CEMCAP

Co-funded by the European Union

Grant Agreement Number:
641185

Action acronym:
CEMCAP

Action full title:
CO_2 capture from cement production

Type of action:

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

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

D4.4
Cost of critical components in CO_2 capture processes

Due delivery date: 2018-08-31
Actual delivery date: 2018-10-20

Organization name of lead participant for this deliverable:
C.T.G. SPA

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:	D4.4
Deliverable title:	Cost of critical components in CO_2 capture processes
Work package:	WP4: Comparative capture process analysis
Lead participant:	CTG

Author(s)		
Name	**Organisation**	**E-mail**
Rahul Anantharaman	SINTEF-ER	rahul.anantharaman@sintef.no
Giovanni Cinti*	C.T.G. SPA	g.cinti@itcgr.net
Edoardo De Lena	Politecnico di Milano	edoardo.delena@polimi.it
Chao Fu	SINTEF-ER	chao.fu@sintef.no
Stefania Osk Gardarsdottir	SINTEF-ER	stefania.gardarsdottir@sintef.no
Helmut Hoppe	VDZ GmbH	helmut.hoppe@vdz-online.de
Armin Jamali	VDZ GmbH	armin.jamali@vdz-online.de
Matteo Romano	Politecnico di Milano	matteo.romano@polimi.it
Simon Roussanaly	SINTEF-ER	simon.roussanaly@sintef.no
Maurizio Spinelli	Politecnico di Milano	maurizio.spinelli@polimi.it
Olaf Stallmann[1]	GE-SE	olaf.stallmann@bhge.com
Mari Voldsund	SINTEF-ER	mari.voldsund@sintef.no

*Lead author

[1]Current affiliation Baker Hughes, a GE company.

Keywords

CO_2 capture, CCS, cement, cost estimation, CAPEX

Abstract

In this deliverable the methodology for estimation of equipment costs and installation costs of critical process components is established. A brief overview of the costing method that will be used for standard process components is given, before more detailed descriptions of costing methodologies and assumptions are given for each of the non-standard components. The costing of the non-standard components is either based on experience and databases of the CEMCAP partners, or on information available in the literature.

Please cite this report as: *Cinti, Giovanni; Anantharaman, Rahul; De Lena, Edoardo; Fu, Chao; Gardarsdottir, Stefania Osk; Hoppe, Helmut; Jamali, Armin; Romano, Matteo; Roussanaly, Simon; Spinelli, Maurizio; Stallmann, Olaf; Voldsund, Mari; 2018. Cost of critical components in CO_2 capture processes (D4.4).*
Refer to the CEMCAP community in Zenodo.org for citation with DOI.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

TABLE OF CONTENTS

Page

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

1 INTRODUCTION

The purpose of this deliverable is to establish the methodology for calculation of cost of process components (equipment and installation costs) in CEMCAP, and thus give a basis for techno-economic evaluations to be presented in CEMCAP deliverable D4.6.

The title of this deliverable is "Cost of critical components in CO_2 capture processes". The word *critical* is here used in the context *critical for the cost estimations*. Thus, the critical components are the process components where cost evaluations are not straightforward, and a detailed description of the cost estimation methodology is required.

The methods for cost estimation of the critical components are based on literature or developed in discussion with the industry partners. In particular, VDZ has provided information on the DeNOx system, ThyssenKrupp and Italcementi have been involved in the chapter about the main burner and IKN has provided cost estimates for the oxyfuel clinker cooler and the oxyfuel kiln hood. The cost methodologies presented for oxyfuel auxiliary burners, oxyfuel calciner, fans, rotary kiln inlet and outlet sealings, exhaust gas recirculation system, calcium looping reactors, and the preheating tower are developed based on the experience and internal databases of Italcementi. Reference cost and size for the liquid CO_2 pump is based on data compiled from industrial references in the COCATE EU FP7 project.

A short overview of the cost estimation methodology is given in Chapter 2, an overview of "standard" process equipment and the key characteristics required for cost estimation is given in Chapter 3, and design criteria are presented in Chapter 4. In Chapters 5 – 27 cost estimation methodologies for each of the critical components are described. Finally, the estimation of capital costs for the chilled ammonia process, which is preformed by GE, is described in Chapter 28.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

2 COST ESTIMATION IN CEMCAP

The general method for cost estimation in CEMCAP is explained in the CEMCAP framework document (D3.2). A summary of the method showing the different cost elements that are included is shown for the total plant costs in Figure 2.1.

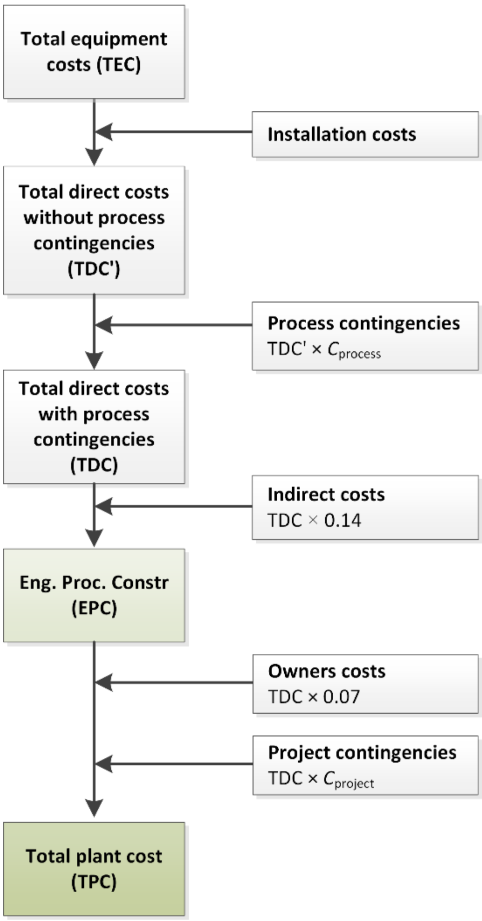

Figure 2.1. The Bottom-Up approach for estimation of total plant costs.

An overview of the cost elements included in the operational costs is given in Figure 2.2. For some equipment, e.g. the $DeNO_x$ and $DeSO_x$ systems, it is relevant to present estimation of consumable consumption/costs in addition to the equipment costs in this document.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

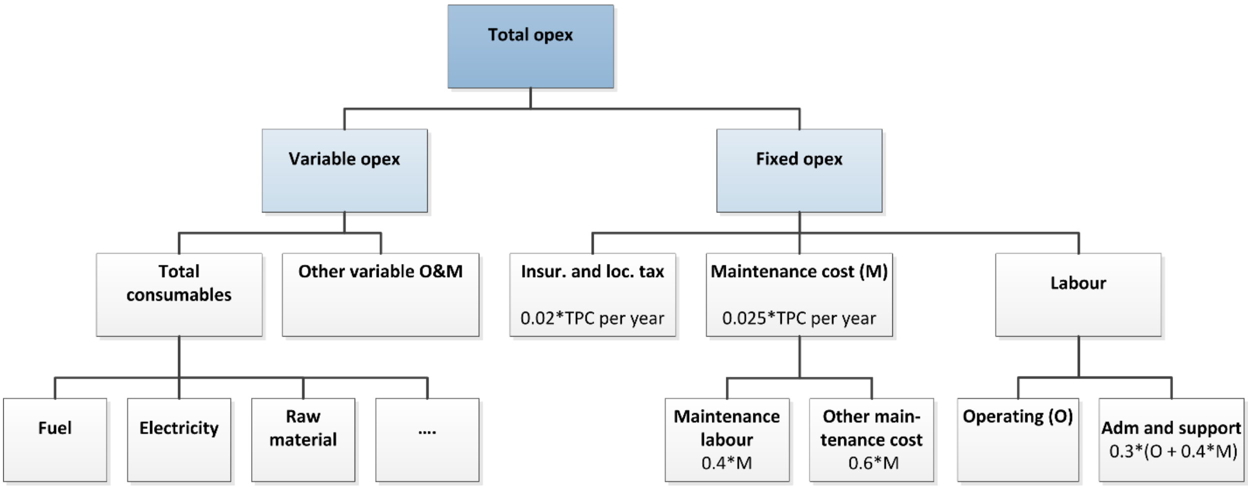

Figure 2.2. Overview of the elements of operational expenses.

