Thermodynamic and Economic Evaluation of a 1000MWth Chemical Looping Combustion Power Plant

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Thermodynamic and Economic Evaluation of a 1000MW$_{th}$ Chemical Looping Combustion Power Plant

Master's thesis work

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Declaration of authorship

I hereby declare that, to the best of my knowledge and belief, that this thesis contains no material previously published or written by any other person except where due reference is made in the text of the thesis.

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(Maxime Olausson)

Abstract

Chemical looping combustion (CLC) is a CO_2-capturing technology that has a potential to provide competitive CO_2-capturing costs for solid fuel fired power plants. A number of CLC power plants of pilot scale are available at research institutes, but still no in industrial scale. When scaling up from pilot scale to full industrial scale, an important step is to determine if the technology is economically feasible or not. The aim of this work is to provide a deeper economic analysis of CLC for industrial scale applications than what has been done previously, with help of results from process simulation modelling and an economic model.

Previously developed process models by the EST department at TU Darmstadt of a 1000 MW_{th} hard coal fired CLC power plant have been utilized and further developed in order to improve feasibility. Relevant data has been extracted from the model results, and used as input when predicting the plant economy. In the economic evaluation, a number of economic key performance indicators have been estimated. Furthermore, a sensitivity analysis has been carried out, to identify which parameters that have the greatest impact on the plant economy.

The process models predict the net electrical efficiency of the plant to 41%. Results from the economic evaluation show that the levelized cost of electricity is expected to be 48 $\frac{€}{MWh}$, the CO_2 avoidance cost 23 $\frac{€}{tonne\ CO_2}$ and the payback-period ten years. The economic evaluation shows that the single greatest contributor to the plant lifetime cost is the fuel cost, which accounts up to 25 % of the plant lifetime cost. A conclusion from this work is that CLC offers competitive CO_2-capturing costs in comparison with other CO_2-capturing technologies.

Keywords: chemical looping combustion, CLC, CO_2-capturing, techno-economic evaluation, LCOE, CO_2-avoidance cost

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Abbreviations and symbols

Chemical substances

CO	Carbon monoxide
CO_2	Carbon dioxide
H_2	Hydrogen
H_2O	Water
Me_xO_y	Oxidized metal oxide
Me_xO_{y-1}	Reduced metal oxide
N_2	Nitrogen
O_2	Oxygen
SO_2	Sulphur dioxide

Abbreviations

AR	Air Reactor
ASU	Air separation unit
CAPEX	Capital Expenses
CCS	Carbon dioxide capture and sequestration
CFB	Circulating Fluidized Bed
CLC	Chemical Looping Combustion
DM	Deutsche Mark
EPC	Engineering, Procurement and Construction
EST	Institut Energiesysteme und Energietechnik
FORTRAN	Formula Translation (programming language)
FR	Fuel Reactor
IEAGHG	International Energy Agency Greenhouse Gas R&D Programme
iG-CLC	In-situ Gasification Chemical Looping Combustion
KPI	Key Performance Indicator
LCOE	Levelized Cost of Electricity
OC	Oxygen carrier
OPEX	Operating expenses
POC	Post Oxidation Chamber
SI	Solid inventory
TU	Technische Universität

A	Area	$\left[m^2\right]$
AD	Annual depreciation	$\left[\frac{€}{year}\right]$
AEP	Annual electricity production	$\left[\frac{MWh_{el}}{year}\right]$
bhp	Power of motor shaft	$[hp]$
$CAPEX_n$	Capital expenses at year n	$[€]$
CO_2AC	CO_2-avoidance cost	$\left[\frac{€}{tonne\ CO_2}\right]$
ccf	Cumulative cash flow	$[€]$
ct	Construction time	$[years]$
d	Diamater	$[m]$
dp	Depreciation period	$[years]$
d_1	Operating pressure factor	$[-]$
d_2	Material factor	$[-]$
F	Force	$[N]$
F_d	Heat exchanger type factor	$[-]$
F_m	Material factor	$[-]$
F_p	Operating pressure factor	$[-]$
f_{fins}	Finned tube correction factor	$[-]$
f_{FR}	Fuel reactor correction factor	$[-]$
f_{pipes}	Curvature correction factor of pipes	$[-]$
g	Acceleration of gravity	$\left[\frac{m}{s^2}\right]$
h	Height	$[m]$
I_n	Interest payment at year n	$[€]$
IC	Investment cost	$[€]$
ir	Assigned rate of return (interest rate)	$[-]$
l	Total length of tubes	$[m]$
lt	Plant lifetime	$[year]$
LAR	Levelized cost of asset retirement	$\left[\frac{€}{year}\right]$
LC	Levelized cost	$\left[\frac{€}{year}\right]$
$LCOE$	Levelized cost of electricity	$\left[\frac{€}{MWh_{el}}\right]$
L_1	Equipment dimension factor	$[-]$
L_2	Equipment dimension factor	$[-]$
m	Exponent of the six-tenths factor rule	$[-]$
m_{si}	Solid inventory	$[kg]$
$\dot{m}$	Mass flow	$\left[\frac{kg}{sec}\right]$
$M\&S$	Marshall & Swift cost index	$[-]$
n	Year of operation	$[-]$
NPV	Net present value	$[€]$

