primarily consisting of ammonium carbonate and bicarbonate, may form in the absorber and in related units. In the existing CAP plants the solid formation has been strictly avoided due to the complexity of handling solids and the operational issues related to the clogging of equipment. Nevertheless, solid formation offers different advantages: i) reduction of stripper heat requirement; ii) reduction of the stripper dimension; and iii) reduction of the ammonia slip from the absorber. Accordingly, the performance of the process can be improved when exploiting the formation of solids in a so-called 2nd generation CAP.

As shown in Figure 4.5, the plant layout of the 2nd generation CAP differs from the standard CAP in the CO_2 capture section. The rich solution leaving the absorber is introduced to a solid formation section where the solution is cooled and chilled in two subsequent crystallizer units. Solid phase forms at low temperature and the generated suspension is separated in a hydro-cyclone into (i) a rich slurry, which is sent to the regeneration, and (ii) a clear solution, which is recycled to the absorber. The absorber with its sophisticated structured packing, however, remains free of solids.
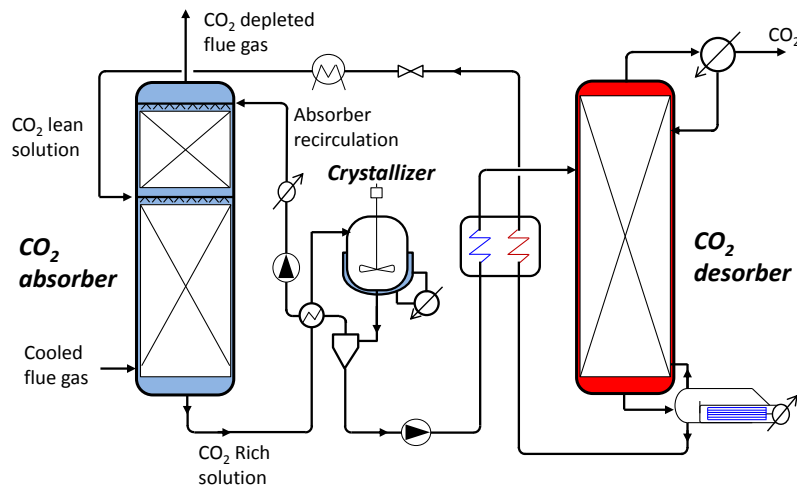
Figure 4.5: CO2 capture section layout of the 2nd generation CAP

A list of process units involved in this technology is given in Table 4.3, where case specific units are those which need to be specifically designed in CEMCAP while general specifications fixed at this stage of the project will be used for general units.

Table 4.3: List of process units required in CAP

Case specific units	General units
Direct contact cooler	Pump
CO_2 absorber	Heat Exchanger
CO_2 desorber	Flue gas blower
CO_2 post-treatment column	
NH_3 absorber	
NH_3 desorber	
Crystallizer*	
Hydrocyclone*	

* only required for 2nd generation CAP

4.4 Membrane assisted liquefaction

Membrane-assisted CO_2 liquefaction is based on the principle of combining two different separation technologies, none of which are perfectly suited for stand-alone capture of CO_2 at low to medium concentration in flue gases, so that each can carry out a partial separation within its favorable regime of operation. Polymeric membranes are generally suited for bulk separation of CO_2 resulting in moderate product purity. Low-temperature separation, on the other hand, is very well suited for purification of CO_2 of moderate-to-high feed purity through condensation of CO_2 and removal of the volatile components nitrogen, oxygen etc. by phase-separation.

A simplified process scheme for the combination of membrane bulk separation and low-temperature CO_2 purification is shown in Figure 4.6. An exemplified process configuration for the liquefaction unit is shown in Figure 4.7.
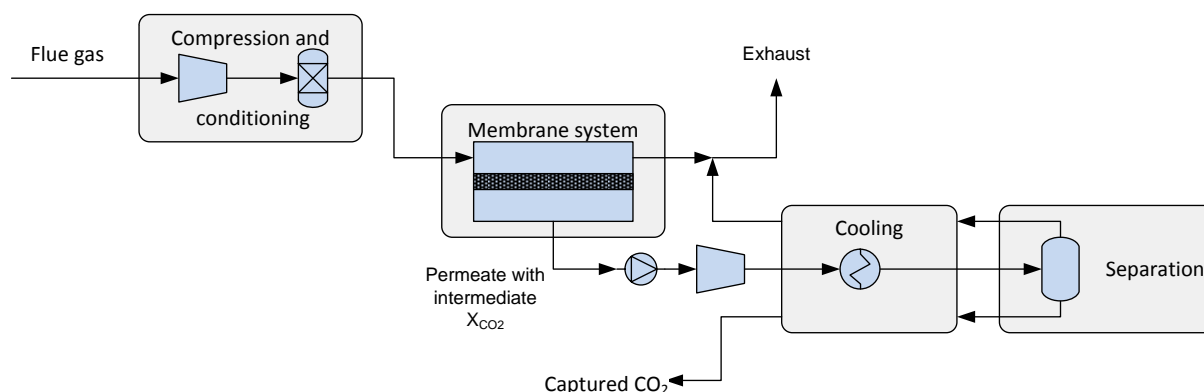
Figure 4.6: Simplified process scheme for membrane-assisted CO_2 liquefaction
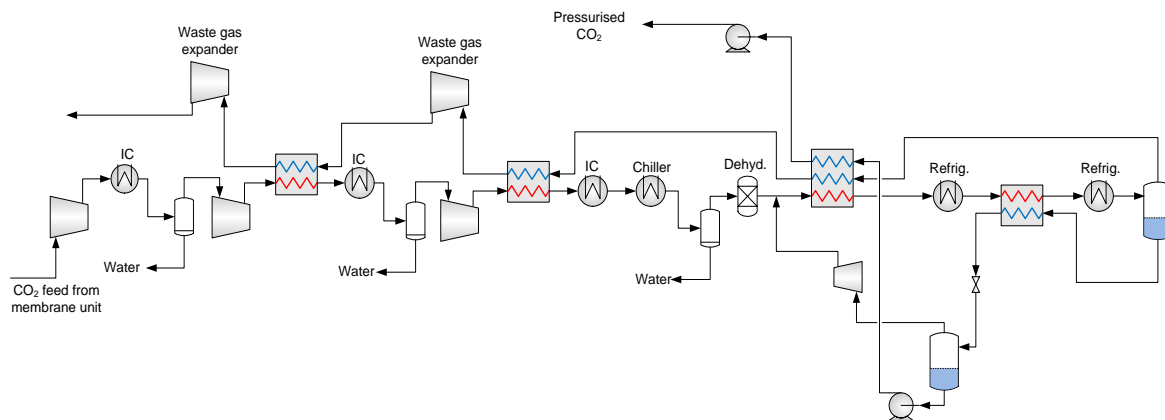
Figure 4.7: Example of possible process configuration for low temperature CO_2 condensation

Membrane separation is conceptually a very simple process where the driving force for separation is the component partial pressure differential across the membrane. The purity and capture ratio attainable from a single stage membrane process depends on the membrane properties such as membrane permeability and selectivity and involves a trade-off between driving forces across the membrane (attained by pressure manipulations of the feed and permeate streams) and the membrane area.

Efficient separation of CO_2 from volatile components, in this case primarily nitrogen and oxygen, by liquefaction requires compression and cooling of the feed stream. The obtainable CO_2 capture ratio is governed by the vapour–liquid equilibrium for the system. The CO_2 capture ratio strongly depends on the separation pressure (typically 20 - 40 bar) and temperature (typically around -55 °C). The energy- and/or cost-optimal CO_2 capture ratio is also highly dependent on the CO_2 concentration in the permeate gas from the membrane unit. The trade-off involved in the liquefaction process is thus between compression required prior to liquefaction and the refrigeration utility.

Thus, for a given CO_2 product purity and capture ratio of the overall process, the following membrane and process parameters will require to be "optimised":

- CO_2 capture ratio of the membrane process and liquefaction process
- Membrane properties (permeability and selectivity)
- Membrane permeate CO_2 purity
- Membrane feed pressure and permeate pressure
- Membrane area
- Compression level prior to liquefaction process[1]
- Refrigeration utility

---

[1] For a given capture ratio defined for the liquefaction process this is set by the condensation temperature

A significant advantage for the principle of membrane-assisted CO_2 liquefaction is that there is no requirement for process steam, which is normally not available in cement factories. The process will in principle depend on electric power only, which is readily available in most cases.

A list of process units required in the process is given in Table 4.4. For the membrane part several commercial membrane types exist, but it is likely that facilitated-transport membranes, currently under development, will become the preferred technology.

Table 4.4: List of process units required in membrane-assisted CO_2 liquefaction

Case specific units	General units
Membrane unit:	*Membrane unit:*
Membranes	Filters and gas cleaning units
	Fan or compressor
	(Vacuum pump)
	Liquefaction unit:
	Compressors and intercoolers
	Dehydration unit
	Heat exchangers
	Phase separators
	Gas expanders
	Liquid CO_2 pumps
	Industrial refrigeration units

4.5   Calcium looping

The Calcium Looping (CaL) process has been proven as an effective and cost-effective technology for post-combustion CO_2 capture in power plants [ARI, 2013]. It is based on the reversible reaction between a CaO-based sorbent and CO_2 in a flue gas, and in the subsequent calcination of the $CaCO_3$ formed in a separate reactor to regenerate the CaO and release the CO_2 as a concentrated gas stream. This technology has been developed in the last years from a concept paper to a pilot-scale of up to 1.7 MW_{th} in the framework of different European, national and industrial research projects, which have allowed carrying out progress at different levels: particle reaction fundamentals [ABA, 2003], [GRA, 2006] reactor testing and scale-up, [ARI, 2013], [ABA, 2004], [ALO, 2010], [CHA, 2011], [DIE, 2013], and process modeling [ROD, 2008], [ROD, 2012].

Main advantage of this technology is that it is a high temperature process (carbonator: ~ 650 °C; calciner: ~ 900 °C) so it allows recovering most of the heat introduced for sorbent regeneration as high temperature heat, which can be used for steam production. Due to its wide availability, cost and CO_2 carrying capacity, limestone is usually proposed as natural precursor of CaO. Coal is typically proposed to be burnt under oxy-combustion conditions in the calciner for supplying the energy needed for regenerating the sorbent. Consequently, a solid purge would be needed in the CaL process to avoid coal ashes and sulphur accumulation in the solid loop, as well as to eliminate deactivated CaO. This fact allows a synergy with cement industry by using purged material (mainly CaO) as raw material for clinker production.

In this project, the direct integration of a CaL system into a cement kiln is analysed, whose implications for cement and CaL process have only been theoretically analysed up to date [ROD, 2008], [ROM, 2013], [ROM, 2014]. The CaL configurations that are sufficiently mature to be demonstrated within CEMCAP are those involving oxy-combustion conditions in the calciner (represented in Figure 4.8 and Figure 4.9), and they are analysed in this project. Basically these CaL configurations differ on the type of reactors used for the carbonator and calciner. In both cases, heat recovery (not shown in the figures) should be performed from the walls of the carbonator and from cooling the carbonator and calciner off-gases. While heat from the carbonator will be recovered by producing steam, heat in the gas streams may be recovered both by preheating the raw meal in direct contact suspension preheaters or by feeding the heat recovery steam cycle. The optimal heat recovery strategy will depend on techno-economic parameters and will be object of the activity performed in CEMCAP.
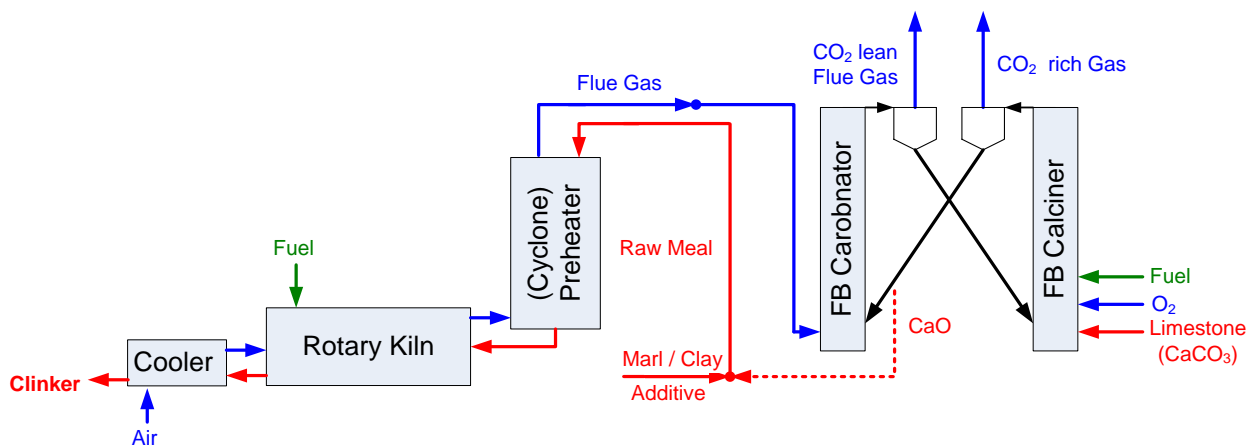

Figure 4.8: CaL process integration into a cement kiln operated in twin circulating fluidised bed reactors
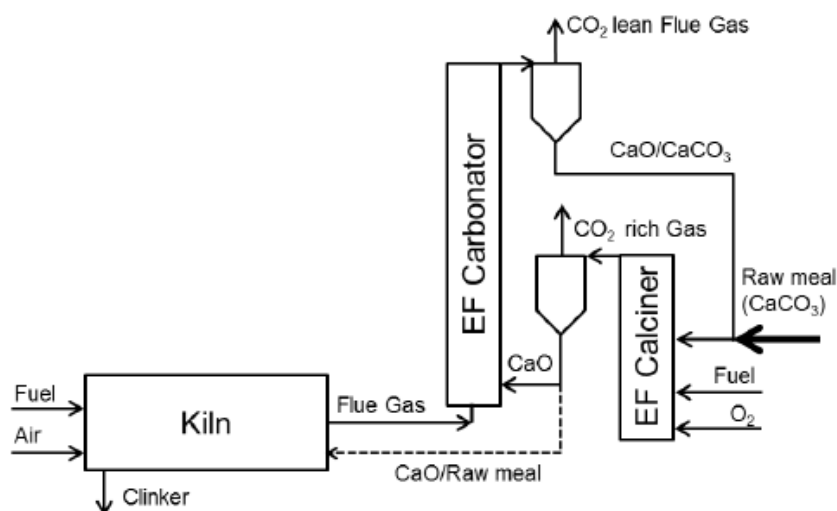
Figure 4.9: CaL process integration into a cement kiln operated in entrained-flow reactor technology

In the first analyzed configuration (Figure 4.8), carbonator and calciner reactors are two interconnected fluidised bed reactors. As mentioned before, this type of reactor technology has already been demonstrated at a pilot scale of 1.7 MW_{th} in post-combustion CO_2 capture applications in power plants. When integrated into a cement kiln, solids purged from calciner (dashed line in Figure 4.8), which are basically composed of CaO, are incorporated into the cement preheater for partially substituting the raw meal needed for clinker production. Doing so, CO_2 emissions associated with raw meal calcination are consequently reduced. This type of reactor technology usually operates with particle sizes in the range of 100 to 600 μm, which is high compared to the particle size of 50 μm needed for producing clinker. Therefore, to be able to use the solid purge into the cement production process, CaL solid purge may need to be cooled down and milled. One goal of the project is to investigate if smaller sized particles up to 200 µm can be used in existing fluidized bed reactors and if these particle sizes can be directly used for cement production.