All costs estimations will be performed in euro on a 2014 basis. In cases where costs are not directly available in 2014 prices, capital costs are adjusted to the correct level using the Chemical Engineering Plant Cost Index (CEPCI, Table 2.1), and operational costs are adjusted for inflation. The CEPCI values for recent years are given in Conversion rates between euro, US dollars and GB pounds for recent years can be found the same place. In this document, all costs are given in 2014 values unless something else is specified.

Table 2.1. Yearly average Chemical Engineering Plant Cost Index [CEP, 2017].

Year	CEPCI
2001	394.3
2002	395.6
2003	402.0
2004	444.2
2005	468.2
2006	499.6
2007	525.4
2008	575.4
2009	521.9
2010	550.8
2011	585.7
2012	584.6
2013	567.3
2014	576.1
2015	556.8
2016	541.7
2017	567.5

The cost estimates will be for 'nth plants' based on current knowledge of the technology, i.e. for commercial plants built after the initial technology demonstration plants. Additional costs normally associated with 1st-of-a-kind commercial plants shall be excluded. The estimate accuracy is expected to be +35%/-15% (AACE Class 4) for all technologies, except for the chilled ammonia process where all cost estimation will be done by GE with expected uncertainty of ±30% (also AACE Class 4).

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

3 "STANDARD" PROCESS EQUIPMENT

For all capture and conditioning technologies, the direct cost of "standard" equipment (such as pumps, heat exchangers, packed columns, vessels, compressors, expanders, blowers, electric generator, vacuum pump, amine filter) is assessed using Aspen Process Economic Analyzer® (APEA). In such cases, the cost evaluation is based on the key characteristics of the equipment based on the process simulation and design criteria. A list of the input key characteristics required for the cost assessment in APEA is presented in Table 3.1.

However, it is worth noting that in the case of the chilled ammonia process, GE will do a cost estimation based on GE's in-house tools and following the cost assessment guideline established in D3.2.

Table 3.1. List of input key characteristics for cost evaluation of the different equipment in APEA.

Equipment type	Key characteristics input
Columns	Type
	Size – internal diameter
	Size – height
	Design pressure
	Design temperature
	Material
Heat exchangers	Type
	Size – duty
	Size – area
	Design pressure
	Design temperature
	Material
Fans, compressors and expanders	Type
	Size – flow
	Size – power
	Design pressure
	Design temperature
	Material
Pumps (above atmospheric pressure)	Type
	Size – flow
	Design pressure
	Design temperature
	Material
Vacuum pumps	Type
	Size – flow
	Size – power
	Design temperature
	Material
Tanks and vessels	Type
	Size – internal diameter
	Size – height
	Design pressure
	Design temperature

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

	Material
Amine filter	Type
	Size – flow
	Design pressure
	Material
Electrostatic precipitator	Type
	Size - flow
	Material
Electric generators	Type
	Size – power
	Voltage
Cooling tower	Type
	Size - flow
	Cooling range
	Temperature approach

4 DESIGN CRITERIA

This subchapter presents design criteria for process equipment. This information is needed as input to APEA for cost estimation of standard process equipment and is also relevant for size estimation of the process equipment required in the CO_2 capture units in general. They reflect the general criteria normally adopted for the design of process plants [IEA, 2017]. Exceptions are possible, but they shall be minimized and justified.

4.1 Design pressure

The design pressure is the max. and/or min. pressure for which the equipment is designed. Design pressure is calculated from the operating pressure as given in Table 4.1.

Table 4.1. Design pressure related to operating pressure. Exceptions are listed in the text below.

Pressure range	**Operating pressure**	**Design pressure**
Below atmospheric	Below 1 bara	Not relevant (not required for cost estimation)
Above atmospheric	0 – 12.0 barg	max. norm. operating pressure + 1.8 bar
	12.0 – 21.1 barg	max. norm. operating pressure × 1.15
	21.1 – 31.6 barg	max. norm. operating pressure + 3.2 bar
	31.6 – 70 barg	max. norm. operating pressure × 1.10
	>70 barg	special design

Exceptions to the design pressure given in Table 4.1 are:

- Design pressure defined in accordance with the above criteria does not include liquid static head.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

- For equipment handling liquid, maximum normal operating pressure used for calculation of design pressure shall not be lower than maximum vapour pressure of the liquid at ambient temperature.

- Design pressure for storage vessels for liquids with boiling point lower than 45 °C shall not be less than vapour pressure at 50 °C.

- Pumps should be designed to handle static head and pressure drop from relevant downstream instrumentation and process units.

- Heat exchangers shall be designed by following the "10/13 rule" in accordance with the latest edition of ASME VIII Div.1 [ASM, 2015]. This means to set a design pressure of the lower pressure side of a heat exchanger not lower than 77% of the design pressure of the higher pressure side. This design rule allows to protect the equipment against tube rupture without installing a pressure safety valve. The design pressure set following the "10/13 rule" will be extended also to inlet and outlet piping connected to the lower pressure side of the heat exchanger including the isolation valves.

4.2 Design temperature

The design temperature is the maximum or minimum temperature value for which the equipment is designed.

The following summarizes the criteria to be followed to set the design temperature:

- Operation above ambient temperature: Design temperature is maximum normal operating temperature + 28°C.

- Operation below ambient temperature: Design temperature is minimum normal operating temperature – 5°C.

- When increased operating flexibility is required or future operating conditions are expected, the above design margin can be larger.

4.3 Material selection

Material selection for process equipment is an important factor, both from a material properties and cost perspective. Unless otherwise dictated for process reasons, Stainless Steel 304L is used as the default material for equipment in contact with liquid solvent and CO_2, while Stainless Steel 316L for equipment in contact with hot liquid solvent (above 80ºC) and CO_2 [IEA, 2017]. Stainless Steel 304L is also proven suitable in low-temperature applications [KEY, 2011]. For equipment operating in less severe conditions, e.g. flue gas fans, Carbon Steel can be used.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

4.4 Equipment design life and corrosion allowance

Unless otherwise dictated by process reasons, the corrosion allowance and design life will be specified in accordance with the following guidelines:

- Thick-walled Reactors and Vessel: Corrosion allowance for 25 yrs design life.

- Column & Vessel Carbon Steel and Alloy Construction (including non-removable internal): Corrosion allowance for 25 yrs design life.

- Heat exchanger
    - Shell of carbon steel and alloy construction (including non-removable internal): corrosion allowance for 25 yrs design life.
    - Bundle of stainless steel construction: corrosion allowance for 10 yrs design life.
    - Bundle of carbon steel and alloy steel construction: corrosion allowance for 5 yrs design life.

- Other Equipment and Removable Vessel Internals

  Carbon Steel Construction
  Corrosion allowance for 10 yrs design life.

  Alloy Steel Construction
  Corrosion allowance for 15 yrs design life.

  Stainless Steel Construction
  Corrosion allowance for 15 yrs design life.