OC_{lt}	Oxygen carrier lifetime	$[hour]$
OC_{oxc}	Cost of oxygen make-up stream	$\left[\frac{€}{year}\right]$
$OPEX_n$	Operating expenses at year n	$[€]$
P	Power	$[W]$
p	Price	$[€]$
r	Annuity factor	$\left[year^{-1}\right]$
rev_n	Revenue at year n	$[€]$
SI	Solid inventory	$[tonne]$
sAR	Specific asset retirement price	$\left[\frac{€}{kW_{el}}\right]$
tax	Corporate tax rate	$[-]$
$w_{superheaters}$	Total weight of all superheater tubes	$[tonne]$
x_n	Percentage of payment at year n	$[-]$
ΔP	Pressure drop	$[Pa]$
η	Efficiency	$[-]$
ρ	Density	$\left[\frac{kg}{m^3}\right]$

1 Introduction

One of the absolutely most important issues of the world is global warming caused by human activity. The main cause of this is an increased use of fossil fuels as coal, oil and gas since 1850, leading to rapidly growing concentrations of greenhouse gases in the atmosphere [1]. Power generation from fossil fuel fired power plants is the largest single source of contribution [2]. As the vision to replace the total power generation by renewable energy sources still is hindered by economic and practical constraints, it is necessary to find clean methods to produce energy from fossil fuels. One alternative is to capture the greenhouse gases emitted from the processes that take use of fossil fuels, before it is released to the atmosphere.

Chemical Looping Combustion (CLC) is a promising technology for carbon dioxide capturing, that has been under significant development since early 2000, when the European Union funded research projects related to this field. The purpose of this technology is to capture CO_2 from large emitters in the energy sector. The core idea is that oxygen is captured from air by oxygen carriers, and then transferred to a fuel reactor where the oxygen is released. Subsequently, the oxygen carrier is returned to the air reactor to re-start the cycle. Metals or metal oxides are suitable for being used as oxygen carriers, and constitute the fluidized bed in the fuel reactor. This described configuration is the one used in the majority of the CLC units that exist worldwide, and is known as *interconnected fluidized-bed reactors*. A basic sketch of the concept is shown in Figure 1.1. Other configurations have also been investigated, but this thesis work will focus on interconnected fluidized-bed reactors, as it is the one with most information available. In addition to the two reactors, heat exchangers, post combustion system, separation units, gasification system if the fuel is a solid, and a CO_2 compression unit to ease storage and transportation of the CO_2, are basic components of a CLC plant.

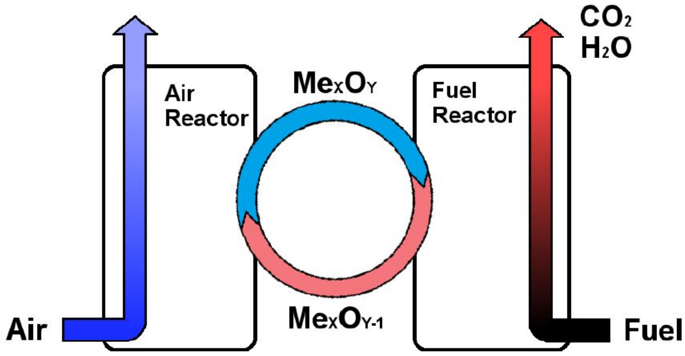


Figure 1.1: Illustration of a CLC interconnected fluidized-bed reactor system. Figure from Gunnarson, A. [48]

What is achieved by utilizing this technology, is that the exhaust gas from the combustion gas mainly consists of CO_2, instead of being diluted by nitrogen and other gases which is the case if air directly is used as the oxygen source for combustion of the fuel.

There is currently no CLC plant operating in industrial scale, but there have been studies about how such a plant configuration should look like, and estimations about costs in comparison to other state-of-art technologies [3, 4, 5]. One of the largest CLC pilot plants are located at Technische Universität Darmstadt (TU Darmstadt). [6]

One advantage with CLC is that due to the low content of nitrogen and sulfur species in the flue gas stream from the fuel reactor, heat from this mentioned stream can be recovered by cooling it down to temperatures lower than possible for conventional coal fired power plants. This because for conventional coal fired power plants, the exhaust gas stream contains significant amounts of nitrogen and sulphur compounds that may condensate and cause corrosion. The dew points for the relevant acids are around $130°C$, and that is consequently the lower limit of how far the flue gas stream may be cooled down for conventional coal fired plants [32].

Furthermore, a key environmental advantage of CLC, compared to other CO_2 capture technologies, is that oxygen is separated from air without the extensive use of air separation units (ASUs) which demand big amounts of energy that significantly reduces the efficiency of the process [10]. In CLC, ASUs are needed for a post-combustion phase to maximize the conversion of fuel, but the requirement of pure oxygen is far lower than what is needed in for instance oxy-fuel combustion.

2 Project objectives

The main goal of the project is to estimate process related parameters and economic parameters of a grassroot or semi-grassroot full industrial scale coal fired 1000 MW_{th} CLC power plant. The process models developed in project ACCLAIM [4] by Ohlemüller, P. and John, M. will be improved and a process flow sheet of the main CLC components and the steam cycle will be proposed.