The second analyzed configuration (shown in Figure 4.9) proposes the use of entrained flow reactors for the carbonator and the calciner reactors in the CaL process [MAR1, 2012], [MAR2, 2012] This reactor concept, which has not been experimentally demonstrated yet, fulfills cement process requirement of fine calcined materials (30-50 μm) and any intermediate solid handling device is therefore avoided. Moreover, it exploits the synergies with mature cement process technologies like raw meal precalciner systems.

A list of process units required in the process is given in Table 4.5. In addition to the case specific units, which are the two reactors characterizing the process, general units include fans, cyclones for gas-solid separation at reactors outlet and a heat recovery steam cycle, including its heat exchangers to recover excess heat from the carbonator and calciner off-gases.

Table 4.5: List of process units required in the Calcium Looping process

Case specific units	General units
Oxyfuel calciner	Fans
Carbonator reactor	Cyclones
	Heat recovery steam cycle and heat exchangers

5 PROCESS UNIT SPECIFICATIONS

In this chapter operating specifications are given for standard process units that are used in several of the capture processes. This is done to ensure consistency between process simulations of the respective processes.

5.1 Heat exchangers

Minimum pinch point temperature differences vary with the heat exchanging fluids. Values for different types of heat exchangers are specified in Table 5.1. The values are also dependent on factors such as size, pressure, composition and metallurgy, and the values presented here should be considered as guidance.

Table 5.1: Minimum pinch point temperature difference (ΔT) for heat exchangers

Operating temperature	Fluids	Minimum ΔT
Above ambient temperature	Dusty gas/gas	40 °C
	Dusty gas/liquid	80 °C
	Liquid/liquid	10 °C
	Reboiler	10 °C
	Condenser[1]	10 °C
Below ambient temperature	General minimum	3 °C
	Boiling liquid/condensing liquid[2]	5 °C

[1] Intercoolers in multistage compression is included in this category due to some condensation of water.
[2] For sections of heat exchangers with single-component mixtures at constant boiling and condensation temperature.

Pressure drop in heat exchangers is dependent on phase. Usually liquid phase pressure drop is absolute while gas pressure drop depends on the gas pressure.

- Liquid phase pressure drop for cold and hot side: 0.04 MPa
- Gas phase pressure drop for cold and hot side: 2 % of inlet pressure

Assumptions related to HRSGs can be found in Table 6.8 in Chapter 6.3.

5.2 Separators

Pressure drop in knock-out drums in compressor trains can be neglected. Pressure drop in other types of separators has to be estimated for each particular case.

5.3 Pumps, fans, compressors and expanders

Guidance values for efficiencies of pumps, fans, compressors and expanders are given in Table 5.2. The exact efficiencies will depend on the exact type of unit, the size, and on the operating conditions. For the sake of simplicity, one approximate value is set for each type of unit.

Table 5.2: Efficiencies for pumps, fans and compressors; the values given can be considered as BAT values

Efficiency	Value
Pump efficiency	80%
Isentropic efficiency compressors	85%
Isentropic efficiency expanders	85%
Fan efficiency	85%

Driver efficiencies should be 95%. In case expanders will be used to generate electric power, similar mechanical and generator efficiencies as used in integrated power production (Chapter 6.3) should be used.

5.4   Filters

For process dust abatement two main technologies are available: electrostatic filters (ESP) and bag filters. Both technologies are suitable for the cement kiln gas, having different strong and weak points. The ESP has low efficiency when the gas stream parameters are far from the steady operating conditions, namely start up and shut down periods, while the bag filter is very sensitive to the gas temperature and moisture content; different types of filtration media are available, but in any case the highest temperature can't be higher than 260 °C. Filtration efficiency is very high for both technologies, with residual dust content after treatment below 10 mg/m$^3_{STP}$ (dry state, 10% O_2). The total pressure drop for an ESP is in general, at a gas velocity of 1 m/s, around 0.6 – 0.7 kPa, while for a fabric filter it is 1.2 – 1.5 kPa (0.7 – 0.8 kPa on the filter media and the rest on the filter casing). Additional electrical energy is necessary for ESP to create the electrostatic field between the electrodes, while for fabric filters compressed air is necessary to clean the bags and to operate a gas conditioning system (by water injection or by air to gas heat exchanger). In the end both solutions are consuming 1.5 – 2 kWh/t$_{clk}$ for the kiln-raw mill application (main process emission). This range is valid for installation at a wide range of plants, with plant sizes from 2,000 up to 5,000 t$_{clk}$/d. Maintenance costs are in favour of bag filters, mainly due to the higher complexity of an ESP.

6 CO$_2$ CAPTURE UTILITIES

6.1 Steam

Steam generation is needed in case of CO$_2$ capture by absorption, in order to regenerate the CO$_2$-rich solvent. Waste heat available from cement production process will be used to raise steam for this scope. However, it is expected that waste heat available will not be sufficient to provide the entire heat duty required. Therefore, additional steam generation systems will have to be included.

Two options will be considered in CEMCAP for this scope, namely: (i) steam import from an external CHP plant, and (ii) steam generation in a natural gas-fired boiler. In the last case, preference to a natural gas-fired systems is given in order to limit additional CO$_2$ emissions from fuel combustion and avoid high cost dedicated devices for the control of the emissions of other pollutants such as SO$_x$, NO$_x$ and particulate matter (PM).

6.1.1 Steam produced from heat recovery in the cement power plant

In this case, the heat required to produce the steam is available in the process and needs to be recovered to produce the steam for the capture process. While no climate impact or fuel cost is considered, investment and operating cost associated with the recovery are included. The investments for the waste heat recovery system is based on an investment cost of 0.45 M€/MW suggested from industrial experience.

6.1.2 Steam import from external coal CHP plant

The steam, in this case, is assumed to be bought from a coal power plant near the cement plant location. The steam is assumed to be extracted from the steam flow entering the low pressure (LP) steam turbine of the power plant. Heat to power conversion efficiencies are calculated by considering expansion of steam at initial pressure and temperature of 4 bar and 220 °C, condenser pressure of 0.1 bar, turbine isentropic efficiency of 90% and electrical-mechanical efficiency of the generator of 98%. The cost and climate impact of the steam are based on the lost electricity revenues that the power plant would have received from producing and selling the corresponding electricity. While the climate impact of the steam is based on the climate impact of the electricity it would have produced [ANA, 2011], the economic value of the electricity is based on the average European price presented in Chapter 6.2. The results are summarised in Table 6.1.

In practice, less than 10 % of European existing cement plants are located near a coal power plant and therefore this option is not the most representative alternative. However in the case of a new cement plant to be built with CCS, this alternative could significantly lower the CO$_2$ capture cost associated with the cement plant.

Table 6.1: Cost and climate impact for steam

Steam temperature [°C]	Calculated efficiency for conversion to power [%]	Steam cost [€/MWh$_{th}$]	Steam climate impact [kg$_{CO2}$/MWh$_{th}$]
100	13.3	7.7	101
120	17.8	10.3	136
140	22.3	13.0	170

6.1.3 Natural gas-fired boiler

In case of steam generated in a natural gas fired boiler, natural gas consumption will be evaluated by simply assuming a boiler efficiency of 90% (LHV basis). Boiler efficiency will not depend on the pressure of the steam generated, assuming that air preheating temperature can be adjusted depending on the temperature of the combustion gases exiting the evaporator tubes. The cost of this option is largely dominated[2] by the cost of natural gas.

Steam produced from this option is more expensive than the external steam supply by a coal CHP plant. However, this option results in lower CO_2 emissions, is more flexible in term of range of capacity available and can easily be installed on-site. The cost of this option is based on the fuel cost presented in Chapter 9.5.2, the boiler efficiency and the investment and the operating cost are based on the cost of a natural gas boiler. The climate impact of this option is based on a natural gas emission factor defined in Chapter 6.2.1 and the boiler efficiency of 90%.

6.1.4 CO₂ emissions and cost

The cost and climate impact associated with the different options to produce steam presented above are gathered in Table 6.2. By default, the steam from heat recovery of the cement kiln is considered up to the amount available. If steam is required in addition to what can be extracted from the heat recovery in the cement kiln, it is assumed to be produced by a Natural Gas Boiler as this option is more probable than the import of steam imported from a coal power plant. However sensitivity analyses will be performed to address the influence of the steam cost on CO_2 capture cost of the different technologies.

Table 6.2: Cost and climate impact for steam

Steam source	Steam cost [€/MWh$_{th}$]	Steam climate impact [kg$_{CO2}$/MWh$_{th}$]
Waste heat available on the plant	8.5	0
External CHP steam plant at 100°C	7.7	101
External CHP steam plant at 120°C	10.3	136
External CHP steam plant at 140°C	13.0	170
Natural gas boiler	25.3	224

---

[2] An overnighted CAPEX of 9.5 M€ is estimated for a 150 MW$_{th}$ boiler resulting in investment and maintenance cost of 1.3 €/MW$_{th}$h over the project economic duration.

6.2 Electricity

6.2.1 Reference efficiency and CO_2 emission

Reference efficiency and emissions associated to power generation are needed to quantify the indirect primary energy consumption and indirect CO_2 emissions associated to electricity consumptions or production within the cement plant (see Chapter 10.1).

A first approach for defining the reference efficiency and emissions is to consider the average values for electricity production in the EU-28. In this approach, fuels emission factors in Table 6.3 [IEA, 2005] and fuel consumptions and electricity production in Table 6.4 [EUR1] have been considered as primary data. From these data, the average electric production efficiency and CO_2 emissions have been calculated, as reported in the last two columns of Table 6.4. In particular, average electric efficiency of 45.9% and average CO_2 emissions of 262 kg/MWh have been calculated for the whole EU-28 non-CHP power generation system in 2014. Considering only fossil fuel power generation, efficiency of 40.6% and emissions of 776 kg/MWh have been calculated.

A second approach for defining reference efficiency and emissions is to consider typical values for representative technologies and fuels. Such an approach is justified by the fact that, from an LCA (life cycle assessment) perspective, the actual emissions and primary energy consumption associated to flows of electricity between the cement plant and the grid will depend on the specific technologies used to balance the variation of the power exchanged with the grid. Therefore, a wide variability can be expected from case to case and from country to country, depending on the local market conditions and the regional power generation technologies. Due to this variability, it is recommended to perform a sensitivity analysis considering the different power generation cases reported in Table 6.5. Note that in the case of renewables, the definition of primary energy is not straightforward. Therefore, two different approaches are considered. The first approach is the "physical energy content method", which is used for energy statistics by IEA and Eurostat [IEA2, 2016], [IEA, 2017], [EUR, 2017]. In this approach the primary energy is defined as the first energy flow in the production process that has a practical energy use. For geothermal and solar thermal power the primary energy is then heat, while for solar photovoltaic, wind, hydro, tide, wave, and ocean power, the primary energy is the produced electricity. Since the share of geothermal and solar thermal power plants for electricity generation is small in Europe, this contribution is neglected, and we end up with a reference efficiency of 100%. In the other approach, only fossil energy is counted as primary energy. This corresponds to setting the reference efficiency of renewables to infinity.

Table 6.3: Reference CO_2 emission factor from using different fuels [IEA, 2005]

Fuel source	CO_2 emission factor, c_{CO2}, t_{CO2}/TJ_{fuel_LHV}
Hard coal	94.6
Lignite	101.2
Oil shale and oil sand	106.7
Oil	78.5
Natural gas	56.1
Refinery gas	66.7
Nuclear/Renewables	0
Waste (non-renewable)	98.1

Table 6.4: Electricity generation in EU-28 in 2014 in non CHP power plant

Fuel	2014 electricity production, E, GWh [EUR1]	2014 fuel consumption, E, TJ$_{LHV}$ [EUR1]	2014 average electric efficiency, % (calculated)	CO$_2$ emission factor, kg/MWh (calculated)
Hardcoal and derivatives	352,931	3,246,292	39.1	870.1
Lignite and derivatives	229,815	2,227,256	37.1	980.8
Oil shale and oil sand	10,022	110,523	32.6	1176.7
Oil w/o refinery gas	11,116	101,658	39.4	717.7
Natural Gas	231,142	1,712,945	48.6	415.7
Refinery gas	151	1,188	45.8	525.0
Nuclear	849,485	9,267,109*	33.0*	0
Renewables	830,117	2,988,421*	100.0*	0
Waste (non-renewable)	11,530	143,447	28.9	1220.0
Total	2,526,309	19,798,839	45.9	262
Total, fossil fuels only	835,177	7,399,862	40.6	776

* efficiency of 33% for nuclear and 100% for renewables is assumed, in agreement with [IEA2, 2016]. Primary energy consumption calculated from the assumed efficiency.

Table 6.5: Reference efficiency and CO$_2$ emission factor for power generation technologies, to be considered in the calculation of the key performance indicators. For fossil fuel power generation, CO$_2$ emissions factors have been calculated from the assumed electric efficiency and the fuel emission factors in Table 6.3.