4.5 Equipment sizing

The criteria of design oversizing factors to be used to determine the dimensions and/or rating capacity of equipment and machinery are listed herein below.

4.5.1 Drums

No oversizing is considered. Suitable hold-up times are considered according to the services the drums are specified for.

The following minimum liquid surge time shall be provided between low and high liquid level in new drums, unless otherwise dictated by process reason:

- Drums feeding other equipment for further processing: 15 min
- Overhead receivers: 5 min hold up shall be considered with respect to the reflux flow rate or 2 min with respect to the net product flow rate – whichever is the greater.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

The same sizing rules shall be applied also to hold-up capacity of columns.

4.5.2 Columns

Unless otherwise dictated for process reason, columns shall be sized in order to obtain a percent flooding not higher than 75% based on the maximum normal loading rate.

4.5.3 Pumps

Unless otherwise dictated for process reasons, the rated pump capacity, for new pumps, shall be the maximum normal flow-rate multiplied by the factors in Table 4.2.

Table 4.2. Oversizing factors for pump capacity.

Capacity/Control	Factor
Pumps capacity under flow control	1.10
Pumps capacity under level control or heat removal	1.20

4.5.4 Compressors

Unless otherwise dictated for process reasons, design oversizing factors for compressors shall be 1.10.

4.5.5 Drivers and generators

Unless other dictated for process reasons, motors shall have a power rating of 0.8 [AGR, 2001].

Steam and gas turbine drivers shall be sized to deliver continuously 110% of the maximum power required for the purchaser's specified conditions, while operating at corresponding speed with specified steam conditions.

4.5.6 Heat exchangers

Unless otherwise dictated for process reasons, a design margin of 10% shall be specified on required heat exchanger area for new coolers/condensers (including air) and reboilers.

In addition to the above requirements, for heat recovery systems (i.e. feed/effluent and feed/bottom) the following overdesign criteria shall also be met:

- Effluent cooler and feed heater: 10% of cooler or heater duty or 5% of feed/effluent duty, whichever is greater;
- Reboiler: 5% of feed/bottom duty or 10% of reboiler duty, whichever is greater;

Heat exchanger area is estimated using the following equation:

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

$$A = \frac{\dot{Q}_{des}}{U \cdot \Delta T_{lm}} \tag{4.1}$$

where Q_{des} (heat exchanger duty) and ΔT_{lm} (logarithmic mean temperature difference) are based on process simulations (considering design oversizing factors) and the average U-value (overall heat transfer coefficient) is assumed from Table 4.3.

Table 4.3. Assumed overall heat transfer coefficients (U-values).

Type of heat exchanger	$U_{assumed}$ ($Wm^{-2}K^{-1}$)
Reboiler, kettle type	1500
S&T condenser	1000
P&F condenser	3000
Gas-gas heat exchanger (Atmospheric pressure)	35
Gas (dirty) – liquid heat exchanger (Atmospheric pressure)	70
Gas (clean) – liquid heat exchanger (Atmospheric pressure)	400
Gas-gas heat exchangers (Above atmospheric pressure)	300
Gas-liquid heat exchangers (Above atmospheric pressure)	500
Liquid-liquid heat exchangers	1500

5 DENO$_X$ SYSTEM

The production of cement clinker is a high-temperature process (gas temperatures up to 2,000 °C and material temperatures above 1,400 °C) which results in the formation of nitrogen oxide. At most cement kilns, emission abatement measures have to be applied to meet the given NO_x emission limit value. The SNCR process (selective non-catalytic reduction) is a state-of-the-art technology in the cement industry and is applied in many cement kilns in Europe. The NO_x reduction is achieved by injection of a reducing agent into the flue gas stream, at a point in the process where the flue gas is between 900 and 1,050 °C. In most cases aqueous ammonia solution (25%) is used, but also urea solution (40%) is a common reducing agent. The reduction of NO_x is carried out according to the following chemical reaction:

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

In cement kilns the appropriate temperatures for injection of the ammonia solution are given in the riser duct above the kiln inlet or in the calciner. If the injection is carried out at lower temperatures emissions of unreacted ammonia can occur, the so-called "NH_3 slip". If the temperatures are too high the reducing agent can be oxidized which results in lower NO_x reduction rates and decreased efficiency of the SNCR process.

The available operating experiences show that NO_x reduction rates of 50-80% are achievable with the SNCR process. To achieve high reduction rates, the reducing agent has to be injected in overstochiometric amounts.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

The TDC for a standard SNCR plant are 0.5-1.0 M€$_{2017}$, in most cases closer to 1 M€$_{2017}$. The installation of a so-called "high efficiency SNCR" plant (he-SNCR) would require higher costs. The costs for the storage of the ammonia solution are included in these figures. The costs of ammonia solution (25%) are currently between 120 and 140 €$_{2017}$/t in central Europe.

A standard SNCR system, with a long-term average reduction rate of 60% and ammonia solution (25%) as reducing agent, is assumed to be already installed in the CEMCAP reference cement kiln (TDC of 1 M€$_{2017}$). The consumption of ammonia solution is assumed to be 5 kg/t_{clk}. The cost of increased NO_x removal (max 80% reduction rate) can be estimated as the increased reagent cost by assuming an ammonia consumption of 1.5 times the stoichiometric amount, and ammonia solution cost of 130 €$_{2017}$/t.

6 DESO$_X$ SYSTEM

DeSOx systems are normally not installed in conventional cement plants since most SOx are absorbed in the clinker and the raw mill [IEA, 2008]. However, when the MEA CO_2 capture process is included, the MEA can react with acid gases such as SOx and NOx to form amine salts. The SOx in the flue gas should typically be limited to 10 ppmv prior to the MEA CO_2 capture process. SOx should possibly also be removed prior to the membrane units in the membrane-assisted liquefaction process.

6.1 Scrubbing with NaOH

The reduction of SO_x to acceptable levels is achieved through injection of 50% NaOH solution in the direct contact cooler (DCC) present in both the MEA and the MAL processes. The cost of NaOH solution is taken as 370 € [ONA, 2017]. The reduction takes place according to the following chemical reaction, where all SO_x is assumed to be in the form of SO_2:

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

The SO_x concentration of the cement kiln flue gases is 200 mg/Nm^3 on dry basis at 10% O_2 content, which translates to 236 mg/Nm^3 during the first part of the year (in low air leak conditions), and 182 mg/Nm^3 during the second part of the year (in medium air leak conditions). All SO_x is assumed to be removed in the process, with stochiometric amount of the NaOH solution sufficient to achieve the reduction. An example of estimated costs of the DeSOx system for the low air leak MEA capture case is given in
Table 6.1.

Table 6.1. Example of estimated costs for SOx removal with NaOH scrubbing for MEA capture.

	Low air leak
Total direct costs (TDC), M€$_{2014}$	0
Total plant cost (TPC), M€$_{2014}$	0
Total operating cost (variable only), M€$_{2014}$/yr	0.27

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

6.2 Wet limestone FGD

The most common DeSOx technology in the power industry is the limestone based wet flue gas desulfurization (FGD) technology, which is included here for comparison with NaOH scrubbing.

The total plant cost and operating costs for this technology can be calculated based on the method of Sargent and Lundy [SAR, 2013] and the cost estimation methodology defined in CEMCAP, with sizing of equipment based on McCabe et al. [MCC, 2005], Joseph and Beachler [JOS, 1998] and IEA [IEA, 2014a]. An example of estimated costs of the DeSOx system for the low air leak MEA capture case is given in Table 6.2. The SOx at the outlet of the FGD has been limited to 10 ppmv (dry basis).