Based on the results, an economic evaluation will be carried out to determine if an upscaled CLC plant would be economically feasible and competitive with other CSS-technologies. Previous techno-economic studies about full-scale CLC plant have been made by Lyngfelt, A. in [3] and in [27], in which they invite the scientific public for further studies within the field. This previous work will be used to conduct a more detailed economic evaluation by estimating the number, design and dimensions of the required process components; taking into account indirect capital costs; looking at the operational expenses; considering how the cost varies during the plant lifetime; and by taking into account the time value of money. Other similar techno-economic studies made about CCS technologies [11, 12, 18] will be highly useful as references.

Other objectives are to identify which sources that have the greatest impact on the total cost and to compare the operating expenses with the capital expenses. It is also of high relevance to understand how changes in some operating parameters and how economic parameters as price fluctuations and changes in interest rate affect the financial results. A sensitivity study will therefore be done. The main key performance indicators that will be focused on, are listed in Table 2.1.

Table 2.1: Project objective key performance indicators (KPI) to be identified

Process KPI	**Economic KPI**
Plant overall electrical efficiency	Investment cost
Process stream properties	Operating cost
Utility demand ‡	Levelized cost of electricity
Internal plant electricity demand *	CO_2-avoidance cost
Rough design, dimensions, performance and number of each process unit in the proposed plant design	Payback period

‡ Demand of oxygen carrier, fuel, cooling water

* Electricity demand for CO_2-compression, air separation unit, compressors, pumps, motors etc

3 Process fundamentals of chemical looping combustion

In CLC, direct contact between fuel and air is avoided by taking use of an oxygen carrier that transfers oxygen from air to the fuel reactor where the combustion takes place, as illustrated in Figure 3.1. In the fuel reactor, the fuel is combusted to CO_2 and H_2O by an oxidized metal oxide, denoted Me_xO_y. The metal oxide is reduced in the fuel reactor to Me_xO_{y-1} and then transported to the air reactor, where it is re-oxidized to Me_xO_y.

Both gaseous fuels and solids can be used as fuel. In case of a gaseous fuel, e.g refinery or natural gas, the fuel reacts directly with the oxygen carrier. If instead a solid material is used as fuel, e.g coal or biomass, other operations are required. In a syngas-CLC process, the oxygen carrier reacts with gasification products (syngas), that is the product from a gasifier. With other words, a solid fuel is used as the primary fuel, but must be pyrolized before coming into contact with the oxygen carrier. The gasification may occur in an external gasifier, or be gasified directly in the fuel reactor. The latter technique is called in-situ gasification CLC (iG-CLC). In this case, the solid fuel is mixed together with the oxygen carrier in the fuel reactor. iG-CLC is the setup that is focused on in this study. [6]

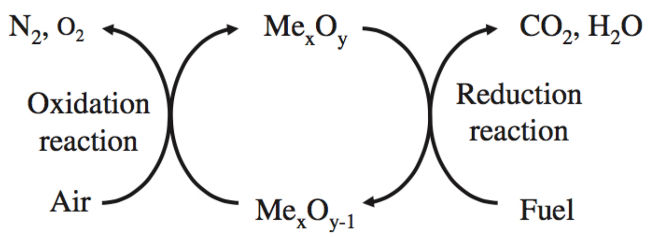


Figure 3.1: Chemical Looping Combustion system for gaseous fuels with a metal oxide as oxygen carrier. Figure from [6]

The fuel reactor in an iG-CLC system is a fluidized bed with the solid bed made up by oxygen carrier, solid fuel and burnt solid fuel. Figure 3.2 shows a typical configuration if coal is used as fuel. The bed is fluidized with H_2O, CO_2 or a mixture of these gases, and are the gasifying agents. CO_2 is particularly proposed for highly reactive solid fuels. Otherwise CO_2 as gasifying agent results in slow gasification rates. [6]

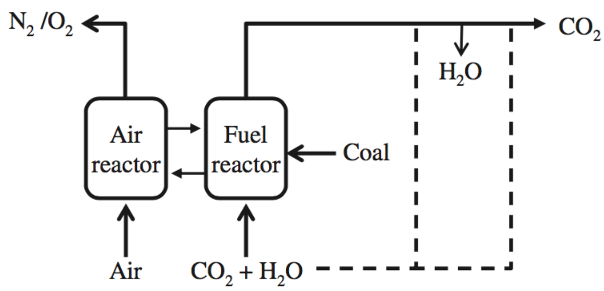


Figure 3.2: Schematic layout of a solid fuelled iG-CLC. Figure from [6]

A significant difference with iG-CLC systems compared to when the fuel is directly fed into the fuel reactor as a gas, is that the syngas is obtained from the char particles that move inside the fuel reactor. This means that a portion of the syngas that is released, particularly in the upper regions of the bed, will have insufficient contact with the oxygen carriers and there are difficulties to achieve complete conversion of the syngas. Therefore, oxy-polishing is needed, i.e to introduce pure oxygen in a later step to oxidize the remaining reactants. If the fuel instead is fed as a gas, the gas is introduced from below and the residence time is more uniform for the gaseous fuel molecules. In this case, oxy-polishing might not be necessary. [3]