Power generation technology	Electric efficiency, %	CO$_2$ emission factor, kg/MWh
Pulverized coal, state of the art [EUD, 2004]	44.2	770
Pulverized coal, sub-critical	35.0	973
Natural gas combined cycle [EUD, 2004]	52.5	385
Renewables (physical energy content method) [IEA2, 2016], [IEA, 2017], [EUR, 2017]	1	0
Renewables (only fossil primary energy)	∞	0

6.2.2 Cost of electricity

The cost of electricity for an industrial customer depends on the amount of energy required in a year. The electricity demand in a cement plant ranges typically from 90 to 150 kWh/t_{cement} [BREF, 2013]; ECRA's reference case consumes 97 kWh/t_{cement}. Considering (i) ECRA's reference plant size, (3,000 t_{clk}/d), (ii) a clinker/cement ratio equal to 0.737, and (iii) 90 % annual run factor, we can obtain the yearly electric consumption as:

$$E = \frac{Q_{cli.}}{s} e_{cement} \cdot \frac{8760 \cdot \varphi}{24} \qquad \text{kWh/y}$$

where
- Q_{clk} is the amount of clinker produced per day [t_{clk}/d]
- s is the clinker to cement mass ratio
- e_{cement} is the specific electric consumption [kWh/t_{cement}]
- φ is the annual cement run factor

Accordingly, ECRA's reference plant consumes about 130,000 MWh/y. Given the plant size variability along with the different specific electric consumptions, three different industrial consumers' ranges are considered for indications on the cost of electricity [EUR2]:

- Band IE: 20,000 < E <70,000 MWh/y
- Band IF: 70,000 < E < 150,000 MWh/y
- Band IG: E > 150,000 MWh/y

Results are shown in Table 6.6.

In cases in which electricity can be produced in the plant and can be sold outside of the battery limit of the plant, it will be credited with the electricity cost corresponding to the associated consumer band.

Table 6.6: Cost of electricity (€/kWh) for three different yearly energy demands in 2013 and 2014 [EUR2]. The cost for each country is the sum of the energy price and tax. The "Median EU28" is the sum of the median energy price and the median tax.

Country	20,000 < E < 70,000 MWh/y		70,000 < E < 150,000 MWh/y		E > 150,000 MWh/y	
	2013	**2014**	**2013**	**2014**	**2013**	**2014**
Belgium	0.0685	0.0642	0.0641	0.0635	n/a	n/a
Bulgaria	0.0469	0.0529	0.0445	0.0474	0.0442	0.048
Czech Republic	0.0503	0.0434	0.0507	0.0411	n/a	n/a
Denmark	0.0491	0.0339	0.0483	0.0339	n/a	n/a
Germany	0.0953	0.0949	0.0818	0.087	n/a	n/a
Estonia	0.0566	0.0529	0.0548	0.0541	n/a	n/a
Ireland	0.0871	0.0799	0.0784	0.0753	n/a	n/a
Greece	0.0868	0.0695	0.0633	0.0487	n/a	n/a
Spain	0.0824	0.0779	0.0723	0.073	0.0627	0.064
France	0.0551	0.0566	0.0459	0.0455	n/a	n/a
Croatia	0.0487	0.0474	0.0524	0.0476	n/a	n/a
Italy	0.1283	0.1288	0.1076	0.1072	0.0956	0.0864
Cyprus	0.1532	0.1404	0.1506	0.1392	n/a	n/a
Latvia	0.0749	0.075	0.0727	0.0772	n/a	n/a
Lithuania	n/a	0.0841	n/a	n/a	n/a	n/a
Luxembourg	0.0531	0.0514	n/a	n/a	n/a	n/a
Hungary	0.0719	0.0649	0.0664	0.0633	0.0702	0.0607
Malta	0.1203	0.1194	0.1229	0.1136	0	n/a
Netherlands	0.0576	0.0501	0.0567	0.0495	n/a	n/a
Austria	0.0719	0.0686	0.07	0.0641	n/a	n/a
Poland	0.0536	0.0486	0.0529	0.0464	0.052	0.0445
Portugal	0.0685	0.0664	0.0602	0.061	n/a	n/a
Romania	0.0418	0.0422	0.0454	0.0493	n/a	n/a
Slovenia	0.0641	0.0578	0.0629	0.0573	n/a	n/a
Slovakia	0.0547	0.0485	0.0569	0.0486	0.0532	0.0472
Finland	0.0547	0.0501	0.0541	0.05	n/a	n/a
Sweden	0.0478	0.0427	0.0455	0.0396	n/a	n/a
United Kingdom	0.0787	0.0868	0.0774	0.0859	0.0767	0.0837
Liechtenstein	n/a	n/a	n/a	n/a	n/a	n/a
Norway	0.0526	0.0469	0.0501	0.0448	n/a	n/a
Montenegro	0.041	0.0388	n/a	n/a	n/a	n/a
Serbia	0.0346	0.0447	0.0312	0.0467	n/a	n/a
Turkey	0.061	0.0596	0.0607	0.0583	0.0625	0.0576
Bosnia and Herzegovina	0.0354	n/a	n/a	n/a	0.0478	n/a
Median EU28	**0.0558**	**0.0548**	**0.0558**	**0.0509**	**0.0569**	**0.0581**

6.3 Integrated power production

For oxyfuel and CaL capture technologies, a significant amount of waste heat is available from the process. In order to achieve good process efficiencies, such heat can be recovered by feeding a heat recovery steam cycle. The size of such a steam cycle may vary depending on the cement kiln configuration and the thermal integration of the capture section with the raw meal preheating process. Depending on the thermal input of the steam cycle, as well as on the temperature profile of the waste heat available, different steam parameters and turbine efficiencies have to be expected according to economic considerations. As a general rule, the smaller the steam cycle, the lower the maximum steam pressure and temperature, the lower the steam turbine isentropic efficiency due to scale effects.

In conventional cement kilns, heat recovery steam cycles are very seldom adopted, due to the small amount of waste heat normally available. In order to define the parameters to be adopted in the simulation of steam cycles for oxyfuel and CaL cement kilns, reference can be made for example to waste-to-energy (WTE), biomass and small fossil fuel-fired power plants, characterized by plant sizes in a wide range.

A rather complete picture of the steam cycle parameters adopted in WTE plants as function of size is available in [CON, 2012], which constitutes the basis for the definition of the steam cycle parameters used in CEMCAP. Since the definition of WTE steam cycle parameters is constrained by hot corrosion issues related to the presence of chloride and other chemically aggressive species in the flue gas, steam cycle parameters have been defined also based on the information reported in [ANS], [SPL, 2010], [JAC, 2009], [TSC, 2003], [ALS].

Steam cycle parameters as function of thermal input are reported in Table 6.7, while steam turbine efficiencies for WTE plants are shown in Figure 6.1. In Table 6.8, the other assumptions for the calculation of the steam cycles are reported. To facilitate the reproducibility of the steam cycle balances, the material and energy balances for three representative sizes of 12.5, 100 and 300 MW of thermal input are reported below as calculated with the Polimi in-house process simulation tool GS [GSS]. The flowsheet of the power plant is shown in Figure 6.2, the energy balances are reported in Table 6.9 and the properties of the streams in Table 6.10, Table 6.11 and Table 6.12.

It should be noted that the material and energy balances of the CEMCAP steam cycles will likely differ from those reported here, since low temperature heat recovery will be possible so that part (or all) the feedwater preheating may be performed through such heat recovery. Therefore, data reported here should be used just as a way to calibrate and validate steam cycle models implemented in different process simulation tools.

Table 6.7: Steam cycle parameters as function of thermal input

Nominal thermal input, MW	12.5	25	50	100	200	300
Steam pressure at turbine inlet, bar	30	40	60	80	100	125
Steam temperature at turbine inlet, °C	350	400	460	480	530	565
LP regenerative condensate preheater	No	No	Yes	Yes	Yes	Yes
Feedwater temperature at boiler inlet, °C	120	120	140	140	140	140
Estimated turbine isentropic efficiency, %	70.0	75.0	78.0	80.8	85.6	86.8
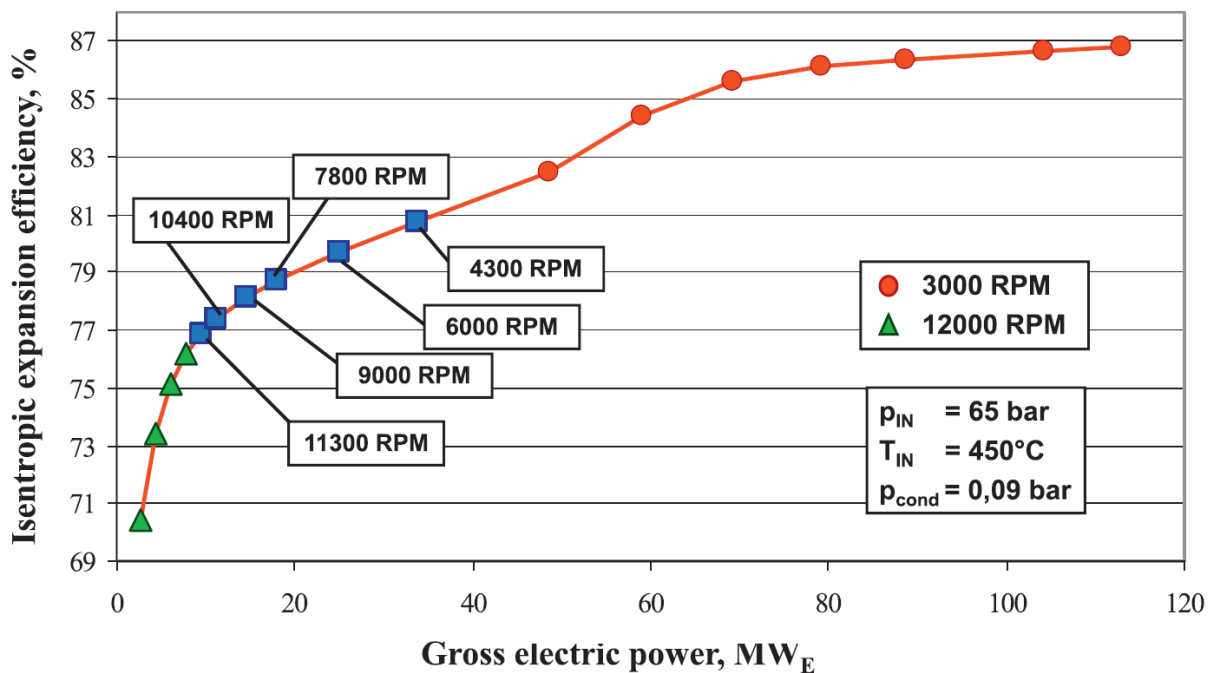
Figure 6.1: Steam turbine isentropic efficiency as function of size for WTE plants [CON, 2012]

Table 6.8: Assumptions for the calculation of the heat recovery steam cycle

Parameter	Value
Reheat	NO
Pressure drop in economizer, bar	20
Pressure drop in superheater	8%
Pressure drop in turbine admission valve	5%
Condensing pressure, bar	0.07
Feedwater pump hydraulic efficiency	75%
Feedwater pump mechanical-electric efficiency	94%
Steam turbine mechanical-electric efficiency @ 3,000 RPM	97%
Steam turbine mechanical-electric efficiency with gearbox	94%
LP steam bleeding pressure (if LP feedwater preheater activated), bar	0.7
Pressure drop of the bled steam for feedwater heating (deaerator)	7%
Pressure drop of the bled steam for feedwater heating (surface preheaters)	5 %
Minimum water-steam temperature difference in feedwater heaters, °C	3
Minimum water-condensate temperature difference in feedwater heaters, °C	5
Electric consumption for heat rejection to ambient as a % of the heat rejected	1%
Deaerator saturation temperature, °C	120 or 140
Number of feedwater preheaters (in addition to the deaerator)	0 or 1
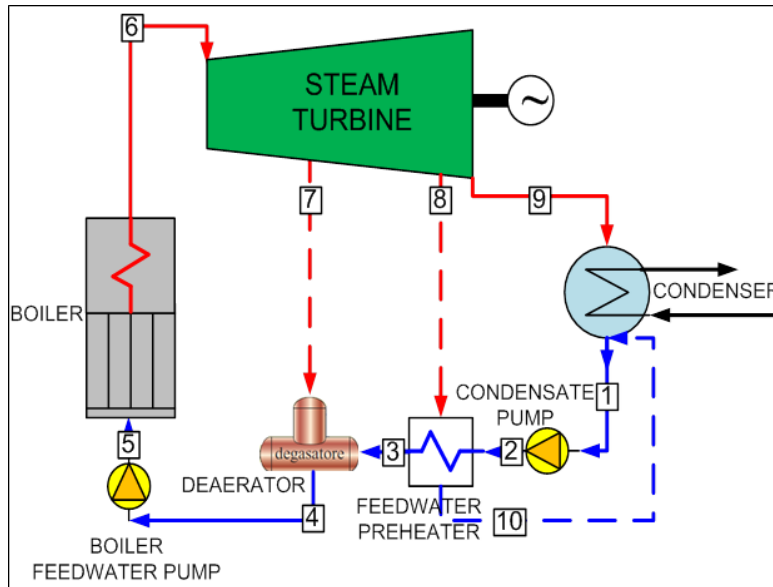
Figure 6.2: Schematic of the steam cycle

Table 6.9: Energy balance of three representative steam cycles with different thermal inputs

Thermal power entering the cycle, MW$_{th}$	12.5	100	300
Energy balance, kW			
Thermal power discharged at condenser	9,366	65,518	182,181
Thermal power exchanged in surface feedwater preheater	-	6,614	18,781
Thermal power exchanged in deaerator	1,418	7,427	21,058
Steam turbine power at blades	3,170	35,050	120,140
Steam turbine electric power at generator	2,979	32,950	116,530
Condensate pump power at blades	1	23	66
Feedwater pump power at blades	35	545	2,258
Condensate pump electric power consumption	1	25	70
Feedwater pump electric power consumption	37	580	2,402
Pumps electric power consumption	38	605	2,472
Condenser auxiliaries consumption	94	655	1,822
Net electric power produced, kW	2,847	31,690	112,236
Cycle gross electric efficiency, %	23.83	32.95	38.84
Cycle net electric efficiency, %	22.78	31.69	37.41

Table 6.10: Properties of the steam cycle streams shown in Figure 6.2, with a heat input of 12.5 MW$_{el}$

Stream ID	Mass flow rate [kg/s]	Temperature [°C]	Pressure [bar]	Enthalpy [kJ/kg]	Entropy [kJ/kg-K]	Steam quality
1	4.16	39.03	0.07	163.40		0
2	4.16	39.03	1.99	163.67	0.558	
3	-	-	-	-	-	
4	4.80	120.02	1.99	503.81	1.528	
5	4.80	120.87	52.61	511.00		
6	4.80	350.00	30.00	3117.54	6.770	
7	0.64	134.97	2.09	2736.47	7.182	
8	-	-	-	-	-	
9	4.16	39.03	0.07	2414.02	7.705	0.9259
10	-	-	-	-	-	