Table 6.2. Example of estimated costs for the wet limestone based DeSOx system for MEA capture.

	Low air leak
Total direct costs (TDC), $M€_{2014}$	19.7
Total plant cost (TPC), $M€_{2014}$	26.7
Total operating cost, $M€_{2014}$/yr	1.22
Variable opex	0.26
Fixed opex	0.96

7 THERMAL RECLAIMER UNIT

Thermal reclaimer units are estimated based on the cost estimates published by IEAGHG [IEA, 2014b]. The cost of the thermal reclaimer unit is in this case scaled based on the heat stable salt (HHS) mass flow going through the thermal reclaimer considering a linear total plant cost of 76,000 $\$_{2013}/(kg_{HHS}/h)$. 0.1% of the solvent circulation rate is assumed to be treated in the thermal reclaimer unit and the heat stable mass flow is assumed to represent 3% of the solvent mass flow [IEA, 2014]. The amount of solvent lost in the reclaimer waste stream is assumed to be 1.4 kg MEA/ton CO_2 captured [KNU, 2009]. The heat requirement is estimated to 10.4 kWh/kg_{HSS}.

8 MEMBRANE UNITS

The membrane module cost is estimated considering a membrane unitary cost of 40 $€_{2014}/m^2$ based on cost update and currency conversion of 50 $\$_{2010}/m^2$ adopted by Zhai and Rubin [ZHA, 2013]. The membrane framework is based on the cost function suggested by van der Sluijs et al. [SLU, 1992] for the framework of the membrane separation system in an ammonia plant of DSM, and modified by Roussanaly et al. [ROU, 2014; ROU, 2016] to take the influence of the design pressure of the module into account. The resulting equation for direct costs (DC) of the framework as function of membrane module area (A_{module}) and membrane module pressure (P_{module}) is shown in the equation below, and information about the reference module is given in Table 8.1. It is worth noting that a maximum limit of 25,000 m^2 of membrane area per module is considered in order to avoid unrealistically large modules and economics of scales.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

$$\mathrm{DC}_{\text{membrane framework}} = \left(\frac{A_{\text{module}}}{2000}\right)^{0.7} \cdot \mathrm{DC}_{\text{reference module}} \cdot \left(\frac{P_{\text{module}}}{55}\right)^{0.875} \tag{8.1}$$

Table 8.1. Direct cost of the membrane framework.

Type of equipment	Value	Unit	Reference
Reference module area	2 000	m^2	[SLU,1992]
Reference pressure	55	bar	[SLU, 1992]
Reference module cost	286	$k€_{2014}$	[SLU, 1992]

In estimating operating costs of the membrane module, an annual replacement rate of 20% of membrane area with a material replacement cost of 8 $€_{2014}/m^2$ is assumed, based on Zhai and Rubin [ZHA, 2013].

9 VACUUM PUMP

Especially large vacuum pumps are required in the membrane-assisted CO_2 liquefaction process. Their cost (TEC and TDC) is estimated by scaling the APEA derived unit cost of a pump with a relatively small maximum capacity of 6750 m^3/h, to the capacity of a large specialized industrial pump. The total cost for the pump stack is equal to the unit cost multiplied with the required number of identical pumps in the process. The reference industrial capacity of 30,000 m^3/h was derived from communications with industrial contacts.

$$Unit\ cost = \left(\frac{\dot{V}_{industry\ reference}}{\dot{V}_{max,APEA}}\right)^{0.6} \tag{9.1}$$

10 DEHYDRATION UNITS

Dehydration by molecular sieves and by TEG units are considered for dehydration in CEMCAP.

10.1 Molecular sieve

Molecular sieves are often utilized for drying gas streams. For example, in the liquefied natural gas (LNG) industry, the water content of the gas needs to be reduced to less than 1 ppmv to prevent blockages caused by solid formation. Molecular sieves are also suitable for drying of the CO_2 product stream.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

The cost of molecular sieve unit is calculated as the sum of the molecular sieve containers cost and the adsorbent cost. The adsorbent cost is evaluated based on the estimated adsorbent total mass and a cost of 3600[1] $€_{2015}/t_{adsorbent}$.

$$DC_{\text{mol.sieve}}\ (€) = DC_{\text{containers}} + 3600\ €_{2015}/\text{t} \cdot \text{m}_{\text{adsorbent}} \tag{10.1}$$

The molecular sieve containers cost is assessed by first estimating the size of two adsorption columns, and then using APEA for cost evaluation of pressure vessels to evaluate the cost. A brief overview of a method for sizing of the adsorption columns is provided below, based on Kohl & Nielsen [KOH, 1997]. The mass of adsorbent can be calculated from the adsorbent bulk density and the estimated bed length and cross-sectional area.

The steps required to size a dehydration adsorption column are:

1. Select the adsorbent material based on requirement and find values for design capacity and bulk density
2. Determine superficial gas velocity based on a specified pressure drop limitation
3. Calculate the required bed cross sectional area based on the superficial velocity calculated in step 2
4. Determine total bed length based on the volume of adsorbent calculated in step 4 and an estimated length of mass transfer zone
5. Check that the resulting overall pressure drop is reasonable
6. Calculate the adsorber column height based on the total bed length calculated in step 5

Step 1

The molecular sieve design capacity can be assumed to be 16.8 kg water per 100 kg adsorbent, and the bulk density 720 kg/m^3.

Steps 2 and 3

The pressure drop per unit length for molecular sieve adsorbents is given by a modification to the Ergun equation for molecular sieve adsorbents:

$$\Delta P / L = B\mu V + CdV^2 \tag{10.2}$$

where ΔP is the pressure drop across the bed (psi) and assume to be 0.33 psi/ft, L is the bed length (ft), μ is the gas viscosity (centipoise), V is the superficial gas velocity (ft/min) and d the gas density (lb/ft^3).

[1] Based on a quote of 1200 €/t (2015) and a logistic/construction factor of 3.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

B and C are constants that take into account the intrinsic properties of the bed. The typical values of B and C are given in Table 10.1.

Table 10.1. Values for B and C for different particle types.

Particle type	B	C
1/8 in. bead	0.0560	0.0000869
1/8 in. extrudate	0.0722	0.000124
1/16 in. bead	0.152	0.000136
1/16 in extrudate	0.238	0.00210

The cross-sectional area can be calculated from the volumetric gas flow rate and the superficial gas velocity.

Step 4

The total length of the bed is the sum of the length of the equilibrium section (LES) and the length of unused bed (LUB):

$$\text{Total length of bed} = LES + LUB \tag{10.3}$$

The LES is

$$LES = 100G/\rho_b \cdot (\Delta Y/\Delta X) \cdot t \tag{10.4}$$

where G is the gas feed rate (kmol/hr.m^2), ρ_b is the adsorbent bulk density (kg/m3), ΔX = Xe-Xo where Xe and Xo are the water loadings in equilibrium with the feed gas and on the regenerated adsorbent (mole per 100 kg) and this delta is equal to the sorbent design capacity, ΔY = Ye-Yo where Ye and Yo are the concentration of water in the gas feed and in the gas in equilibrium with the regenerated adsorbent (mole/mole) and t is the time from start of adsorption (hr).

The time of adsorption in one cycle is called the cycle time and it is normally around 4 hours.

LUB = ½MTZ where MTZ is the mass transfer zone or zone of active adsorption. MTZ is evaluated from the following equation:

$$MTZ = (V/35)^{0.3} \cdot Z \tag{10.5}$$

Where V is the superficial gas velocity in ft/min and Z = 1.7 for 1/8 in. sieve and 0.85 for 1/16 in sieve.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

Step 5

The overall pressure drop of the adsorbent bed is normally targeted to be in the range of 0.35-0.6 bar. The modified Ergun equation can be used to check if the overall pressure drop of the designed column is within this range.