The gasification in an iG-CLC process begins with reactions (1,2,3). The resulting gases and volatiles are thereafter oxidized through reduction of the oxidized oxygen carrier, Me_xO_y, according to the reaction shown in 4. In addition, the water-gas shift equilibrium in reaction (5), affects the composition of gas in the fuel reactor. In the air reactor, the reduced oxygen carrier, Me_xO_{y-1}, is oxidized with oxygen from air as shown in reaction (6). Thus the oxygen carrier starts a new cycle. An important remark is that the net chemical reaction is equal to the one in usual combustion and has the same combustion enthalpy. [6]

$$Coal \longrightarrow Volatile\ matter + Char \tag{1}$$
$$Char\ (mainly\ C) + H_2O \longrightarrow H_2 + CO \tag{2}$$
$$Char\ (mainly\ C) + CO_2 \longrightarrow 2\ CO \tag{3}$$
$$H_2, CO, Volatile\ matter + n\ Me_xO_y \longrightarrow CO_2 + H_2O + n\ Me_xO_{y-1} \tag{4}$$
$$H_2O + CO \longleftrightarrow H_2 + CO_2 \tag{5}$$
$$Me_xO_{y-1} + \frac{1}{2}\ O_2 \longrightarrow Me_xO_y \tag{6}$$

The pressure of the reactors may have a significant impact on the reaction rate [6]. A higher pressure may affect the absorption properties of surface complexes on the char, which directly has an effect on the reaction rate in a positive way [22]. Other studies show that a higher pressure does not lead to increased reaction rates of the oxygen carrier [23]. Gasification rates are also pressure dependent until a certain threshold [24].

3.2 Components in a chemical looping combustion configuration

The main components of an iG-CLC configuration are described in this section.

3.2.1 Air reactor and fuel reactor circulating fluidized bed boilers

Both the air reactor and the fuel reactor are circulating fluidized bed reactors. They are expected to be of similar design as commercially available technology for conventional CFB power plants, with a few differences [3]. In an ideal case, the whole bed of the air reactor is made up by the oxygen carrier. In the fuel reactor, the oxygen carrier builds up the bed together with the solids from the fuel if there are any, depending on the type of fuel. The fuel reactor is proposed to be built adiabatically with efficient surrounding isolation, while the air reactor is proposed to be enclosed by fluidized bed coolers on the wall sides [3, 4]. It is typical that evaporation occurs in the fluidized bed cooler tubes, because of the sufficient amount of conductive heat transfer that takes place between the bed and the wall [35]. Figure 3.3 shows how the inner wall of a bubbling fluidized bed can be covered with heat transfer tubes. The rest of the heat is transferred in the second section, called the *convective section* or *back pass*, where the superheater, economizer and air pre-heater is located [35]. In the convective section, as the name implies, convective heat transfer dominates.

Superheaters must be built in a way that it superheats approximately the same amount at both low and high loads. A way to do this, is by selecting proper convective heat surfaces. Tube lengths and the passes can be designed in a way to control temperature differences. For instance, the tubes that the hot flue gases first are exposed to, should be shorter than other tubes. A good superheater arrangement also overcomes the problem of uneven or biased flow. Since the heat transfer properties inside the tubes can be different from the corresponding on the outside of the tubes due to different flows or fluid type, as between a gas and a liquid, the heat transfer surface can be manipulated, both through construction material and geometry. An example is to use finned tubes, shown in Figure 3.4, to enhance the heat transfer on the outer side by increasing the surface area on one side. [16]

Figure 3.3: Bubbling fluidized bed reactor with walls covered by evaporator tubes. Figure from Teir, S. [16]

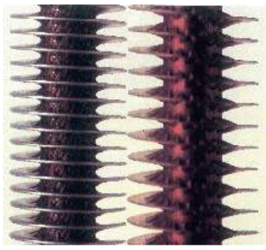


Figure 3.4: Finned heat transfer tubes, with the purpose to increase the heat transfer on the outer side. Figure from Teir, S. [16]

3.2.2 Post Oxidation Chamber

Due to non-perfect mixing, full conversion of the fuel in the fuel reactor is in practice impossible to achieve. Unburned fuel will be present in the flue gases and means an economical loss, but may also be harmful for the environment. Additional measures are thus required to avoid a release of unburned fuel to the environment. Letting the flue gases pass a post oxidation chamber (POC) is in CLC technology a common way to enable close to complete conversion of fuel. There have so far not been many studies made about how a post oxidation chamber should be designed, but in pilot plants and in literature, it has been suggested to be build a POC as a combustion chamber with inlets of pure oxygen that is allowed to be well mixed with the flue gas [3]. Recent modeling studies of data from the CLC pilot plant at Chalmers show that a residence time of approximately 1.3 seconds is reasonable for pilot scale post oxidation chambers [7]. Lyngfelt suggests a residence time between 0.5-1 seconds for a post oxidation chamber for a 1000 MW_{th} power plant [3].

Results from the pilot plant at TU Darmstadt show that 99 % of the unconverted methane from the fuel reactor can be converted in a post oxidation chamber [30]. It also showed that the post oxidation chamber should be operated at temperatures above 900-950 $°C$ to achieve sufficient methane conversion. A possibility to enable adequate temperature control is to perform the post oxidation in several stages.

This method of introducing pure oxygen in a post oxidation chamber is also called *oxygen polishing*. The drawbacks are that an air separation unit for production of oxygen is required and careful control needs to be taken of how much oxygen that is added.