Table 6.11: Properties of the steam cycle streams shown in Figure 6.2, with a heat input of 100 MW$_{th}$

Stream ID	Mass flow rate [kg/s]	Temperature [°C]	Pressure [bar]	Enthalpy [kJ/kg]	Entropy [kJ/kg-K]	Steam quality
1	33.03	39.03	0.07	163.40		0
2	33.03	39.06	4.62	164.10	0.559	
3	33.03	86.96	3.62	364.38	1.157	
4	36.42	140.01	3.62	589.16	1.739	
5	36.42	141.91	106.96	604.00		
6	36.42	480.00	80.00	3349.59	6.684	
7	3.39	143.68	3.89	2780.10	7.007	
8	2.79	91.31	0.74	2555.11	7.169	0.953
9	30.24	39.03	0.07	2327.77	7.428	0.8901
10	2.79	44.06	0.70	184.46	0.626	

Table 6.12: Properties of the steam cycle streams shown in Figure 6.2, with a heat input of 300 MW$_{th}$

Stream ID	Mass flow rate [kg/s]	Temperature [°C]	Pressure [bar]	Enthalpy [kJ/kg]	Entropy [kJ/kg-K]	Steam quality
1	93.70	39.03	0.07	163.40		0
2	93.70	39.06	4.62	164.10	0.559	
3	93.70	86.96	3.62	364.38	1.157	
4	103.38	140.01	3.62	589.16	1.739	
5	103.38	142.80	155.87	611.00		0
6	103.38	565.00	125.00	3512.81	6.632	
7	9.68	155.21	3.89	2764.53	6.971	
8	8.02	91.31	0.74	2526.27	7.090	0.9404
9	85.68	39.03	0.07	2287.54	7.299	0.8734
10	8.02	44.06	0.70	184.47	0.626	

6.4   Air separation unit

There are presently two main different alternatives available in the market concerning oxygen supply:

- oxygen is produced on-site in oxygen generators (air separation units)
- oxygen is produced outside of the cement plant and is delivered afterwards by tank trucks

6.4.1   On-site production

For industrial purposes, atmospheric air separation is the most common oxygen generation process. Air consists mainly of nitrogen, oxygen, argon and small amounts of carbon dioxide. The different air separation technologies are based on different physical and chemical properties of the air components. Three main oxygen generation technologies from air separation are currently known: cryogenic generators, adsorption generators and membrane generators. As cryogenic O_2 production is the most mature technology, adsorption and membrane technologies are not considered in this document.

Cryogenic air separation is currently the most power efficient technology available in the market for the production of large quantities of oxygen. The system working principle relies on the different boiling points of the air elemental components, which can be separated in liquid form at temperatures around -185 °C. The cryogenic technology uses multi-column cryogenic distillation to produce oxygen from compressed air. The air feed is distilled into the elemental streams by a system, which includes a high and low pressure column. Figure 6.3 shows the flow diagram of a cryogenic air separation unit, and Table 6.13 provides technical and economic data.
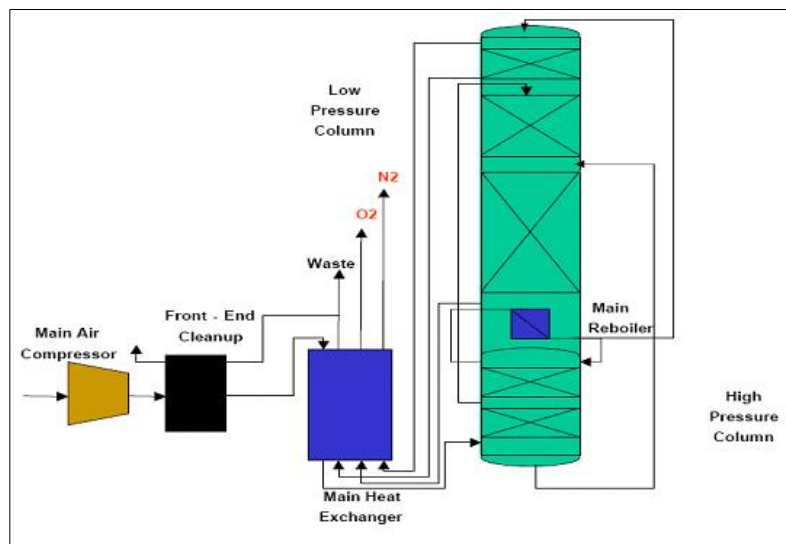

Figure 6.3: Flow diagram of a cryogenic air separation unit

Table 6.13: Technical and economic data of cryogenic air separation

Air separation technology	Oxygen purity [vol.-%]	capacity [t O_2/d]	power consumption [kWh/t O_2]	variable fee [€/t O_2]*
Cryogenic	< 99,9	350 < x < 5,000	220 - 400	> 50

6.4.2 Oxygen supplied by tank trucks

The supply of liquefied oxygen into storage tanks is widely known in the cement industry. Tank trucks coming from an air separation plant located nearby arrive to the cement plant and fill the oxygen storage tanks with liquefied oxygen.

Although the specific costs of liquefied oxygen supply are higher than on-site oxygen generation, depending on the distance of the air separation plant and the required amount of oxygen, it can become a very attractive solution, not only due to its relatively low capital expenditure, but also due to its flexibility.

6.4.3 Oxygen costs

Table 6.14 shows current figures for oxygen costs for the two alternatives considered.

Table 6.14: Oxygen costs for tank delivery system and on-site production

Supply method	Fee	Power costs	Total oxygen costs
Tank system			80 – 100 €/t O_2
Cryogenic separation	>50 €/t O_2	15 – 28 €/t O_2	66 – 91 €/t O_2

6.5 Refrigeration system

The coefficient of performance (COP) for a refrigeration system is defined as the ratio between refrigeration duty and net power requirement:

$$\text{COP} = \frac{\dot{Q}_{\text{refrigeration}}}{\dot{W}_{\text{net}}}$$

The COP for a refrigeration plant is case specific and depends on several factors such as:
- Process layout (single-stage, multiple stages, cascade etc.)
- Refrigerant (single, multiple, mixed)
- Compressor efficiency
- Ambient temperature
- Temperature approaches in heat exchangers
- Plant size and economic trade-off

For the process benchmarking in CEMCAP a simple framework for estimation of the power requirement for refrigeration is made. A generic correlation is developed based on simulations using CoolPack [COO, 2017] of the following four standard refrigeration cycles:

- One-stage cycle, DX evaporator, Refrigerant R134a
- One-stage cycle, DX evaporator, Refrigerant R290
- One-stage cycle, DX evaporator, Refrigerant R717
- Two-stage cascade system, DX evaporators, Refrigerants R290 and R170

The one-stage cycle processes are selected for evaporator temperatures from 233 K to 283 K, and the two-stage cascade process is selected for lower evaporator temperatures. Cooling water temperature (18.2 °C), heat exchanger pinches, and compressor efficiencies are taken as defined in other sections of this document.

The resulting COP is plotted against evaporator temperature in Figure 6.4. It can be observed that the result is very similar for the different refrigerants in the one-stage cycle. A polynomial of sixth degree giving COP as function of evaporator temperature is obtained by regression. It is shown as a grey dotted line in Figure 6.4, and it is used to obtain COP values for evaporator temperatures from 214 K to 283 K in Table 6.15.
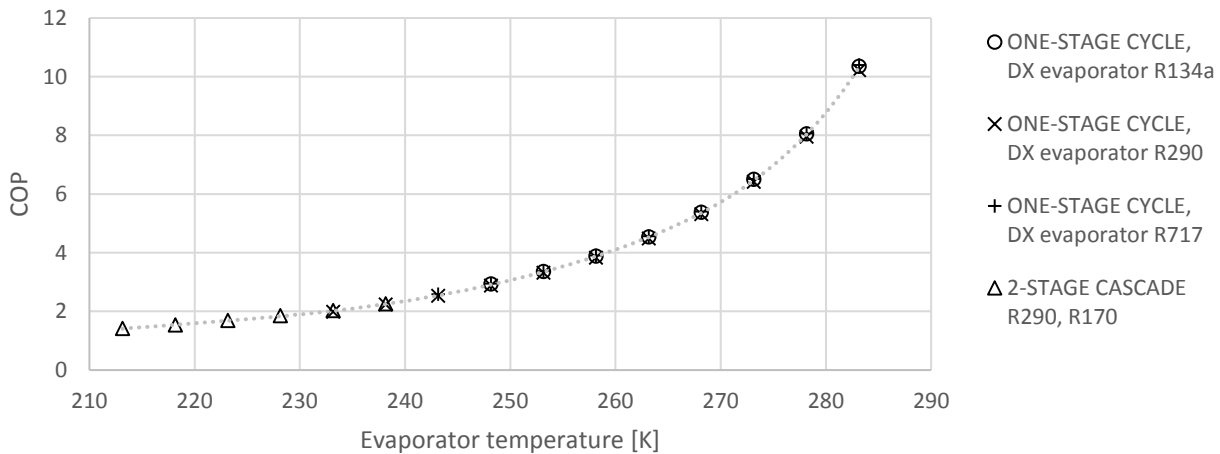
Figure 6.4: Relation between cold-side refrigeration temperature and COP

Table 6.15: Relation between cold-side refrigeration temperature and COP

T [K]	COP		T [K]	COP		T [K]	COP		T [K]	COP
283	10.23		265	4.81		247	2.81		229	1.86
282	9.70		264	4.65		246	2.74		228	1.83
281	9.22		263	4.50		245	2.67		227	1.79
280	8.77		262	4.36		244	2.60		226	1.76
279	8.35		261	4.23		243	2.53		225	1.73
278	7.97		260	4.10		242	2.47		224	1.71

T [K]	COP		T [K]	COP		T [K]	COP		T [K]	COP
277	7.61		259	3.98		241	2.41		223	1.68
276	7.28		258	3.86		240	2.35		222	1.65
275	6.98		257	3.75		239	2.29		221	1.62
274	6.69		256	3.64		238	2.24		220	1.59
273	6.42		255	3.53		237	2.19		219	1.57
272	6.17		254	3.43		236	2.14		218	1.54
271	5.94		253	3.33		235	2.09		217	1.51
270	5.72		252	3.24		234	2.05		216	1.48
269	5.52		251	3.15		233	2.01		215	1.46
268	5.32		250	3.06		232	1.97		214	1.43
267	5.14		249	2.97		231	1.93		213	1.40
266	4.97		248	2.89		230	1.89		212	1.38

By examining the behaviour of the actual COP compared to the theoretical maximum, a general trend that can be used for estimation of COP for low refrigeration temperatures can be observed.

The COP can be looked upon as an efficiency η multiplied with the theoretical maximum coefficient of performance COP_{Carnot}:

$$COP = \eta \cdot COP_{Carnot}$$

The COP_{Carnot} is dependent on the temperature of the two reservoirs that the process transfer heat between. We define the two reservoirs as the cooling water with temperature T_{cw} and the evaporator with temperature T_{evap}. The resulting COP_{Carnot} is:

$$COP_{Carnot} = \frac{T_{evap}}{T_{cw} - T_{evap}}$$

The efficiency η is plotted against evaporator temperature in Figure 6.5. It can be observed that the efficiency η goes towards a constant value when the evaporator temperature decreases (0.5 with the equipment specifications used in CEMCAP). This can be used to estimate COP for low refrigeration temperatures.
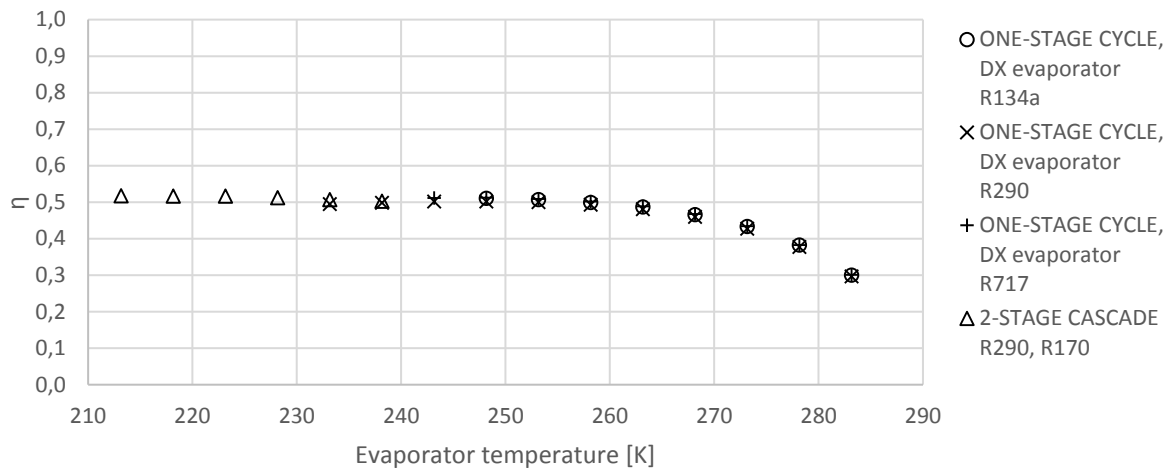
Figure 6.5: Relation between η=COP/COP$_{Carnot}$ and cold-side refrigeration temperature.

6.6 Heat rejection system

The cooling system is assumed to be a closed loop cooling system, where water cooled in a forced draft cooling tower is used as cooling medium. The site is assumed to be at inland with ambient temperature of 15 °C and 60% humidity (see Chapter 2), corresponding to a wet bulb temperature of 10.8 °C. Following EBTF assumptions [EBTF, 2011], cooling water leaves the cooling tower at 18.2 °C, corresponding to a temperature approach of 7.4 °C. The temperature rise of cooling water in process heat exchangers is assumed equal to 11 °C, so that temperature of cooling water returning to the cooling tower is 29.2 °C. A power consumption of 2% of the thermal power rejected is assumed, associated to fans consumption and water pumping. Water consumption of 0.61 kg/MJ of heat rejected is assumed, resulting from assuming a saturated air at 25 °C at the cooling tower outlet and a water make-up flow equal to twice the amount of evaporated water (i.e. a purge water stream from the cooling water circuit equal to the evaporated water is considered to avoid salt build-up).

7   EXTENT OF CO₂ CAPTURE

In order to compare different technologies on the same basis, the cases studied should have the same extent of CO_2 capture. In CEMCAP it is chosen to handle this by fixing the *CO₂ avoided from the flue gas* (AC_{fg}) to 90%. This KPI is defined as the direct CO_2 emission reduction from the flue gas of the cement kiln (see Chapter 10). In addition, an optional case should be selected individually for each technology.