Step 6

The total height of the column is the evaluated to be the length of the adsorbent bed + 2-3m to account for internals etc.

10.2 TEG process

As the equipment of the TEG are generally standard items, the cost estimation of this process is performed based on the designed equipment characteristics and the methodology presented in Chapter 3

11 REFRIGERATION SYSTEM

Refrigeration cycles should be designed specifically for each application. Such cycles consist typically of standard process equipment like heat exchangers and compressors, and costs can be calculated with APEA. If plate and fin heat exchangers are used, the method for cost evaluation is given in Chapter 25.

12 CO_2 PURIFICATION UNIT

CO_2 purification units (CPU) also consist mainly of standard process equipment, in addition to dehydration units and plate and fin heat exchangers. The cost of these units can be found using APEA and the methods described in Chapters 10 and 25.

13 CO_2 PUMP

The total equipment cost of liquid-CO_2 pumps used in CPU units and CO_2 compression section for pipeline transportation is estimated from the following equation, using the pump design capacity as a scaling factor. The reference cost values are derived from communications with several industrial contacts [BUR, 2011a and b].

$$TEC_{CO_2\,pump}[M€_{2011}] = 1.66 \cdot \left(\frac{P_{design}[kW]}{1200}\right)^{0.85} \tag{13.1}$$

Installation cost can be calculated as 2/3rd of the TEC.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

14 AIR SEPARATION UNIT

The oxyfuel and CaL processes rely on the use of high purity oxygen for combustion. Considering the relatively large amount of oxygen needed in these applications, the selected oxygen production technology is the cryogenic air separation unit (ASU) process.

The cost of ASU producing 95% pure O_2 at 1.5 bar can be estimated with the following equation, using the mass flow rate of pure oxygen produced as scaling parameter. This cost refers to the EPC (*cf.* Figure 2.1).

$$EPC_{ASU}\ [M€] = 69.8 \cdot \left(\frac{m_{O2}[tpd]}{2\ 717}\right)^{0.6} \tag{14.1}$$

This cost is established by removing an estimated O_2 compression cost from the EPC cost estimates presented by Rezvani et al. [REZ, 2009]: 75.8 M€$_{2008}$ for a 95% pure O_2 delivery pressure around 40 bar.

15 OXYFUEL MAIN BURNERS

The ongoing testing and modelling of oxyfuel burners within CEMCAP (WP7) may either show that modern burner designs are oxyfuel compatible, or that alteration of burner design is necessary.

If modern burner designs are shown to be oxyfuel compatible, meaning that a conventional burner can be used with a suitable gas matrix (premixed gas with <<100% oxygen), it is assumed that a reuse of the existing burner is allowed.

If it is shown that alteration of burner design is necessary, the cost of a new burner is assumed to be 100 k€$_{2017}$. This is without any additional parts such as primary air fan, igniter, flame detection, control station, piper, burner carriage, etc. The installation costs related to the oxyfuel main burner is negligible.

16 OXYFUEL CALCINER FUEL INJECTION

Fuel injection and combustion in a calciner may be performed either by injection of fuel with simple injection pipes at various locations in the calciner (in the case of solid fuel and no primary air), or by small burners (can handle several types of fuel and combustion with primary air). In the CEMCAP reference kiln, primary air is consumed in the calciner, meaning that fuel injection and combustion is performed by small burners. The fuel type considered is coal, but the use of burners gives the flexibility to use other fuels, and it facilitates better flame control.

However, it can in both cases be assumed that no modifications are required to the fuel injection system in the calciner in the case of oxyfuel retrofit.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

Small oxyfuel burners are required also in the calciner in the highly integrated CaL concept. It can be assumed that a two-burner solution is used. Equipment costs are in this case approximately 40 k€ for both burners, and installation costs are assumed negligible.

17 OXYFUEL CLINKER COOLER

Estimates for equipment and installation costs for a 3000 tpd oxyfuel clinker cooler are provided by IKN. Costs related to the unit both for a new built plant and for retrofit of an existing plant are given in Table 17.1.

Table 17.1. Equipment and installation costs for a 3000 tpd oxyfuel clinker cooler in the case of a new built plant and for retrofit of an existing plant.

	New	Retrofit
Equipment costs	3.3 $M€_{2017}$	2.9 $M€_{2017}$
Installation costs	1.0 $M€_{2017}$	0.7 $M€_{2017}$

In the case of a *new built plant*, the following components are included in the equipment cost:

- Cooler
  - Grate system including drive and surface
  - Supporting structure
  - Roll clinker crusher
  - Clinker discharge system
  - Cooler undergrate fans
  - Electrical parts
  - Engineering
- Housing
  - Upper and lower cooler housing
  - Air ducts from fans to grate
  - Stairs and platforms
- Refractory
  - Refractory of cooler housing
  - Horse shoe
  - Side curbs
- Auxiliaries
  - E.g. spare parts for erection and commissioning

The kiln hood, the oxyfuel recirculation system, and civil works are not included.

In the case of *retrofit of an existing plant*, the following differences to the new built case are taken into account, leading to slightly lower costs:

- Housing

    - The cooler housing is assumed to be in good condition and suitable in size and can mostly be re-used
    - Existing stairs and platforms are assumed to be adapted
- Refractory
    - Refractory of cooler housing is assumed to be re-used

Breakdown of the equipment costs in the two cases are given in Table 17.2.

Table 17.2. Breakdown of equipment costs of a 3000 tpd oxyfuel clinker cooler in the case of a new built plant and for retrofit of an existing plant.

	New	Retrofit
Cooler	71%	83%
Housing	11%	4%
Refractory	13%	7%
Auxiliaries	5%	6%

18 OXYFUEL KILN HOOD

The kiln hood is the connection between the cooler, kiln and tertiary air duct, and it hoists the main burner. It is expected that for retrofit, the existing kiln hood can be reused with some modifications. Estimates for the equipment and installation costs are provided by IKN. The costs for a new built case and for a retrofit case is given in Table 18.1. The equipment costs in the new built case includes refractory.

Table 18.1. Equipment and installation costs for a 3000 tpd kiln hood in the case of a new plant and for retrofit of an existing plant.

	New	Retrofit
Equipment costs	0.4 M€$_{2017}$	0.1 M€$_{2017}$
Installation costs	0.3 M€$_{2017}$	0.1 M€$_{2017}$

19 CONVENTIONAL AND SEALED FANS

Fans are necessary to ensure that the gases overcome the pressure drop in the plant. All fans considered in this chapter are centrifugal fans which can operate continuously at temperatures up to 400°C.

The TEC, for fans with different flow rate and power consumption, can be found by summing the cost of the following items:

1. *Machine cost"* (MC), which can be calculated with the following exponential scaling law of the volumetric flow rate *V*:

$$MC = MC_0 \cdot \left(\frac{V}{V_0}\right)^{0.5} \tag{19.1}$$

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

2. *"Electric motor cost"* (EMC), which can be calculated with the following exponential scaling law, function of the electric power *P*:

$$EMC = EMC_0 \cdot \left(\frac{P}{P_0}\right)^{0.65} \tag{19.2}$$

3. *"Electrical and instrumentation cost"* (EIC), as 30% of sum of the machine and electric motor cost:

$$EIC = 0.30 \cdot (MC + EMC) \tag{19.3}$$

4. *"Other costs"* (OC) (such as vendor's assistance, fan access, etc.) as 15% of the sum of above costs.

$$TEC = MC + EMC + EIC + OC \tag{19.4}$$

For the reference costs of fans (MC_0) and electric motors (EMC_0), the reference costs in Table 19.1 can be used. In case of sealed fans, the machine cost should be increased by 40% with respect to the indicated value.