3.2.3 Air separation

The oxygen demand for the post oxidation is supplied by an air separation unit (ASU). The only production method that is commercially available today for producing large quantities of highly pure oxygen for a reasonable price is cryogenic distillation. Other air separation technologies as vacuum swing adsorption, pressure swing adsorption or separation by polymeric membranes are not economically competitive for production of large amounts of highly pure oxygen [8]. Ceramic membrane air separation has shown promising results lately to reduce oxygen production costs [9]. However, this technology is not commercially mature and will not be taken into account in this work, even though it has a future potential.

The main components of an air separation unit consist of an air compressor, pre-cooling system, purification unit to remove water and CO_2 prior to entering the cryogenic section, heat exchangers, distillation columns, vaporizers and condensers. For air separation units with a capacity of up to 5000 tons/day, a process scheme with double columns is proposed by Air Liquide, consisting of one high pressure and one low pressure column. [8]

3.2.4 Oxygen carrier

The task of the oxygen carrier is to transport oxygen from the air reactor to the fuel reactor. The oxygen carrier has a limited lifetime and can therefore only be used for a certain number of cycles before it should be discarded or recycled. Together with the coal ash, the oxygen carrier constitutes the main solid wastes from the plant. Several materials have been evaluated as oxygen carrier material, as ilmenite, industrial iron oxide scale, process sand and manganese ore. Most of them are environmentally safe to dispose after usage. The main contaminants from the plant is expected to come from the fuel that is introduced. Even if the oxygen carrier is in direct contact with the fuel, the relatively short lifetime of the oxygen carrier is expected to prevent contamination through adsorption. [10]

3.2.5 Cyclones

A cyclone is a stationary device that utilizes the centrifugal force to separate solid or liquid particles from a carrier gas. The gas mixture is introduced to the cone-formed cyclone in the upper part, and gives rise to an axially descending spiral of gas. The centrifugal force forces the particles towards the walls, causing it to fall down and exit through the bottom. In most cyclonic separators, the gas constricts and reverses its axial direction, and exits the cyclone at the top. The separation efficiency is determined by the particle size distribution and dimensions of the cyclone. [34]

A minimum of two or three cyclones are needed in a CLC configuration [14, 15, 3]. One in which the air reactor flue gas passes through, to separate the oxygen carrier from the flue gases. The other one or two are connected to the fuel reactor flue gas stream. In the first cyclone, the oxygen carrier is separated from the gas and fine char particles. The fine char particles could then be separated from the gas in the second cyclone with high separation efficiency, and be sent back to the fuel reactor. However, some of the char particles will be present in the solid stream exiting the first cyclone. In order to improve the separation, a carbon stripper is most likely needed [15].

3.2.6 Carbon stripper

A carbon stripper is a bubbling fluidized bed, with the aim to separate and/or gasify away char particles that have slipped together with the oxygen carrier stream from the fuel reactor. The carbon stripper of the pilot plant at TU Darmstadt is rectangularly shaped and is fluidized with steam [15]. Results from the pilot plant at TU Darmstadt show that the usage of a carbon stripper is necessary to achieve a high conversion of the fuel and to enhance a high CO_2-capture efficiency [14]. Otherwise, it is likely that part of the unconverted char passes through to the air reactor and is burnt there instead of in the fuel reactor as meant.

However in [3], there is a theory that a carbon stripper will not be necessary in a 1000 MW_{th} scale plant. This because the fuel reactor would have a height enough to induce a residence time enough to fully convert the solid fuels, so that unconverted char particles are not slipped through with the oxygen carrier. The answer to this will be given when and if pilot plants of larger scale are constructed.

3.3 Fuel grinding

Something that is essential to enable a high conversion of the fuel, is to allow the char particles to have a high residence time in the fuel reactor. This is possible by using coal with a small particle size, in combination with a high solids inventory and by using a carbon stripper [14]. This means that a finer particle size of the coal is needed than for coal used for conventional coal fired power plants, which induces an additional fuel grinding cost [3]. The disadvantage is that finer particles makes the separation from the gas flow more difficult.

3.4 CO2 capture

In order to efficiently and safely store CO_2, it needs to be compressed to a liquid or supercritical fluid and have a certain purity. The compression is done by compressing the CO_2 rich flue gas from the FR by a series of compressors. As a rule of thumb for modern centrifugal compressors, the compression ratio for each unit should not exceed four [37]. The compression ratio is the factor that a compressor increases the pressure by. This general rule is to avoid excessive heating and to allow cooling steps between the compression stages. The cooling, which is made by a number of heat exchangers, allows separation of water by condensation. If the flue gas stream still contains too high levels of contaminants after the water separation, other separation stages may be necessary.

4 Methodology

Two process simulation models, further described in the Section 5, have been utilized. They are both based on previous models created by Ohlemüller, P. and John, M. from TU Darmstadt [4, 5] with data obtained from experimental results. Aspen Plus was used to simulate the main characteristic process units of a CLC plant, while Ebsilon Professional was used to simulate the steam cycle with a CO_2 compression section included. It would have been preferable to use solely one software for the whole simulation, but the capability of Aspen Plus is rather limited when it comes to steam cycle modeling. Therefore Ebsilon Professional was used to simulate the steam cycle in a more detailed way, which gives room for optimization of the system and a detailed look on each process unit. The Ebsilon model has been improved to be a more realistic representation of a real steam cycle, while in the Aspen model, only the boundary conditions have been altered to match the changes in the Ebsilon model.
The models were used to predict plant performance, operating conditions and design parameters of the process units of a 1000 MW_{th} coal fired CLC plant.