Thus, the following values for CO_2 avoided from the flue gas will be considered:
- 90%
- Optional (other than 90%)

8 CO$_2$ AND FLUE GAS SPECIFICATIONS

In this chapter specifications for CO$_2$ transport and destination, and standard CO$_2$ conditioning processes are presented.

The temperatures and pressures required for the conditioned CO$_2$ are defined by the CO$_2$ transport option. The CO$_2$ purities required are defined by the CO$_2$ transport and destination in combination. The CO$_2$ purification required to reach the purity limits takes place mainly during CO$_2$ capture, but also during the CO$_2$ conditioning. The CO$_2$ purity requirements apply to the conditioned CO$_2$.

8.1 CO$_2$ transport

8.1.1 Pipeline transport

Most international studies on CCS with transport by pipeline have selected CO$_2$ delivery pressures of 11.0 MPa, among others the studies made by IEAGHG, while some studies have selected 10.0 MPa. The CASTOR and ENCAP projects have used 11.0 MPa. The ENCAP project used temperature of <30 °C. These conditions were adopted by the EBTF [EBTF, 2011], and are used also here:

- Pressure: 110 bar
- Temperature: <30 °C

NETL provided in 2012 quality guidelines for energy system studies on CO$_2$ impurity design parameters [NET1, 2012]. They reviewed specifications in other studies, pipeline design guides, pipe transportation specifications and sequestration plant experience, before they set their recommended design parameters (see Table 8.1). The limits are set based on considerations such as toxicity, compression work and corrosion. These parameters are adopted in CEMCAP. An ISO standard on CO$_2$ pipeline transportation systems (ISO 27913) was published in November 2016, i.e. after the CEMCAP simulation work started. In this standard, no standard limits for impurities were set, but indicative levels were given, with some slight differences from the NETL guidelines.

Table 8.1: Minimum CO$_2$ concentration and maximum impurity concentrations for pipeline transport [NET1, 2012]

Component	Limit
CO$_2$	>95 vol%
H$_2$O	<300 ppm$_{wt}$
SO$_2$	<100 ppm$_v$
NO$_x$	<100 ppm$_v$
CO	<35 ppm$_v$
O$_2$	<4 vol%
N$_2$	<4 vol%
Ar	<4 vol%
NH$_3$	<50 ppm$_v$

The following additional impurities can be present in the flue gas: HCl, HF, Hg, other heavy metals, organic carbon and dust. However, the concentrations of these components after the capture processes will be very low. Also, there are no limits set for these components for transport and storage in other studies found in the literature. Therefore the limits for these components are left open in CEMCAP.

8.1.2   Ship transport

For transport of CO_2 by ship it is assumed that the CO_2 will be in liquid form close to the triple point in order to obtain a high CO_2 density and avoid ships with too expensive pressure vessels. The conditions selected by Aspelund et al. [ASP, 2006] are used:

- Pressure: 6.5 bar
- Temperature: subcooled with 3 °C (-52 °C for pure CO_2)

Little work has been done on defining purity requirements for ship transport. It is assumed that the purity requirements should be at least as strict as for pipeline transport. In order to avoid operational problems in the liquefaction process, stricter limits will apply for some components. A tentative limit for water content is <50 ppm [ASP, 2007].

8.2 CO_2 destination

8.2.1   Storage

The specifications for storage as recommended by NETL [NET1, 2012] are adopted, and they are identical to the specifications for pipeline transport (see Table 8.1).

8.2.2   EOR

In CEMCAP the focus is not on EOR. However if EOR for any reason should be considered, the specifications are almost identical to the specifications for storage. The exceptions are stricter limits for N_2, Ar and H_2 (1 vol.-%) and O_2 (0.01 vol.-%). These limits are also adopted from the NETL recommendations [NET1, 2012].

8.2.3   Utilisation

There are several forms of CO_2 utilization, ranging from chemical conversion into various products to use of CO_2 as a solvent. The required specifications of the different routes vary widely in terms of temperature, pressure, and purity. Several uses, their stage of development, and the required pressure of the CO_2 are presented in Table 8.2. For processes which are still under development, the required pressures are not yet known to a high level of certainty, as experiments are still conducted under laboratory conditions. The required purity of CO_2 is in general not known for these processes, except for food applications (99.9 % purity), as industrial grade CO_2 is already 99 % pure and the effect of less pure CO_2 has not been adequately studied. For the processes still at the laboratory stage, examples of tests with CO_2 sourced from capture systems are very few. Some of the catalysts used in these processes may be very sensitive to impurities while others will not be affected.

Table 8.2: Uses of CO_2 and the required CO_2 pressure

Use	Stage of development	CO_2 pressure [bar]	Notes
Urea production	Commercial	140 – 175	CO_2 is sourced both from ammonia production and flue gas (when needed)
Methanol production (conventional)	Commercial	40 – 100	Actual pressure dependent on catalyst and operating temperature
Methanol production (renewable)	Pre-commercial	~50	Processes are still proprietary
Organic carbonates	Laboratory	1 – 100	Processes vary widely
Polyurethanes	Laboratory, Pre-commercial	1 – 100	Processes vary widely; pre-commercial process is proprietary
Technological	Commercial	>74	Usually supercritical CO_2 used for various applications (extraction, solvation, dry-washing, etc.)
Food	Commercial	7 – 8	Mostly for carbonated beverages

8.2.4   Mineralisation

Mineralisation – i.e. the extraction of an alkaline earth metal from a precursor material and its subsequent precipitation with CO_2 in the form of a carbonate mineral – requires little pre-processing of the captured CO_2. The conventional research suggests that CO_2 should be processed at pressures similar to the pipeline pressure. Thus, if captured and compressed CO_2 is pipelined to a mineralization plant next to a distant disposal site, the constraints on pipeline operations are likely to exceed pre-processing needs for mineralisation [IPCC, 2005]. More recent research points towards process concepts that can accept very low CO_2 pressures. In this case, a preliminary capture step becomes optional, only moderate or no compression is needed, and the mineralisation plant has to be built next to the point source/capture plant [WER, 2014]. For both the high and the low pressure approach, CO_2 purity requirements are minimal; acidic components of the flue gas would be neutralized by the alkaline earth metal as they pass through the process and can then be disposed of together with the carbonate product. It may be the case that the purity requirements for the product of the mineralisation process are stringent, depending on the end use/storage of the product compounds. The purity requirements would then be governed by the product quality specifications rather than by processing constraints, but pipeline specifications are expected to still be sufficient.

8.3 CO_2 conditioning

In this section standard CO_2 compression and liquefaction processes are presented. The feed conditions are different for each CO_2 capture technology. The product temperatures and pressures are defined by the transport option. The product purities are defined by the transport option and the CO_2 destination in combination.

8.3.1 CO₂ compression

The various technologies produce CO_2 at different conditions. Therefore the number of stages and pressure ratios for CO_2 compression will be selected for each case individually. However, the following guidelines are set:

- pressure ratios: 2 - 4
- similar pressure ratio for all stages
- isentropic compressor efficiency: 85%
- pump efficiency: 80%
- pump and compressor driver efficiency: 95%

Pressure drop in gas coolers should be 2% of inlet pressure, in liquid coolers 0.4 bar, and in scrubbers neglected, as specified in Chapter 5. Assuming a pinch temperature of 10 °C (since the CO_2 condenses) the intercoolers can cool the CO_2 to 28 °C.

In each scrubber condensate (mainly water) is removed. Before the last compression stage, the gas is dehydrated using TEG if necessary in order to reach the limit of 300 ppm$_{wt}$ of water. Pressure drop in the TEG scrubber is assumed to be 5% of the inlet pressure.

As an example, a suggested compression process for CO_2 after MEA capture (feed ~1.5 bar) is given in Table 8.3 and Figure 8.1. The electric power demand for compression is in this case estimated to 0.33 MJ/kg CO_2. This estimation is for pure CO_2 saturated with water. It is performed with Aspen Plus using the GERG2008 property package. It should be recalculated for each particular case, also taking other impurities into account.

Table 8.3: Pressure ratios and isentropic efficiencies for CO₂ compression after MEA capture

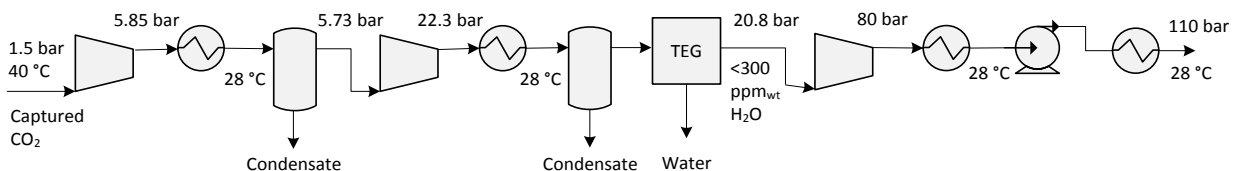|	Inlet pressure [bar]	Discharge pressure [bar]	Pressure ratio [-]	Isentropic efficiency [%]
Stage 1	1.50	5.85	3.9	85
Stage 2	5.73	22.3	3.9	85
Stage 3	20.8	80.0	3.8	85

Figure 8.1: CO₂ compression and dehydration after MEA capture

8.3.2 CO₂ liquefaction

The standard liquefaction cycle for CO_2 captured by MEA (feed ~1.5 bar) shown in Figure 8.2 is assumed to operate with two intercooled compression stages and three dehydration stages. Each intercooling stage consists of ambient cooling to 28 °C and chilling to 4 °C. This is done to maximise the water removal rate in the water separators. After the first compression, intercooling and chilling stage about 97 % of the water content is removed in the water separation drum. The load on the final adsorptive dehydration unit is minimised and has a water removal duty of about 1.5 % of the total water content.
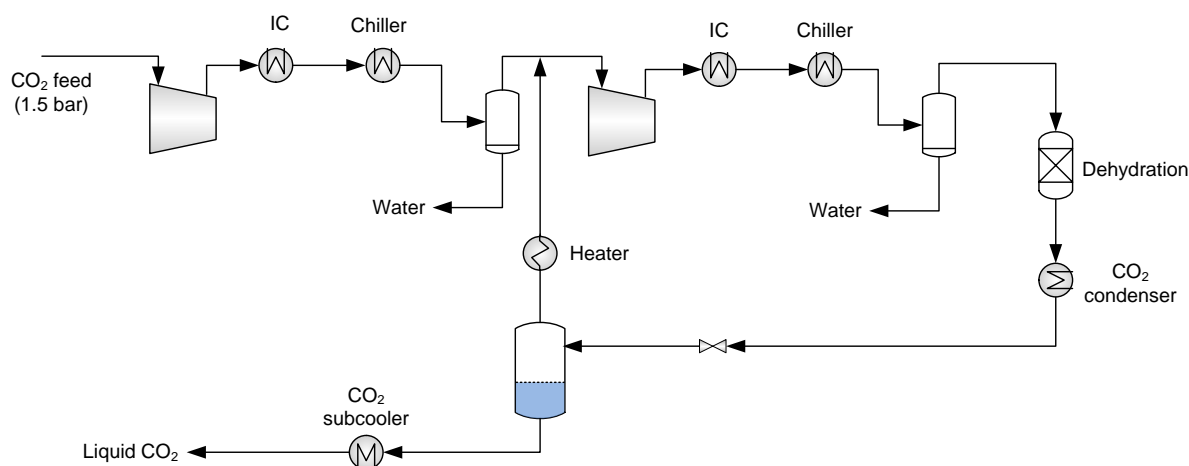

Figure 8.2: Standard CO₂ liquefaction cycle for ~1.5 bar feed pressure

The pressure at the inlet of the CO_2 condenser is 16.2 bar, giving a CO_2 condensation temperature of -24.9 °C. The cold-side evaporation temperature of the auxiliary refrigeration is assumed to be -30.5 °C and cools the CO_2 into subcooled state at -27 °C. At the outlet the pressure is about 15.8 bar and the stream is throttled to 6.9 bar. The vapour fraction of the throttled stream is 0.129 and this stream is recycled to the suction side of the second compressor stage. Since this cold stream is mixed with the feed stream still containing a certain water fraction, the flash gas is assumed to be heated to about 0 °C in order to prevent ice formation and accumulation in the manifold of the gas mixer.

The liquid stream from the flash tank is assumed to hold equilibrium temperature and is subcooled with 3 °C by auxiliary refrigeration (to -52 °C for pure CO_2).

Pressure drops in heat exchangers are specified according to the data in Chapter 5.1. A pressure drop of 50 kPa is assumed for the dehydration unit, while pressure drops have been neglected in the water separation drums. Pressure drops in these drums would increase the power requirement somewhat.

Table 8.4: Pressure ratios, isentropic efficiencies and power consumption for compressors

	Isentropic eff [%]	Pressure ratio	P$_{in}$ [bar]	P$_{out}$ [bar]	Power [kJ$_e$/kg]
Compressor 1	85	3.9	1.5	5.9	114.5
Compressor 2	85	3.1	5.6	17.4	86.8

Table 8.5: Auxiliary refrigeration

	T_{cold} [°C]	COP/COPcarnot	COP	Cold duty [kJ_{th}/kg]	Power [kJ_p/kg]
Chillers	1	0.35	6.85	56.6	8.3
CO_2 condenser	-30	0.48	2.59	384.8	148.5
CO_2 subcooler	-55	0.48	1.49	6.0	4.0

Table 8.6: Overall power requirement

		Specific power
CO_2 compression	kJ_e/kg	201.3
Auxiliary refrigeration	kJ_e/kg	160.8
Sum[a]	kJ_e/kg	362.1
Net[a,b]	kJ_e/kg	381.1

[a] Not including regeneration of dehydration unit
[b] Including 0.95 mechanical efficiency

8.4 Flue gas

The temperature of the flue gas should be sufficiently higher than the dew point in order to avoid condensation and problems with corrosion. Further, a high temperature of the flue gas ensures enough draft in the stack and a wide spread of the flue gas after it has left the stack. If the temperature of the flue gas is not sufficiently high, reheating or alternative technologies should be considered to ensure good dispersion of the plume.

For oxyfuel and calcium looping capture, it is important to pay attention to the operating ranges of the process filter, and that heat is required for evaporation of moisture in the raw mill.

The content of pollutants in the emitted flue gas is subject to regulations. However, since the focus in CEMCAP is on CO_2 capture retrofit, it is assumed that the emission of pollutants is acceptable if it is equal to or lower than the emissions of the reference kiln.