Table 19.1. Machine and motor reference supply cost.

Flow rate [m³/h]	**25,000**	**380,000**	**485,000**
Machine cost, MC_0 [€]	20,800	170,400	288,000
Power [kW]	**75**	**1,100**	**2,000**
Electric motor cost, EMC_0 [€]	12,800	99,600	191,700

As far as the installation cost (IC) is considered, this corresponds to the 40% of the TEC:

$$IC = 0.40 \cdot TEC \tag{19.5}$$

20 ROTARY KILN INLET AND OUTLET SEALING

For the inlet and outlet sealing, ensuring "near-zero" false air ingress, TDC of 320 k€ (i.e. 160 k€ per sealing) can be assumed.

21 EXHAUST GAS RECIRCULATION SYSTEM

Most of the cost of the gas recirculation system is associated to the ducts.

In the oxyfuel cement kiln, cold CO_2 is recycled after exhaust gas cooling and water condensation. Most of the recycled CO_2 is mixed with oxygen and blown to the clinker cooler where it is heated

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

up and then used for combustion in the rotary kiln and the pre-calciner. In order to avoid water condensation in the recycle piping, external insulation is adopted. A total equipment cost of 105 € per m^2 of lateral surface is considered. The diameter of the piping can be calculated by assuming a gas velocity of 15 m/s. The length of the recycle piping will be largely site-dependent, being influenced by the plant layout. In CEMCAP, a piping length of 50 m will be assumed.

In the CaL cement plants, CO_2/H_2O-rich gas is recycled to the calciner inlet after partial cooling. In this case, gas is recirculated at 350-400°C prior to water condensation. A refractory lined recycle duct is assumed, with unit cost of 160 € per m^2 of lateral surface. The diameter of the piping can be calculated by assuming a gas velocity of 15 m/s. The length of the recycle piping will be largely site-dependent, being influenced by the plant layout. In CEMCAP, a piping length of 10 m will be assumed.

Equipment costs for oxygen pipes and the oxygen mixing system is assumed to be approximately 20 k€.

Installation cost can be calculated as the 85% of the duct equipment cost.

22 CALCIUM LOOPING REACTORS

22.1 Fluidized bed calciner

In the Tail end CaL configuration, circulating fluidized bed oxyfuel calciner is used, with an assumed gas superficial velocity of 5 m/s at reactor exit and a reactor height of 20 m. combustion is performed with an oxidant gas composed of oxygen from the ASU (50%vol.) and recirculated inert gas (50%vol.). The cost of the fluidized bed calciner is obtained as the sum of the following items: (i) the cost of the refractory lined reactor, (ii) the cost of cyclone and (iii) the cost of fuel handling equipment. The total equipment cost, expressed in k€$_{2014}$, can be calculated with eq. 22.1 as function of the coal thermal input to the reactor $\dot{Q}_{LHV}$.

$$TEC = 193 \cdot \left(\dot{Q}_{LHV}[MW]\right)^{0.65} \qquad (22.1)$$

Installation costs of 107% of the TEC are assumed.

22.2 Carbonator

Two design options are possible for the carbonator of the CaL process, namely: the adiabatic entrained flow (EF) carbonator, used in the integrated CaL process, and the cooled circulating fluidized bed (CFB) carbonator, used in tail-end CaL.

In the adiabatic EF carbonator, the heat of the exothermic carbonation reaction is stored as sensible heat in the gas-solid stream. Temperature of the gas-solid stream will increase as reaction proceeds from the initial ~600 to about 700°C at the outlet. The reactor will have refractory lined walls and

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

the cost in k€$_{2014}$ can be calculated with eq. (22.2), as function of the gas volume flow rate at carbonator inlet $\dot{V}_{in}$.

$$TEC = 85.9 \cdot \left(\dot{V}_{in}\left[\frac{m^3}{s}\right]\right)^{0.5} \tag{22.2}$$

It is assumed that the entrained flow carbonator is installed on a new tower, as explained in Chapter 27. Installation costs of the entrained flow carbonator are included in the tower cost.

In case of the CFB carbonator, heat transfer surface are placed inside the reactor to maintain a homogeneous temperature of about 650°C. The CFB carbonator closely resemble a CFB boiler with:

- air nozzles at the bottom of the bed, which allow air flow without clogging;
- cyclones for recirculation of solid particles;
- steam drum including downcomer and interconnecting piping for feed water and steam;
- safety valves, primary instrumentation elements and accessories;
- duct work, dampers and expansion joints;
- refractory lining in the (i) lower part of the chamber, (ii) cyclone, (iii) standpipe and (iv) loop seal;
- supporting structures with structural steel and insulation.

For this reason, it is assumed that the TEC of the carbonator is the same of the TEC of a CFB boiler operated with the same combustor temperature and the same thermal power transferred to heat exchangers tubes, producing steam with the same thermodynamic conditions (saturated steam at 100 bar). Commercial software Thermoflex® is used to estimate the TEC of CFB boilers of different size, resulting in a linear correlation between TEC (expressed in k€$_{2014}$) and transferred thermal power $\dot{Q}$ in the range 20-125 MW$_{th}$.

$$TEC = 217 \cdot \dot{Q}[MW_{th}] + 3{,}830 \tag{22.3}$$

Installation costs of 110% of the TEC are assumed.

22.3 Entrained flow oxyfuel calciner

The calciner is made of an external carbon steel shell and an internal refractory lining with fire bricks. The fundamental sizing laws for the calciner are based on the superficial gas velocity, which determines the cross-section, and the residence time needed to ensure burnout of the fuel, which determines the calciner length. Assuming a residence time of 3 seconds for coal combustion, and a gas superficial velocity of 15 m/s at the calciner outlet, the TEC (in k€$_{2014}$) is given by the following equation, where $\dot{V}_{out}$ is the gas volumetric flow rate at calciner outlet.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

$$TEC = 52.5 \cdot \left(\dot{V}_{out} \left[\frac{m^3}{s} \right] \right)^{0.5} \tag{22.4}$$

The cost calculated with the correlation above includes the cost of the auxiliary equipment, such as CO_2 blasters.

In the integrated calcium looping system, it is assumed that the oxyfuel calciner is installed on a new tower, as explained in Section 27. The installation costs of the CaL entrained flow calciner are included in the cost of the tower.

23 SUSPENSION PREHEATER STAGES

In the EF CaL configuration, different preheaters working with hot gases coming from different plant components are needed. Hot gas from the oxyfuel calciner is cooled in a raw meal preheater using CO_2-rich gas as heat source. Hot gas from the rotary kiln, the carbonator and the clinker cooler can be cooled in dedicated parallel preheaters.

Each preheater stage is composed of three parts: (i) the riser duct, where most of the gas-solid heat transfer occurs, (ii) the cyclone, for gas-solid separation and (iii) the feed pipe which transfers the solids separated in the cyclone to the lower stage riser.

The TEC of a single preheater stage is calculated as the sum of the TEC of cyclone, riser duct and feed pipe, which can be calculated as follows:

1. Riser ducts have cylindrical shape and are made of an external carbon steel structure and an internal refractory. Inner diameter is determined by the gas superficial velocity, which can be assumed equal to 17 m/s at the outlet of the riser. The length of each riser can be assumed equal to 14 m, to allow proper contact time and to fit with the size of the associated cyclone and feed pipe. Based on these sizing criteria, the lateral surface of the riser can be calculated. The cost per unit surface is reported in Table 23.1 as function of the riser exit temperature. Higher temperatures involve higher unit cost due to the higher thickness of the refractory.