In addition to the two process simulation models, a number of matlab scrips have been created for the economic assessment. In the scripts, a number of operating parameters are calculated which together with results from the process models give information about the overall plant efficiency and electricity output. The capital cost, or also called capital expenses (CAPEX), has been estimated as a function of the total installed equipment cost, similar to the "Percentage of Delivered-Equipment Cost method" described in [31], with guidelines from the Global CCS Institute about how to perform cost estimations on CCS technologies [18]. The total installed equipment cost can be evaluated by selecting proper cost functions for each process unit. With this method, the total capital costs can then be approximated by multiplying the total installed equipment cost with a certain factor to cover direct plant costs as piping and control systems, and indirect owner costs as contingency, legal expenses and engineering. More about this is explained in Chapter 6.

Operating expenses have been estimated with information from literature and data extracted from the developed process models, such as stream flows. Labor and maintenance costs were approximated as functions of the total installed equipment cost, in accordance to what is done in [11, 31].

With cost functions for the capital and operating expenses determined, the key performance indicators mentioned in Chapter 2 were evaluated. Various sources of scientific literature have been used to determine reasonable cost functions and assumptions. In particular, previous work made by employees of TU Darmstadt [11, 17] and Max S. Peters' Plant Design and Economics for Chemical Engineers [31] have been helpful to a great extent to this part of the work.

Once all cost functions were determined, a sensitivity analysis was carried out, to get an understanding of which parameters, e.g oxygen carrier lifetime, fuel price, annual operating time, that have the greatest influence on the plant economy. The working procedure is shown below.

1. Improvements and modifications of the Aspen and Ebsilon models simultaneously
2. Determination of process key performance indicators from modelling results and calculations in Matlab
3. Development of capital cost functions for each process unit and determine installation costs in Matlab
4. Summation of all the equipment and installation costs
5. Approximation of the total capital expenses by usage of the Percentage of Delivered-Equipment Cost method
6. Development of operating cost functions
7. Analysis of all expenses each year during the plant lifetime, including capital expenses (CAPEX) through depreciation and operating expenses (OPEX). All expenses are evaluated with the Net present value method

8. Estimation of the revenue each year during the plant lifetime

9. Evaluation of the economic key performance indicators

10. Sensitivity analysis performed by varying variables of choice

5 Process models

Two process models have been used in order to provide the data necessary to perform cost evaluations. Both are based on models developed by John, M. in his Master's thesis work in 2014 under supervision of Ohlemüller, P. [5]. The process model in Aspen plus is used to simulate the main CLC-plant components, with the steam cycle and CO_2-compression excluded. The steam cycle and the CO_2-compression unit is instead modeled in Ebsilon Professional 10. The connection points of where heat is transferred between the main CLC-setup and the steam cycle is shown in Figure 5.1. Initially, Ebsilon Professional 11 was planned to be used, but as the original model was made with the previous version of the program and could not be run accordingly with the latest version of the software, the decision was taken to continue working with Ebsilon Professional 10. For future users of the model, it should be made clear that it does not run properly with newer versions than Ebsilon Professional 10.

In order for the models in the two softwares to work in a combined way, the boundary conditions must be equal. For example, the entering and exiting energy streams represent the heat obtained and transferred to the steam cycle. Moreover, the temperatures and compositions of the streams existing in both models should be equal. To achieve this relationship, an iterative procedure is needed when matching the models. For instance, for a set of operating conditions in the Aspen model, certain stream compositions, temperatures, mass flows and heat duties are calculated. These results should then be entered in the Ebsilon, leading to new properties of some streams. The new properties should be entered in Aspen, which then gives rise to a new set of input variables that should be transferred to the Ebsilon model. This procedure should be carried on until convergence is obtained.

The two process models are listed in Table 5.1.

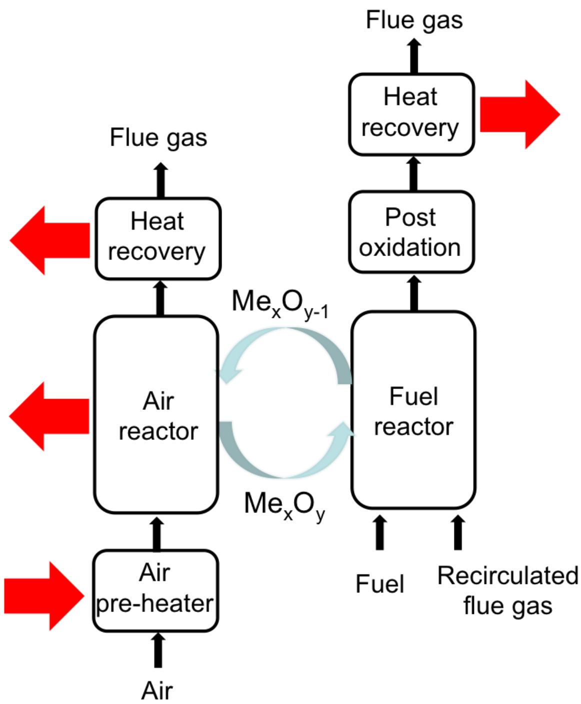


Figure 5.1: Heat transfer points between the steam cycle and the main CLC plant model