9 ECONOMIC ANALYSIS

9.1 Introduction

One of the objectives of CEMCAP is to evaluate the economic viability of CO_2 capture from cement using different capture technologies. In order to perform the assessment and comparison in a consistent manner, a cost methodology has been developed and is presented in the following sections.

The economic assessment consists mainly of estimations of investment costs, fixed and variable operation and maintenance costs of the cement plant with and without CO_2 capture. Investment costs are evaluated based on a Bottom Up Approach for the CO_2 capture and conditioning process while the cement plant cost are based on a Top Down Approach. The operation and maintenance costs are estimated based on material replacement and a factor approach, while the variable operating costs are based on material, fuel and energy consumptions.

Based on these costs, relevant key performance indicators such as the break-even cement cost with CO_2 capture and the break-even CO_2 avoided cost. Finally, sensitivity analyses shall be performed in order to quantify the impact of several parameters on the economic performances of the cement plant with CO_2 capture.

9.2 Economic boundaries

This section specifies the main assumption considered in order to perform the economic evaluation of the cement kiln with and without CO_2 capture and conditioning.

All economic assessments will be reported on a 2014 basis. In cases in which all costs are not directly available in 2014 prices, investment costs are adjusted to the correct level of prices using the Chemical Engineering Plant Cost Index (CEPCI) for the CO_2 capture and conditioning facilities and the cement kiln costs. The CEPCI is developed in the US, but it is assumed that it is representative also for Europe, and that it can be used directly to convert from euro for a specific year to euro in 2014. Yearly average values for the CEPCI are given in Figure 9.1 and Table 9.1, and the exchange rates from euro to US dollars and GB pounds are given in Table 9.1.

The cement kiln and the CO_2 capture and conditioning process are considered to have an economic lifetime of 25 years [KOR, 2013]. However the economic lifetime of the plant may differ from the expected lifetime of certain equipment and material of the process, such as membranes and sorbent. Therefore cost associated with periodic replacement of such parts shall be included. As in practice the operating lifetime of the plant and certain equipment can be longer or shorter than here considered, sensitivity analyses on these parameters shall be performed.

A plant construction period of 3 years for the capture process and 2 years for the cement kiln are here considered. The annual allocation of construction costs used for evaluation of the overall cost is presented in Table 9.2.
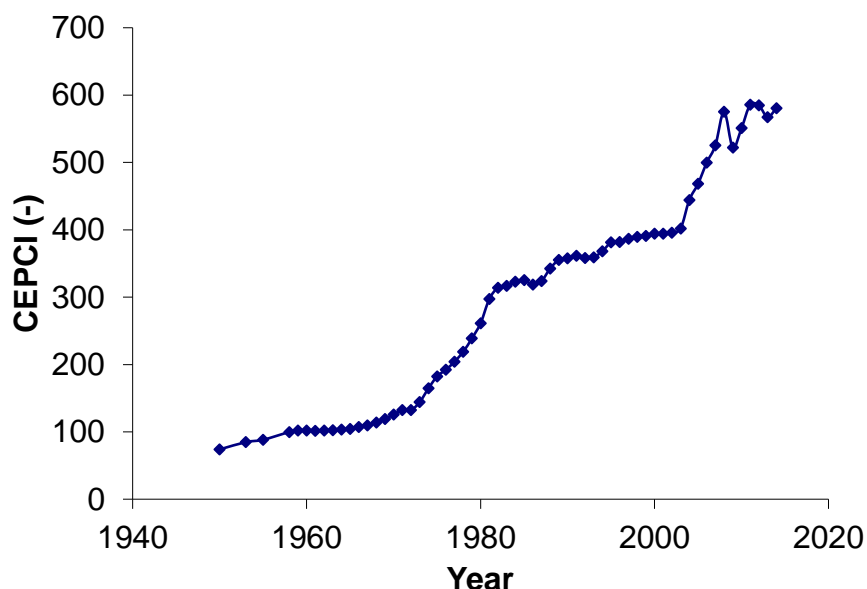
Figure 9.1: Chemical Engineering Plant Cost Index [CEP, 2014].

Table 9.1. Yearly average Chemical Engineering Plant Cost Index [CEP, 2014] and exchange rates [OFX, 2017] for 2001-2014.

Year	CEPCI	USD/EUR	GBP/EUR
2001	394.3	0.8960	0.6220
2002	395.6	0.9425	0.6281
2003	402.0	1.1341	0.6933
2004	444.2	1.2441	0.6787
2005	468.2	1.2464	0.6846
2006	499.6	1.2563	0.6819
2007	525.4	1.3704	0.6845
2008	575.4	1.4714	0.7961
2009	521.9	1.3945	0.8916
2010	550.8	1.3274	0.8585
2011	585.7	1.3927	0.8679
2012	584.6	1.2857	0.8112
2013	567.3	1.3285	0.8491
2014	571.6[1]	1.3292	0.8065

[1] In deliverables D4.1 and D4.2 a preliminary CEPCI value for 2014 (580.2) was used since the final number was not available yet. In remaining deliverables the final 2014 CEPCI (571.6) will be used.

Table 9.2: Annual allocation of construction costs [ANA, 2011]

Year	Allocation of cement kiln cost [%]	Allocation of CO_2 capture cost [%]
1	50	40
2	50	30
3		30

9.3 Financial parameters

A discounted cash flow approach is here considered to perform the project valuation. A real discount rate (i.e. without inflation) of 8 % is considered in order to perform the project financial valuation. The discount rate represents an average cost of capital which takes into account both the return rates on equity and interest rates on loans. As the discount rate is specific to the type of investor, sensitivity analyses will be performed on this parameter. For example, the State uses lower discount rates than average companies while Oil & Gas companies and companies dealing with risk use a higher discount rate.

As both corporate taxes and depreciation fiscal accounting varies significantly among European member states, the project evaluation is performed on a pre-taxation basis. Finally, as the aim of the CEMCAP project is to evaluate the cost of capturing CO_2 from cement kilns, no CO_2 emission taxation is here considered.

9.4 Capital investment

Two approaches are considered in order to evaluate the Total Plant Cost (TPC): a Bottom-Up approach and a Top-Down approach [ANA, 2011]. As only a very limited number of cement kilns is built every year in Europe, the cost provided in the project will be given for a retrofit case.

The Bottom-Up approach
A Bottom-Up approach (BUA) is used to estimate the EPC costs for the CO_2 capture and conditioning processes. A schematic overview of the BUA is given in Figure 9.2.
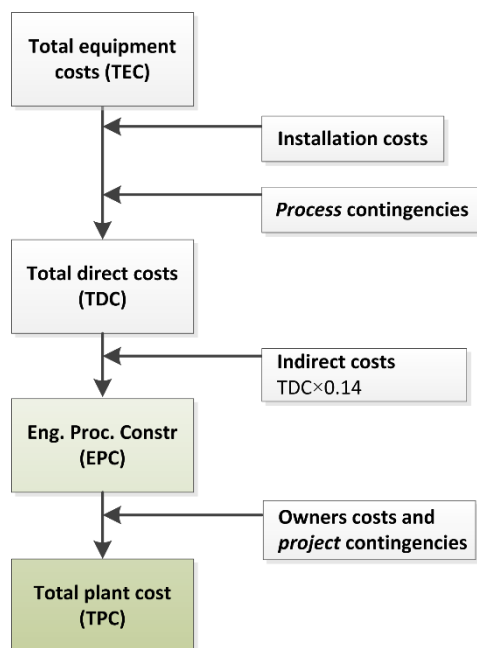

Figure 9.2: The Bottom-Up approach for estimation of total plant costs

The following cost elements are included:

- *Equipment Costs* – The Equipment Cost for each main basic equipment of the different processes can be estimated based on a step-count exponential costing method, using the dominant or a combination of parameters derived from mass and energy balance computations, combined with cost data obtained from equipment suppliers and/or other available data. The *Total Equipment Cost (TEC)* is the sum of all Equipment Costs in the plant.

- *Installation costs* – The installation costs are estimated as additional expenses to integrate the individual equipment into the plant, such as costs for piping/valves, civil works, instrumentations, electrical installations, insulations, paintings, steel structures, erections and OSBL (outside battery limits).

- *Total Direct Cost (TDC)* – The Direct Cost is the sum of the Equipment Costs and the Installation Costs and shall also include the appropriate process contingency factor in order to reflect the differences in technology maturity between the considered CO_2 capture concepts as shown in Table 9.3.

Table 9.3: AACE guidelines for process contingency [NET, 2011]

Technology Status	Process Contingency [% of Associated Process Capital]
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

- *Indirect Costs* – The indirect expenditures are fixed to 14 % of the TDC for all CO_2 capture and conditioning technologies considered and include the costs for the yard improvement, service facilities and engineering costs as well as the building and sundries. An indicative breakdown of the Indirect Costs is given in Table 9.4.

Table 9.4: Breakdown of indirect costs

Indirect costs	% of TDC
Yard improvement	1.5
Service facilities	2
Engineering/consultancy cost	4.5
Building	4
Miscellaneous	2

- *Engineering, Procurement and Construction Costs (EPC)* – The EPC is the sum of Total Direct Cost and Indirect Costs.

- *Owner's Costs and Contingencies* – The owner's costs for planning, designing and commissioning the plant and for working capital, together with project contingencies, are set to 19 % of the total EPC cost, in the case of commercial technologies, following AACE 16R-90 guidelines for AACE Class 4 budget estimates[3].

  However, as most of the capture concepts considered in this study are not fully mature, the owner's costs and contingencies shall be corrected as presented in Table 9.5[4] specifically based on the level of maturity of the concept considered. The maturity levels corresponding to each concept can for example be extracted from the literature [IEA, 2014].

Table 9.5: Owner's costs and contingencies guidelines

TRL level	Level of maturity	Correction factor	Owners costs and contingencies [% total EPC costs]
9	Mature	1	19
8	Successful	1.04	24
7	Proven	1.09	30
5-6	Growing	1.14	36
4	Promising	1.20	42
1-3	Experimental	1.25	49

- *Total Plant Cost (TPC)* – The total capital investments is the sum of EPC, owner's costs and contingencies.

The Top-Down approach

A Top-Down approach is used to estimate the EPC of the cement kiln:

- *Engineering, Procurement and Construction cost (EPC)* is estimated directly, based on equipment supplier estimates of EPC costs for the entire cement kiln.

- *Owner's Costs and Contingencies* – The owner's costs for planning, designing and commissioning the plant and for working capital, together with contingencies, are set to 15% of the total EPC cost for all the technology options.

- *Total Plant Cost (TPC)* – The total capital investments consist of EPC, owner's costs and contingencies considering the concept level of maturity.

The cost estimates are for 'nth plants' based on current knowledge of the technology, i.e. they are commercial plants built after the initial technology demonstration plants. Additional costs

---

[3] AACE 16R-90 states that project contingency for a "budget-type" estimate (AACE Class 4) should be 15 % of the sum of the TDC (including process contingency) which combined to the 7% [IEA, 2016] of the TDC considered for the owner's cost lead to 19% of the EPC costs.

[4] Correction factor are estimated based on the learning curve approach suggested by NETL [NET2, 2012] considering the differences in cost for the different maturity levels in the case of the 20th plant (which is expected to have cost representative of the Nth Of A Kind cases).

normally associated with 1st-of-a-kind commercial plants shall be excluded. The estimate accuracy is expected to be +35/-15 % (AACE Class 4).

If not directly provided for 2014 prices, investment costs are adjusted to the correct level of prices using the CEPCI for the CO_2 capture and conditioning facilities and the cement kiln costs.

It is worth noting that the development and land costs are not considered in the project estimates.

9.5   Main operating cost parameters

The designed capture processes are set to operate in base load operation. However, in practice technical issues in the first years lead to a lower capacity factor before reaching the base load. Therefore, during the first two years of operation, capacity factors of 40% and 65% are presumed respectively, while a capacity factor of 87 % is achieved afterwards.

An overview of the elements of the included operating costs is given in Figure 9.3, and more details about the cost estimations are given in the following text.
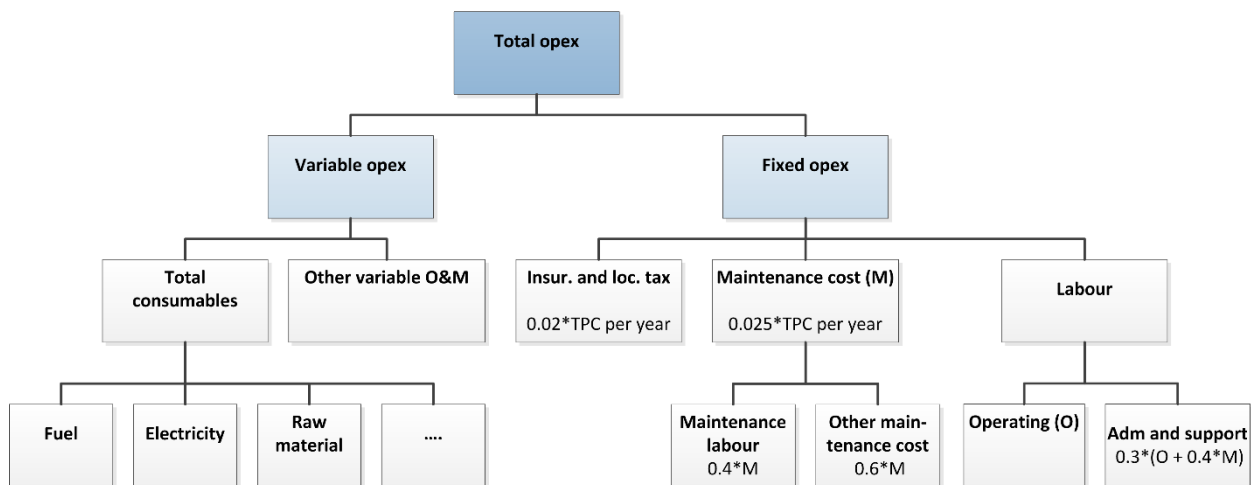

Figure 9.3: Elements of operating costs in CEMCAP

9.5.1   Maintenance, insurance and labour costs

The fixed operating costs which include maintenance, insurance and labour costs are estimated as follows:

- *Insurance and local property taxes:* The total annual cost of insurance, local property taxes and miscellaneous regulatory and overhead fees is to be a total of 2% of TPC [ANA, 2011].

- *Maintenance cost:* Maintenance costs include cost of preventive maintenance, corrective maintenance (repair and replacement of failed components) and periodic replacement of materials. A maintenance cost corresponding to 2.5 % of the TPC excluding periodic replacement of materials which are defined based on the process [ANA, 2011]. This cost includes the maintenance labour cost, corresponding to 40% of the total maintenance cost.