Table 23.1. Cost of the riser surface as a function of gas temperature.

Riser exit temperature	< 500°C	500°C < T < 750°C	> 750°C
Unit cost [€/m^2]	160	230	350

2. The diameter and length of feed pipes are assumed equal to 900 mm and 10 m respectively, for each stage. The same cost per unit of lateral surface considered for riser ducts (Table 23.1), can be used for feed pipes as well.
3. The diameter of the cyclone is a function of volumetric gas flow rate inlet and can be calculated by the following equations. The first equation refers to the high efficiency first

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

stage of the preheater tower, while the second equation refers to the other stages. In both equations, the volumetric flow rate is expressed in m^3/s. Each preheater string features one cyclone per stage, with the exception of the first stage, where two parallel high efficiency cyclones are used.

$$D_{cyclone}[mm] = 647 \cdot \dot{V}^{0.422} \tag{23.1}$$

$$D_{cyclone}[mm] = 1548 \cdot \dot{V}^{0.28} \tag{23.2}$$

The cost of the cyclone can be assumed proportional to the diameter, with an increasing cost with the operating temperature, as indicated in Table 23.2.

Table 23.2. Specific cyclone cost as a function of gas temperature.

Cyclone inlet temperature	**< 500°C**	**500°C < T < 750°C**	**> 750°C**
Unit cost [€/m]	135,000	200,000	260,000

Therefore, the TEC is given by:

$$TEC = \sum_{i=1}^{\#stages} \left(Cost_{cyclone} + Cost_{riser} + Cost_{feed\ pipe}\right)_i \tag{23.3}$$

The IC of a preheater stage is 166% of TEC.

This costing methodology can be also used to estimate the cost of the sorbent cooler of the integrated CaL process, which is a riser where hot solids and cooling gas are contacted. For this component, which is assumed to be installed on an additional tower, as explained in Section 27. Installation costs of the sorbent cooler are included in the tower cost.

24 TUBULAR HEAT EXCHANGERS

In CaL cases, tubular heat exchangers are necessary for heat recovery, mainly to produce steam for the heat recovery steam cycle. Two different types of heat exchangers can be distinguished based on the nature of the fluids involved in the heat transfer process:

- Gas – water heat exchangers;
- Gas – gas heat exchangers.

When the steam cycle The cost of these heat exchangers is estimated as a part of the heat recovery steam cycle using the Thermoflex® software if the full cycle is simulated there.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

For cost estimation of single heat exchangers, the total equipment cost (TEC) of heat exchangers is calculated by summing the heat exchange surface cost and the auxiliary equipment (drums, valves, casing, etc.) cost, which is assumed equal to 150% of the surface cost.

The IC are 25% of the TEC:

$$IC = 0.25 \cdot TEC \tag{24.1}$$

24.1 Gas – water heat exchanger

Commercial software Thermoflex® is used for the sizing of gas-water heat exchangers, i.e. to calculate the heat transfer area, the number and the size of the tubes, etc. for a given tube material. The weight of the heat exchanger resulting from Thermoflex calculation is used together with assumed material cost for estimating the cost of the heat exchange surface. In Table 24.1 the assumed material cost and type as function of steam temperature is reported.

Table 24.1. Material cost and selection criteria as function of steam temperature.

Steam temperature	Material	Specific cost [€/kg]
T < 350°C	Carbon steel	3.0
350°C < T < 450°C	T22	6.0
T > 450°C	T91	10.5

24.2 Gas – gas heat exchanger

In both CaL configurations, an oxygen pre-heater recovering heat from the CO_2-rich gases is considered, where O_2 is preheated to moderate temperature (<200°C). The commercial software Thermoflex® is used to size this component. Due the high O_2 concentration in the flow from ASU (95%), stainless steel is for tubes material, with a specific cost of 7 €/kg.

Auxiliary equipment cost is assumed equal to 100% of the exchange surface cost. The IC are assumed equal to 25% of the TEC.

25 PLATE AND FIN HEAT EXCHANGER

Plate and fin heat exchangers are necessary for low temperatures in the CO_2 liquefaction process. Such heat exchangers are ideal for sub-zero temperatures due to their high surface areas per unit volume, and they are capable of handling multiple streams.

The cost estimation of plate and fin heat exchangers is done as suggested by ESDU [ESD, 2003]: for two-stream heat exchangers the C-value method is used, and for multi-stream heat exchangers the B-value method is used. Installation costs are assumed as 100% of the TEC.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

25.1 C-value method

In the C-value method, a cost per unit ($Q/\Delta T_{lm}$) is estimated, where Q is the rate of heat transfer between the two streams and ΔT_{lm} is the logarithmic mean temperature of the heat exchanger. C-values are presented by ESDA for the various fluid pairs for a range of fixed values of ($Q/\Delta T_{lm}$). Based on this, approximate values for heat exchange between gaseous CO_2 and liquid CO_2 are assumed, as shown in Table 25.1 (corrected for the cost year of €$_{2014}$ with the CEPCI).

Table 25.1. C-values (€$_{2014}$/(W/K)) for brazed aluminium plate-fin heat exchangers.

$Q/\Delta T_{lm}$ [W/K]	Pressure	30,000	100,000	1,000,000
Gaseous O_2/N_2 – Liquid O_2/N_2	< 25 bar	0.70	0.30	0.11
Gaseous hydrocarbon – Liquid hydrocarbon		0.68	0.27	0.09
Assumed for gaseous CO_2 – Liquid CO_2		0.69	0.29	0.10
Gaseous O_2/N_2 – Liquid O_2/N_2	25-40 bar	0.77	0.33	0.12
Gaseous hydrocarbon – Liquid hydrocarbon		0.75	0.30	0.10
Assumed for gaseous CO_2 – Liquid CO_2		0.76	0.32	0.11

C-values for other values of ($Q/\Delta T_{lm}$) than the ones presented in Table 25.1 can be found by logarithmic interpolation. Thus, if a value C_1 is known for a value $(Q/\Delta T_{lm})_1$ and the value C_2 is known for a value $(Q/\Delta T_{lm})_2$, then the C-values for a ($Q/\Delta T_{lm}$) lying between $(Q/\Delta T_{lm})_1$ and $(Q/\Delta T_{lm})_2$ can be estimated from the equation:

$$C = \exp\left(\ln C_1 + \frac{\ln\left(\frac{C_1}{C_2}\right) \ln\left[\left(\frac{Q}{\Delta T_{\mathrm{lm}}}\right)/\left(\frac{Q}{\Delta T_{\mathrm{lm}}}\right)_1\right]}{\ln\left[\left(\frac{Q}{\Delta T_{\mathrm{lm}}}\right)_1/\left(\frac{Q}{\Delta T_{\mathrm{lm}}}\right)_2\right]} \right) \tag{25.1}$$

Finally, the cost estimate is given by:

$$\mathrm{TEC} = (Q/\Delta T_{lm}) \times C \tag{25.2}$$

The TEC for stainless steel heat exchanger can be estimated as five times the TEC of a brazed aluminum heat exchanger.

25.2 B-value method

In the B-value method, B-values are used to estimate active volume, which again is used to estimate cost. The following steps are taken:

1. Hot and cold composite curves are constructed.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

2. The composite curve diagram is split into zones, and for each zone values of the mean overall volumetric heat transfer coefficient for the zone, B_z, are estimated from the following equation:

$$\frac{Q_z}{B_z} = \sum_{i=1}^{n} \frac{Q_i}{\beta_i} \tag{25.3}$$

   where Q_i is the amount of heat transferred to or from the i^{th} stream in the zone and β_i is the volumetric film coefficient relating to the i^{th} stream.