Table 5.1: The two process models

File name	Description
Modified FR 990 1GW full scale CLC plant 2 Zone.apwz	Referred to as "The Aspen model". Simulates the main components of a coal fired CLC plant, excluding the steam cycle
FR 990 full scale CLC Steam Cycle 25 June.ebs ‡	Referred to as "The Ebsilon model". Simulates the steam cycle and the CO_2-compression with heat recovery of the coal fired CLC plant that is modelled in the Aspen model

‡ Remark: It can only be run by Ebsilon Professional 10

5.1 Process model in Aspen Plus

As mentioned, the main process model is made with the software ASPEN PLUS, which is a program for simulation of chemical processes. The first step towards the current model was taken when a model to predict the behaviour of the 100 kW_{th} pilot CLC plant at Chalmers University of Technology was developed [5]. The model was validated with data obtained from an experimental campaign by Ohlemüller et al. [25]. That model was then used by John et al. [5] for developing an upscaled 1 GW_{th} version. When upscaling from 100 kW_{th} to 1 GW_{th}, naturally it is not as easy as multiplying all the model parameters with a factor of 10000. For instance, multiplying the height of 5m of the fuel reactor in the small scale model for the pilot plant at Chalmers, would result in a reactor height of 50000m. Therefore, careful consideration must be made when assuming plant dimensions of an up-scaled model. This was done by John et al. by comparison with experimental data, scientific literature and development of new a number of new FORTRAN codes. The magnitude of the streams were scaled up to meet the new power output requirement.

In [4], variations of the model have been tested, with respect to dimensions and temperatures of the reactors, oxygen carrier mass flow and fluidization medium. The version concluded as most suitable, denoted Var4 in [4], was chosen to select proper reactor dimensions and fluidization medium. Table 5.2 shows these reactor properties.

Table 5.2: Properties of the reactors, according to Var4 in [4]

Reactor properties	
AR cross sectional area	254 m^2
FR cross sectional area	176 m^2
AR temperature	1049 $°C$
FR temperature	990 $°C$
AR fluidization medium	Air
FR fluidization medium	Recirculated flue gas from FR

Figure A.1 in Appendix shows the flowsheet of the Aspen Plus model. Table A.2 and Table A.1 provide information about the stream and block abbreviations. In the CLC configuration chosen in this study, pyrolysis of coal takes place in the fuel reactor, but as it is not possible to include this into the fuel reactor (FR) component, it is represented by the decomposer (DECOMP), see Figure 5.2.
It was concluded by doing an overall heat balance that the default lower heating value (LHV) of coal in Aspen Plus provides a value too high, compared to experimental data of "El Cerrejón" coal from the Cerrejón mine in Colombia [5].

The default LHV value for coal shows a value as much as 6% higher than the actual experimental value. This could be due to different amounts of moisture. To correct this error, an energy stream equal to the surplus of LHV, is extracted from the stream that enters the fuel reactor. This is represented by the energy stream QHU in Figure 5.2.

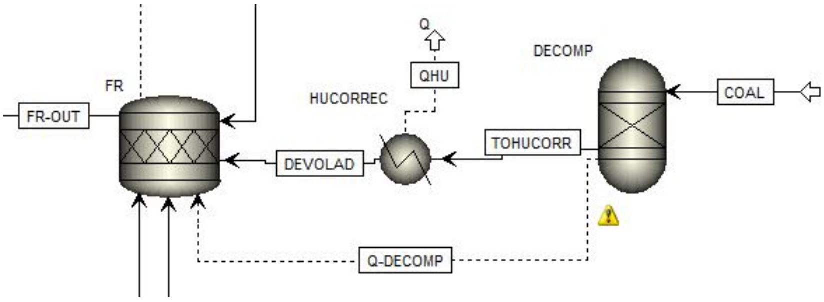


Figure 5.2: Pyrolysis of coal in the Aspen model

Both reactor temperatures are set to fixed values, but since the fuel reactor is modeled as an adiabatic reactor in contrast to the isothermal air reactor, the mass flow in circulation is adjusted to meet the desired temperature of the fuel reactor. The air reactor is isothermal with fluidized bed coolers installed to control the temperature. The fluidized bed cooler transfers the heat to evaporate and superheat the steam in the steam cycle.
The post oxidation chamber, POC, is for simplicity modeled as an adiabatic chamber, and the heated up flue gases are cooled down by dilution with recirculated flue gas, and afterwards used for superheating and pre-heating of the steam cycle. In a real case, the POC would probably be cooled stage-wise to maintain the temperature below a limit due to material constraints.
For further explanations about the model development and the FORTRAN codes of the blocks, the reader is directed to [5].

5.1.1 Modifications of the original Aspen Plus model

Only small justifications of the stream flows and/or temperatures have been done compared to the original model [5], in order to take into account the modifications that were made in the Ebsilon model. Furthermore, the energy stream UIOP is added to clarify that heat from the steam cycle is necessary to be provided to the air pre-heating, seen in Figure 5.1.