- *Labour costs:* Labour costs include operating labour, administrative and support labour. 100 persons are assumed to be employed for operative labour [IEA, 2013]. The 'fully burdened' cost of labour, including social security payments is assumed to be 60 k€ per employee. Administrative and support labour are also included and assumed to be 30 % of the operating and maintenance labour cost while maintenance labour are assumed to be 40 % of the overall maintenance cost [KOR, 2013].

9.5.2   Utilities and consumables cost

The variable operating costs include material utilities consumption such as electricity, natural gas, process water, chemicals, sorbent, etc. The costs of the main utilities and consumables are evaluated based on the process energy and mass balance and the costs presented in Table 9.6. As discussed in Chapters 6.2 and 6.1, certain utilities such as electricity and steam can be obtained at different costs depending on the source, amount and location. Considering the electricity of the plant with CCS, it seems reasonable to assume that the electricity cost which shall be considered for the cost evaluation correspond to the highest electricity consumption band presented in Chapter 6.2, while the steam cost depends on the steam demand for the CO_2 capture plant as presented in Chapter 6.1. Sensitivity analyses will be performed on these values in order to evaluate their impacts on the key performance indicators.

Table 9.6: Costs of main utilities and consumables

Utilities and consumables	Cost	Unit	Range
Electricity	58.1	€/MWh	
Natural gas [KOR, 2013]	6	€/GJ	
Coal [ANA, 2011, KOR, 2013]	3	€/GJ	
Steam from heat recovery	3.5	€/MWh	Until the amount which can be produced from the cement kiln wasted heat
Steam from Natural Gas Boiler	25.3	€/MWh	Above the amount which can be produced from the cement kiln wasted heat
Process water [KOR, 2013]	6.65	€/m^3	
Cooling water [KOR, 2013]	0.39	€/m^3	
MEA	1,450	€/t	
Raw meal, incl. grinding (limestone, iron oxide, sand, etc.)	5.0	€/t$_{clk}$	
Alternative fuels	15 - 25	€/t$_{clk}$	

9.5.3   Other variable Operating costs

In addition to the aforementioned operating costs, other miscellaneous Opex of 0.8 €/t$_{cement}$ (or 1.09 €/t$_{clk}$) are assumed, coherently with [IEA, 2013].

10 KEY PERFORMANCE INDICATORS

In this chapter energy, environmental and economic indicators are defined for a subsequent benchmarking of capture technologies. In this connection, also the Benchmark Analysis from the Norcem CO_2 Capture Project will be taken into account.

10.1 Energy and environmental performance indicators

In the reference configuration, the cement kiln production requires the use of fuel for sustaining calcium carbonate decomposition in the precalciner and the formation of the main clinker constituents in the rotating kiln. Such processes entail direct fuel consumptions and direct CO_2 emissions within the cement plant boundaries. Such consumptions and emissions are defined as follows:

- Direct specific primary energy consumption:

$$q_{clk} \left[\frac{kJ_{LHV}}{kg_{clk}} \right] = \frac{\dot{m}_{fuel} \cdot LHV_{fuel}}{\dot{m}_{clk}}$$

- Direct specific CO_2 emissions at cement kiln stack[5]:

$$e_{clk} \left[\frac{kg_{CO2}}{t_{clk}} \right] = \frac{\dot{m}_{CO2}}{\dot{m}_{clk}}$$

In addition to these values, cement manufacturing brings some additional indirect contributions associated to electric power import, reflecting in indirect primary energy consumption ($q_{el,clk}$) and CO_2 emissions ($e_{el,clk}$). These values depend on the specific electric power consumed by the cement kiln auxiliaries ($P_{el,clk}$), which is characterized by the efficiency (η_{el}) and specific CO_2 emissions (e_{el}) of the power generation technology or the country energy mix considered. Such indirect consumptions and emissions are defined as follows:

- Specific electric power consumption:

$$P_{el,clk} \left[\frac{kWh_{el}}{kg_{clk}} \right] = \frac{P_{el}}{\dot{m}_{clk}}$$

- Indirect specific primary energy consumption:

$$q_{el,clk} \left[\frac{kJ_{LHV}}{kg_{clk}} \right] = \frac{P_{el,clk}}{\eta_{el}}$$

---

[5] Only fossil carbon emissions should be accounted for in computing this index (i.e. CO_2 emissions from biomass combustion are considered as neutral emissions)

- Indirect CO_2 emissions:

$$e_{el,clk} \left[\frac{kg_{CO2}}{t_{clk}}\right] = P_{el,clk} \cdot e_{el}$$

As a result, equivalent values for primary energy consumption ($q_{clk,eq}$) and CO_2 emission ($e_{clk,eq}$) result from the sum of the direct and indirect values mentioned above:

$$q_{clk,eq} = q_{clk} + q_{el,clk} \quad \left[\frac{kJ_{LHV}}{kg_{clk}}\right]$$

$$e_{clk,eq} = e_{clk} + e_{el,clk} \quad \left[\frac{kg_{CO2}}{t_{clk}}\right]$$

It has to be highlighted that in case of net power export from the cement kiln ($P_{el} < 0$ according to its definition above) when a sufficiently large heat recovery steam cycle is possible, fuel and CO_2 credits accounted with the previous equations result in negative indirect fuel consumptions and CO_2 emissions.

CO_2 capture ratio is defined as the ratio between the CO_2 captured $\dot{m}_{CO2,capt}$ and the total CO_2 generated in the cement kiln $\dot{m}_{CO2,gen}$:

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

CO_2 avoided from flue gas (AC_{fg}) is defined as the direct CO_2 emission reduction from the flue gas of the cement kiln. In this definition, CO_2 emitted with the vent gas from the CO_2 purification unit should be accounted for:

$$AC_{fg} = 1 - \frac{e_{clk,fg}}{e_{clk,fg,ref}}$$

CO_2 avoided gives a better indication on the reduced impact of a cement kiln with CO_2 capture. CO_2 avoided is defined as follows, through the ratio between the CO_2 emission of the cement kiln with CO_2 capture e_{clk} and the emissions of the reference cement kiln without capture $e_{clk,ref}$:

$$AC = 1 - \frac{e_{clk}}{e_{clk,ref}}$$

Similarly, the equivalent CO_2 avoided index, taking into account the indirect CO_2 emissions associated to electric power consumption/generation can be defined through the equivalent CO_2 emissions:

$$AC_{eq} = 1 - \frac{e_{clk,eq}}{e_{clk,eq,ref}}$$

In order to compare different technologies from an energy and environmental point of view, the SPECCA (Specific Primary Energy Consumption for CO_2 Avoided) index can be used, as already done diffusely in the literature in power plants applications. The SPECCA index is defined by the following equation, quantifying the increased equivalent fuel consumption to avoid the emission of CO_2 in a cement kiln with CO_2 capture with respect to a reference cement kiln without capture (ref):

$$SPECCA\left[\frac{MJ_{LHV}}{kg_{CO2}}\right] = \frac{q_{clk,eq} - q_{clk,eq,ref}}{e_{clk,eq,ref} - e_{clk,eq}}$$

One advantage of the SPECCA index is to allow a comparison between CO_2 capture technologies, independently of the field of application. For example, SPECCA indexes calculated for cement kilns can be compared with the same indexes calculated for power plants and other industrial applications.

In Figure 10.1 a graphical representation of the connections between the cement plant and the electric grid with the performance indexes described above is shown.
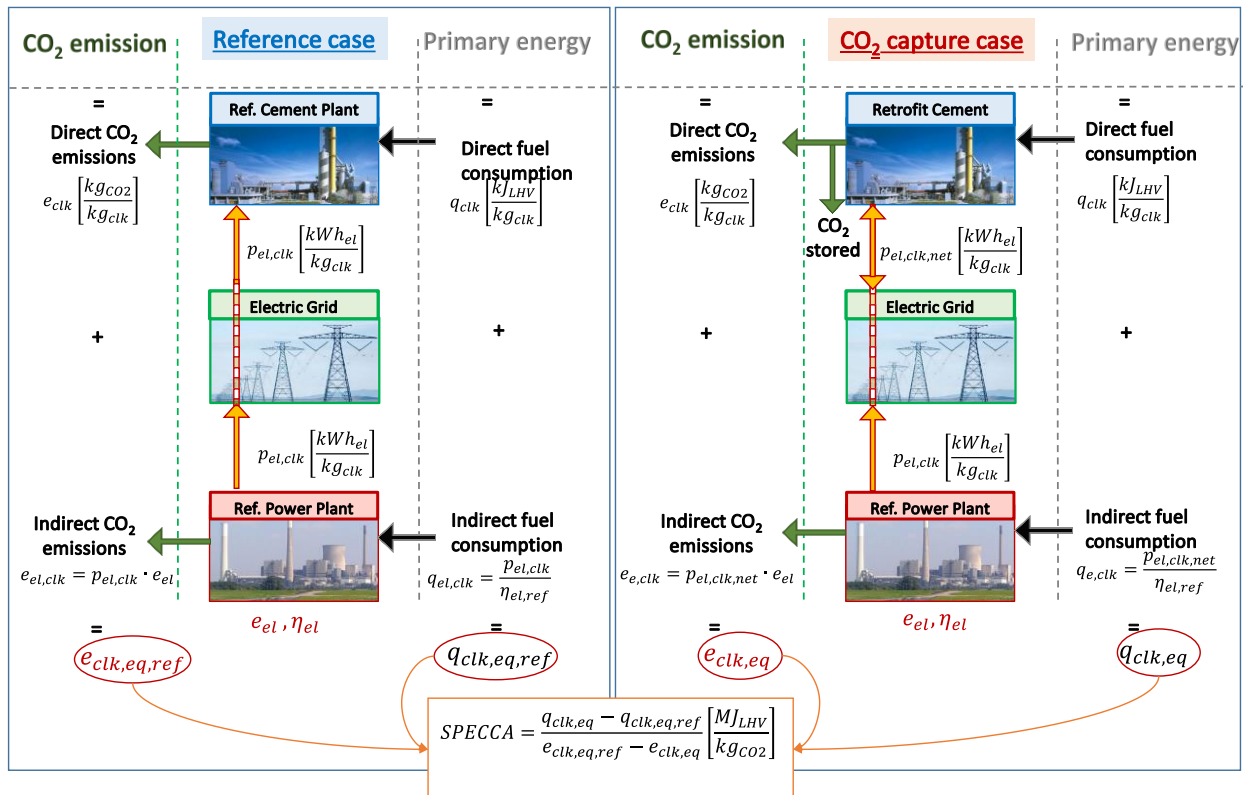

Figure 10.1: Representation of the energy flows involving a cement plant in reference cement plants and in plants with CO_2 capture

10.2   Economic indicators

The economic performance of the cement plant with CO₂ capture will be evaluated with the following indexes:

- Cost of clinker (COC): this is evaluated by summing the contributions of the investment cost C_{inv}, 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}, of the other O&M cost $C_{O\&M}$, all referred to the 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) has to be considered.

$$COC = C_{inv} + C_{Fuel} + C_{RM} + C_{el} + C_{O\&M} \quad \left[\frac{€}{t_{clk}}\right]$$

- Cost of CO₂ avoided (CAC): this 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 CO₂ capture:

$$CAC = \frac{COC - COC_{ref}}{e_{clk,eq,ref} - e_{clk,eq}} \quad \left[\frac{€}{t_{CO2}}\right]$$

- Cost of clinker with carbon tax (COC_{Ctax}): in order to compare the cost of clinker for cement kilns with different CO₂ capture rates, it is important to evaluate the economic impact of the CO₂ emitted on the cost of clinker. This cost can be calculated with the following equation, where the cost of the CO₂ emitted as consequence of the carbon tax (C_{Ctax}) is added to the previous COC.

$$COC_{Ctax} = C_{inv} + C_{Fuel} + C_{RM} + C_{el} + C_{O\&M} + C_{Ctax} \quad \left[\frac{€}{t_{clk}}\right]$$

where the following equation is used to calculate the C_{Ctax} term, as consequence of the carbon tax c_{tax}, expressed as € per ton of fossil CO₂ emitted:

$$C_{Ctax} = c_{tax} \cdot e_{clk,eq} = c_{tax} \cdot (e_{el,clk} + e_{clk}) \quad \left[\frac{€}{t_{clk}}\right]$$

The cost associated to CO₂ emissions takes into account the equivalent CO₂ emissions (i.e. both direct and indirect emissions). Based on this approach, the cost of electricity C_{el} to be used in the two COC indexes does not change. The increase of the expected cost of electricity as consequence of the presence of a carbon tax in the economic system is however taken into account through the contribution of the term $c_{tax} \cdot e_{el,clk}$.

10.3   Other indicators

Besides carbon dioxide emissions, the cement kiln exhaust gases contain other contaminants such as SO$_x$, NO$_x$, particulate, etc. released from fuel combustion or by the raw meal. Their final concentration may depend on the carbon capture technology considered. It is expected that reliable predictions on the emissions of such contaminants may not be completely clarified by

the activities performed in CEMCAP, which focus on CO_2 emissions. For such contaminants, qualitative considerations may be made for each technology assessed.

Other indicators that should be indicated for the technologies assessed are related to space requirement, in particular the footprint and the total height of the capture plant.

11 CASE OVERVIEW

Several different cases will be studied in CEMCAP in addition to the base case. The cases are related to the amount of air leak, the extent of CO_2 capture, the transport option, and the steam scenario. The cases are described under the relevant topic in this document, and an overview of the cases with a reference to the relevant chapter is given in Table 11.1.

Table 11.1: Overview of cases studied in CEMCAP.

Case	Air leak	CO_2 avoided from flue gas (AC_{fg})	Transport	Steam scenario	Relevant chapter
Base case	Increasing	90%	Pipeline	NG boiler and waste heat	
Constant low air leak	Constant low	90%	Pipeline	NG boiler and waste heat	3.8
Optional AC_{fg}	Increasing	Optional	Pipeline	NG boiler and waste heat	7
Ship transport	Increasing	90%	Ship	NG boiler and waste heat	8
Steam import	Increasing	90%	Pipeline	Steam import and waste heat	6.1

Other cases may be studied if they appear to be interesting or particularly relevant for one or several of the studied technologies.