3. For each zone, the value of $\Delta T_{lm,z}$ is calculated using the zone hot and cold end temperatures.

4. The active heat transfer volume per zone is calculated by:

$$V_z = \frac{Q_z/\Delta T_{\mathrm{lm},z}}{B_z} \tag{25.4}$$

5. The total active heat transfer volume, V, of the heat exchanger is calculated by summing the volumes of the zones and adding an allowance of 15% for headers and distributors:

$$V = 1.15 \sum_{z=1}^{nz} V_z \tag{25.5}$$

where nz is the total number of zones.

The TEC of the heat exchanger is obtained by multiplying V by the cost per unit volume that can be found in Figure 1 in [ESD, 2003], and correcting for the cost year using the CEPCI and for the use of stainless steel by multiplying the TEC with five. The cost curves presented in Figure 1 in [ESD, 2003] are valid for pressures below 25 bar, for pressures in the range of 25-40 bar, the TEC should be multiplied with 1.1.

26 HEAT RECOVERY STEAM CYCLE

The steam generated in the CaL plant is converted into electric power in a Heat Recovery Steam Cycle (HRSC). Commercial software Thermoflex® is used to size the HRSC, according to the assumptions presented in D3.2.

27 ADDITIONAL TOWER

In the integrated CaL configuration an additional tower is required to support the CaL reactors and the sorbent cooler, whose total weight exceeds the weight of the existing equipment supported

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

by the existing preheater tower. The cost of the additional tower, which includes the installation cost of the above mentioned CaL system components, is assumed to be proportional to the weight of the equipment to be supported, according to the following equation (in k€$_{2014}$).

$$TDC = 52.2 \cdot W[ton] \tag{27.1}$$

28 CHILLED AMMONIA PROCESS

The chilled ammonia CO_2 capture and conditioning process is costed by GE using the proprietary tool QFACT which is based on an extensive database of executed projects. QFACT is a computerized in house developed model that was developed as a tool to perform equipment factored estimates based on the correlation of construction quantities. In this tool, physical equipment data, based on simplified design correlation verified against statistical return data, and general project documentation (Process Flow Diagram and Plot Plan) are used as basis of the estimates. This cost estimation will be done, following the guidelines of the AACE, for a budget cost estimate with a targeted accuracy of ±30%.

A breakdown of the numbers into single positions would disclose GE confidential information about cost structure and/or pricing strategy and as such no detailed cost breakdown will be provided for this technology. However, the elements of the cost estimation are lumped into the two main categories (see Figure 2.1):

- Equipment cost (incl. Bulk Material cost)
- Installation cost (incl. Erection)

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

29 REFERENCES

[AGR, 2011] Agrawal, K.C., 2001. *Industrial Power Engineering and Applications Handbook*. Newnes Power Engineering Series. (Chapter 1.8).

[ASM, 2015] ASME VIII, 2015. Section 1: *ASME Boiler and Pressure Vessel Code*, Section VIII, Division 1: *Rules for Construction of Pressure Vessels*.

[BUR, 2011a] Bureau-Cauchois, G., Roussanaly, S., Husebye, J., 2011. *Economic Assessment of Flue Gas & CO_2 Collecting Networks*. Deliverable N° D4.1.1, COCATE EU FP7 Project.

[BUR, 2011b] Bureau-Cauchois, G., Roussanaly, S., Husebye, J., 2011. *Economic Assessment of CO_2 Export Systems and comparison of implementation strategies*. Deliverable N° D4.1.2, COCATE EU FP7 Project.

[CEP, 2017] Chemical Engineering, 2017. Economic Indicators: Chemical Engineering Plant Cost Index (CEPCI).

[ESD, 2003] ESDU, 2003. *Selection and cost of heat exchangers*. Plate-fin type. ESDU 97006.

[IEA, 2008] IEA GHG, 2008. *CO_2 capture in the cement industry*. Report 2008/03.

[IEA, 2013] IEA GHG, 2013. *Deployment of CCS in the cement industry*. Report 2013/19.

[IEA, 2014a] IEA GHG, 2014. *CO_2 capture at coal-based power and hydrogen plants*. Report 2014/03.

[IEA, 2014b] IEA GHG, 2014. *Evaluation of reclaimer sludge disposal from post-combustion CO_2 capture*. Report 2014/02.

[IEA, 2017] IEA GHG, 2017. *Evaluating the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery*. Report 2017/08.

[JOS, 1998] Joseph, G.T., Beachler, D.S., 1998. *Scrubber Systems Operation Review, Lesson 9: Flue Gas Desulfurization (Acid Gas Removal) Systems*. U.S. Environmental Protection Agency (EPA) APTI Course SI: 412C.

[KEY, 2011] Keys to Metals AG, 2011. *Steels for Cryogenic and Low-Temperature Service.* Available from: http://www.keytometals.com/page.aspx?ID=CheckArticle&site=kts&NM=61

[KNU, 2009] Knudsen, J. N., Jensen, J. N., Vilhelmsen, P-J., Biede, O., 2009, *Experience with CO_2 capture from coal flue gas in pilot-scale: Testing of different amine solvents.* Energy Procedia, 1: 783-790.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

[KOH, 1997] Kohl, A., Nielsen, R., 1997. *Gas Purification Handbook*, 5th edition. Gulf Publishing Company

[MCC, 2005] McCabe, W. L., Smith, J. C., Harriott, P., 2005. *Unit Operations of Chemical Engineering*, McGraw-Hill, New York, U.S.

[ONA, 2017] Onarheim, K., Santos, S., Kangas, P., Hankalin, V., 2017. *Performance and costs of CCS in the pulp and paper industry part 2: Economic feasibility of amine-based post-combustion CO_2 capture.* International Journal of Greenhouse Gas Control, **66**: p. 60-75.

[PAN, 1997] Panos, P. J., 1997. *Fluid Bed Ash Cooler*. Patent US005954000A.

[REZ, 2009] Rezvani, S., Bolland, O., Franco, F., Huang, Y., Span, R., Keyser, J., Sander, F., McIlveen-Wright, D., Hewitt, N., 2009. *Natural gas oxy-fuel cycles—Part 3: Economic evaluation.* Energy Procedia 1, 565-572.

[ROU, 2014] Roussanaly, S., Lindqvist, K., Anantharaman, R., Jakobsen, J., 2014. *A Systematic Method for Membrane CO_2 Capture Modeling and Analysis.* Energy Procedia, **63**(0): p. 217-224.

[ROU, 2016] Roussanaly, S., Anantharaman, R., Lindqvist, K., Zhai, H., Rubin, E., 2016. *Membrane properties required for post-combustion CO_2 capture at coal-fired power plants.* Journal of Membrane Science, **511**: p. 250-264.

[SAR, 2013] Sargent & Lundy L.L.C., 2013. *IPM Model - Updates to Cost and Performance for APC Technologies: Wet FGD Cost Development Methodology*, Project 12847-002.

[SLU, 1992] Van Der Sluijs, J. P., Hendriks, C. A., Blok, K., 1992. *Feasibility of polymer membranes for carbon dioxide recovery from flue gases.* Energy Conversion and Management, **33**(5–8): p. 429-436.

[ZHA, 2013] Zhai, H., Rubin, E., 2013. *Techno-Economic Assessment of Polymer Membrane Systems for Postcombustion Carbon Capture at Coal-Fired Power Plants.* Environmental Science & Technology, **47**(6): p. 3006-3014.

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185