12   SENSITIVITY ANALYSIS

Due to the intrinsic uncertainty and the time-place dependency of some assumptions, a sensitivity analysis should be performed on the following parameters in the suggested range, estimating the consequent techno-economic indicators variation:

- Indirect fuel consumption, indirect emission and SPECCA should be calculated with the efficiency and specific emissions of the power generation technologies indicated in Table 6.5, in addition to the values average European mix of Table 6.4.
- Fuel price: +/- 50% of the reference cost
- Electricity price: +/- 50% of the reference cost
- Steam supply: +/- 50% of the reference cost
- Carbon tax: 0-100 €/t$_{CO2}$
- Capex of the new technologies: +/- 30%

APPENDIX

APPENDIX A – THERMODYNAMIC PROPERTIES

A.1 Air and flue gases

Any EOS suitable for expanded gases, including the ideal gas EOS can be used for temperatures above 0 °C.

A.2 CO_2

Prediction of properties for CO_2 rich mixtures should be consistent with GERG-2008 [KUN, 2008]. This equation of state covers the gas phase, liquid phase, supercritical region, and vapour-liquid equilibrium stated for mixtures of 21 different components typically found in natural gas. In this model, pure CO_2 is modelled by means of the Span and Wagner (1996) equation of state, which is considered state-of-the-art for this component.

A.3 Reaction enthalpies during the production of cement clinker

Table A.3.1: Kiln feed reactions and reaction enthalpies during the production of Portland cement clinker [LOC, 2006]

Reaction	Reaction equation	Standard enthalpy of reaction	
		[kJ/kg] *	[kJ/kg clk]
I. Formation of the oxides, and decomposition reactions			
1. Evaporation of H_2O	$H_2O_{liq} \rightarrow H_2O_{vap}$	+ 2,453	+ 4
2. Kaolinite decomposition **	$Al_2O_3 \cdot 2\ SiO_2 \cdot 2\ H_2O \rightarrow Al_2O_3 + 2\ SiO_2 + 2\ H_2O$	+ 780	+ 78
3. Organic clay constituents	$C + O_2 \rightarrow CO_2$	- 33,913	-136
4. $MgCO_3$ dissociation	$MgCO_3 \rightarrow MgO + CO_2$	+ 1,395	+ 22
5. $CaCO_3$ dissociation	$CaCO_3 \rightarrow CaO + CO_2$	+ 1,780	+ 2,111
		total 1 – 5	+ 2,079
II. Formation of intermediate products			
6. Formation of CA	$CaO + Al_2O_3 \rightarrow CA$	- 110	- 8
7. Formation of C_2F	$2\ CaO + Fe_2O_3 \rightarrow C_2F$	- 114	- 6
8. Formation of β-C_2S	$2\ CaO + SiO_2 \rightarrow$ β-C_2S	- 732	- 493
		total 6 - 8	- 507
III. Sintering reactions			
9. Formation of C_4AF	$CA + C_2F + CaO \rightarrow C_4AF$	+ 25	+ 3
10. Formation of C_3A	$CA + 2\ CaO \rightarrow C_3A$	+ 25	+ 1
11. Formation of C_3S	β-$C_2S + CaO \rightarrow C_3S$	+ 59	+ 35
		total 9 to 11	+ 39
IV. Overall reaction a) Including combustion b) Without combustion	kiln meal → clinker	total of 1 - 11 total of 1,2,4 - 11	+ 1,611 + 1,747

* related to substance in left-hand column
** between 500 and 600 °C in the preheater

REFERENCES

[ABA, 2003]   J.C. Abanades, et al.: Conversion Limits in the Reaction of CO_2 with Lime. Energy Fuels, 17(2), p. 308-315 (2003)

[ABA, 2004]   J.C. Abanades, et al.: Capture of CO_2 from combustion gases in a fluidized bed of CaO. AIChE J., 50(7), p. 1614-1622 (2004)

[ALO, 2010]   M. Alonso, et al.: Carbon dioxide capture from combustion flue gases with a calcium oxide chemical loop. Experimental results and process development. Int. J. Greenhouse Gas Control, 4(2), p. 167-173 (2010)

[ALS]   Alstom: Industrial steam turbines – Sengkang (Indonesia) case study (leaflet)

[ANA, 2011]   R. Anantharaman, O. Bolland, N. Booth, E.V. Dorst, C. Ekstrom, F. Franco, E. Macchi, G. Manzolini, D. Nikolic, A. Pfeffer, M. Prins, S. Rezvani, L. Robinson: D1.4.3 European best practice guidelines for assessment of CO_2 capture technologies. In: DECARBit Project, 2011

[ANA, 2014]   R. Anantharaman, D. Berstad, S. Roussanaly: Techno-economic Performance of a Hybrid Membrane – Liquefaction Process for Post-combustion CO_2. Energy Procedia, Volume 61, p. 1244–1247 (2014)

[ANS]   Ansaldo Energia Steam Turbines brochure

[ARI, 2013]   B. Arias, et al.: Demonstration of steady state CO_2 capture in a 1.7 MWth Calcium looping pilot. Int J Greenhouse Gas Control, 245, p. 237-245 (2013)

[ASP, 2006]   A. Aspelund et al.: Ship Transport of CO_2: Technical Solutions and Analysis of Costs, Energy Utilization, Exergy Efficiency and CO_2 Emissions. Chemical Engineering Research and Design, 09/2006, p. 847-855

[ASP, 2007]   A. Aspelund and K. Jordal: Gas conditioning – The interface between CO_2 capture and transport. 2007

[BIP, 2006]   Bureau International des Poids et Mesures: The International System of Units (SI). 8th edition, 2006

[BOU, 2013]   K. Stéphenne: Start-Up of World's First Commercial Post-Combustion Coal Fired CCS Project: Contribution of Shell Cansolv to SaskPower Boundary Dam ICCS Project. Energy Procedia, Volume 63, 2014, p. 6106-6110

[BREF, 2013]   JRC Reference Reports - Best Available Techniques (BAT) Reference Document for the Production of Cement Lime and Magnesium Oxide. European Commission, 2013

[CEP, 2014]   Chemical Engineering, Economic Indicators: Chemical Engineering Plant Cost Index (CEPCI). 2014

[CHA, 2011]   A. Charitos, et al.: Experimental Validation of the Calcium Looping CO_2 Capture Process with Two Circulating Fluidized Bed Carbonator Reactors. Ind Eng Chem Res, **50**(16), p. 9685-9695 (2011)

[CON, 2012]	S. Consonni, F. Viganò: Waste gasification vs. conventional Waste-To-Energy: A comparative evaluation of two commercial technologies. Waste Management, 32, p. 653-666 (2012)
[COO, 2017]	CoolPack. Software. Available online at: http://www.en.ipu.dk/Indhold/refrigeration-and-energy-technology/coolpack.aspx
[DIE, 2014]	H. Dieter, et al.: Development of the calcium looping CO_2 capture technology from lab to pilot scale at IFK, University of Stuttgart. Fuel, **217**, p. 23-37 (2014)
[EBTF, 2011]	EBTF: European best practice guidelines for assessment of CO_2 capture technologies. 2011
[ECR, 2009]	ECRA CCS Project – Report about Phase II (TR-ECRA-106/2009). European Cement Research Academy, 2009
[EUD, 2004]	European Union – L32/183: establishing harmonised efficiency reference values for separate production of electricity and heat in application of Directive 2004/8/EC of the European Parliament and of the Council – Official Journal of the European Union
[EUR1]	Statistical Office of the European Union (Eurostat) - Supply, transformation, consumption - electricity - annual data
[EUR2]	Statistical Office of the European Union (Eurostat) - Electricity prices components for industrial consumers - annual data
[EUR, 2017]	Eurostat: Statistics Explained – Calculation methodologies for the share of renewables in energy consumption. http://ec.europa.eu/eurostat/statistics-explained/index.php/Calculation_methodologies_for_the_share_of_renewables_in_energy_consumption [Accessed: 14.09.17]
[FRA, 2009]	Franco et al. "Common Framework Definition Document". 2009
[GAL, 2006]	E. Gal: Ultra cleaning of combustion gas including the removal of CO_2. Patent WO2006022885, USA, 2006
[GAL, 2008]	E. Gal: Ultra Cleaning of Combustion Gas Including the Removal of CO_2. Patent US2008/0072762A1, USA, 2008
[GAL, 2011]	E. Gal, et al.: Chilled ammonia based CO_2 capture system with water wash system. Patent US2011/0048239A1, USA, 2011
[GÖT, 2004]	G. Göttlichter: The energetics of carbon dioxide capture in power plants. National Energy Technology Laboratory, 2004
[GRA, 2006]	G. Grasa, et al.: CO_2 Capture Capacity of CaO in Long Series of Carbonation/Calcination Cycles. Ind Eng Chem Res, 45(26), p. 8846-8851 (2006)
[GSS]	GS software, GECoS, Politecnico di Milano. www.gecos.polimi.it/software/gs.php
[GTW, 2015]	Gas Turbine World 2015 Performance Spec

[IEA, 2005]	CO₂ emissions from fuel combustion. International Energy Agency (IEA), Paris, France, 2005.

[IEA, 2005] CO_2 emissions from fuel combustion. International Energy Agency (IEA), Paris, France, 2005.

[IEA, 2008]   IEAGHG: CO_2 Capture in the Cement Industry (Report No.: 2008/3). IEA Greenhouse Gas R&D Programme, 2008

[IEA, 2013]   IEAGHG. Deployment of CCS in the cement industry (Report 2013/19). EA Greenhouse Gas R&D Programme, 2013.

[IEA, 2014]   IEAGHG: Assessment of emerging CO_2 capture technologies and their potential to reduce costs (Report No.: 2014/4). IEA Greenhouse Gas R&D Programme, 2008

[IEA, 2016]   IEAGHG: CO_2 Capture in Natural Gas Production by Adsorption Processes for CO_2 Storage, EOR and EGR (Report No.: 2016/XX). IEA Greenhouse Gas R&D Programme, 2016[IPCC, 2005]   IPCC Special Report on Carbon Dioxide Capture and Storage. 2005

[IEA2, 2016]   Statistic: Electricity Information. International Energy Agency (IEA), 2016

[IEA, 2017]   IEA: Frequently Asked Questions. https://www.iea.org/statistics/resources/questionnaires/faq/ [Accessed: 14.09.17]

[IUP, 1996]   IUPAC: Terminology for membranes and membrane processes. Pure & Appl. Chem., Vol. 68, No. 7, p. 1479-1489 (1996)

[JAC, 2009]   J.A. Jacobs, M. Schneider: Cogeneration application considerations. GE Energy report, 2009. Available online at http://site.geenergy.com/prod_serv/products/tech_docs/en/downloads/GER3430G.pdf

[KOR, 2013]   K. Koring, V. Hoenig, H. Hoppe, J. Horsch, C. Suchak, V. Klevenz, B. Emberger: Deployment of CCS in the Cement industry. In: IEAGHG (Ed.), 2013

[KOZ, 2009]   F. Kozak, et al.: Chilled ammonia process for CO_2 capture. Energy Procedia, **1**(1), p. 1419-1426 (2009)

[KUN, 2008]   O. Kunz, W. Wagner: The GERG-2008 Wide-Range Equation of State for Natural Gases and Other Mixtures: An expansion of GERG-2004

[LIN, 2014]   K. Lindqvist, K. Jordal, G. Haugen, K.A. Hoff, R. Anantharaman: Integration aspects of reactive absorption for post-combustion CO_2 capture from NGCC (natural gas combined cycle) power plants. Energy, 78, p 758-767 (2014)

[LOC, 2006]   F.W. Locher: Cement – pronciples of production and use. Verlag Bau + Technik, Düsseldorf, 2006

[MAR1, 2012]   M. Marchi, et al.: Improved process for the production of cement clinker and related apparatus (in Italian). Patent application MI2012 A00382, Italy, 2012

[MAR2, 2012]   M. Marchi, et al.: Process and improved plant for the production of cement clinker (in Italian). Patent application MI2012 A00383, Italy, 2012

[MEL, 2007]   T. Melin, R. Rautenbach: Membranverfahren – Grundlagen, Verfahren und industrielle Anwendungen, Wiley-VCH Verlag, 2007

[NET, 2011]   NETL: "Quality guidelines for energy system studies: Cost estimation methodology for NETL assessments of power plant performance". 2011

[MÜL, 1993]   D. Müller: "Erstellung eines modellhaften Wärmenutzungskonzepts an einer Anlage zur Zementproduction". Master thesis. 1993

[NET1, 2012]   NETL: "Quality Guidelines for Energy System Studies: CO_2 Impurity Design Parameters". 2012

[NET2, 2012]   NETL: "Quality guidelines for energy system studies: Technology learning curve (FOAK to NOAK)". 2012

[OFX, 2017]   OFX: "Historical Exchange Rates", 2017. Webpage. https://www.ofx.com/en-gb/forex-news/historical-exchange-rates/ [accessed 13.03.17].

[ROB, 2008]   L.M. Robeson: The upper bound revisited. Journal of Membrane Science, 320(1–2), p. 390-400 (2008)

[ROD, 2008]   N. Rodríguez, et al.: Process for Capturing CO_2 Arising from the Calcination of the $CaCO_3$ Used in Cement Manufacture. Environ Sci Technol, **42**(18), p. 6980-6984 (2008)

[ROD, 2012]   N. Rodríguez, et al.: CO_2 Capture from Cement Plants Using Oxyfired Precalcination and/or Calcium Looping. Environ Sci Technol, **46**(4), p. 2460-2466 (2012)

[ROM, 2013]   M.C. Romano, et al., The Calcium Looping Process for Low CO_2 Emission Cement and Power. Energy Procedia, **37**, p. 7091-7099 (2013)

[ROM, 2014]   M.C. Romano, et al.: The Calcium looping process for low CO_2 emission cement plants. The 6th International Conference on Applied Energy. 2014: Taipei, Taiwan

[SPL, 2010]   H. Spliethoff: Power Generation from Solid Fuels. Springer-Verlag, Berlin Heidelberg, 2010

[TEL, 2012]   V. Telikapalli, et al.: CCS with the Alstom chilled ammonia process development program–Field pilot results. Energy Procedia, **4**, p. 273-281 (2011)

[TSC, 2003]   I. Tschanun, M. Mineur: Biomass Combustion With State Of The Art Bubbling Bed Steam Generators. Power-Gen Europe 2003

[VDI, 2003]   VDI 2094: Emission Control Cement Plants. VDI-Handbuch Reinhaltung der Luft, Band 3, 2003

[WER, 2014]   M. Werner et al.: Flue gas CO_2 mineralization using thermally activated serpentine: from single- to double-step carbonation. Phys. Chem. Chem. Phys., 16, p. 24978-24993 (2014);
DOI: 10.1039/C4CP02786H