5.2 Process model in Ebsilon Professional

Ebsilon Professional is a software suitable for process simulations of steam cycles for power plants. It is used mainly for designing and optimizing thermodynamic cycles.
The design of a CLC plant implies a few major differences compared to a conventional coal fired power plant. First of all, it has two reactors instead of one, resulting in two hot flue gas streams. Since the purpose of CLC is to capture the CO_2 in the flue gases from the fuel reactor, the two flue gas streams must not be mixed. Moreover, a CO_2 compression section, consisting of several compressors and heat exchangers with cooling water system, is not present in a conventional plant. These factors makes the design more complex than for a conventional coal fired power plant, and a "normal" process flow configuration can not directly be applied.

The steam cycle configuration proposed in this work is based on the model first developed by John, M. et al [5] by comparison with a state of the art coal fired power plant designed by E.ON. The backbone of the model remains the same, but has been improved to be more feasible for a real plant design. This section describes the current model. Exactly what changes that have been made compared to the original model is discussed in Section 5.2.1. Figure B.1 in Appendix B shows the new steam cycle design developed in this study. Because of the large picture size, the important parts will be described with zoomed in figures here below.

The two flue gas streams from the CLC plant in the Aspen model work as input streams to the Ebsilon model, shown in Figure 5.3. The composition, mass flow and temperature is determined by the Aspen model and is entered as stream properties. Since the post oxidation chamber in the Aspen model is modeled as adiabatic, the fuel reactor flue gas temperature is high. To protect the construction materials from being exposed to harmfully high temperatures, part of the cooled down fuel reactor flue gas is recirculated and mixed with the hot flue gas before it enters the first heat exchanger. The temperature after the mixing point is set to a value of 1200° C, and the mass flow of the recirculation is adjusted thereafter.

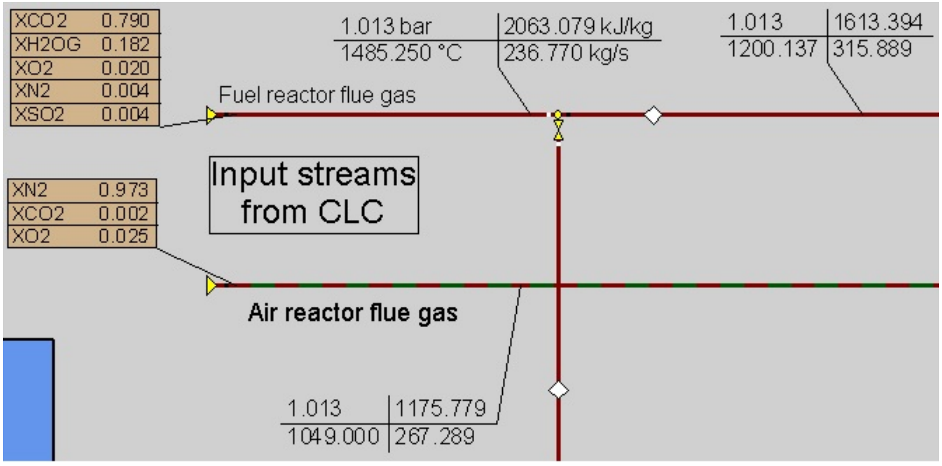


Figure 5.3: Flue gas input streams to the steam cycle

The feedwater is heated, evaporated and superheated by heat transferred from the both flue gas streams in a configuration shown in Figure 5.4. This arrangement was chosen after several trials, and was selected as the best choice with respect to efficiency and feasibility. The feedwater, represented by the blue line, enters from the bottom of the figure and is pre-heated twice by the FR flue gas stream and twice by the AR flue gas stream, to take use of the temperature gradients. A greater number of heat transfer units could have been chosen, but it increases the investment cost and the difficulty of construction. After being pre-heated, the feedwater is evaporated and superheated by the AR fluidized bed cooler, illustrated by a steam generator named "Evaporator AR tube walls" in the Ebsilon model. The outlet steam, represented by the red line, then enters two superheaters and reaches a temperature of 620° before entering the steam turbines. Before being released to the atmosphere, the AR flue gas stream is used to pre-heat the air stream that enters the AR, illustrated by the yellow line in Figure 5.4. When there is a choice, compressors are placed at locations where the stream being compressed has the lowest temperature. This to allow the lowest possible investment and operating costs, since it is related to the volumetric flow, which is directly proportional to the temperature.

At a temperature of 620° and a pressure of 270 bar, the steam enters a two-stage high pressure turbine, and is followed by a four-stage intermediate pressure turbine. The outlet steam is then lead to two parallel double flow low pressure turbines. The whole steam turbine arrangement was developed by John, M. in [5] and is shown in Figure 5.5. The steam properties are shown in Table 5.3.

Table 5.3: Properties of the steam entering the steam turbines

Type of steam	**Temperature** [$^\circ C$]	**Pressure** [bar]
High pressure steam	620	270
Intermediate pressure steam	620	75
Low pressure steam	212	3.5

Steam is extracted at different temperature and pressure levels to enable an efficient internal pre-heating of the feed water, which may be seen in Figure 5.6. Cooling water is necessary to condensate the lowest pressure steam. The terminal temperature differences have been defined with Table 5.4 as guideline. Note that the values for the terminal

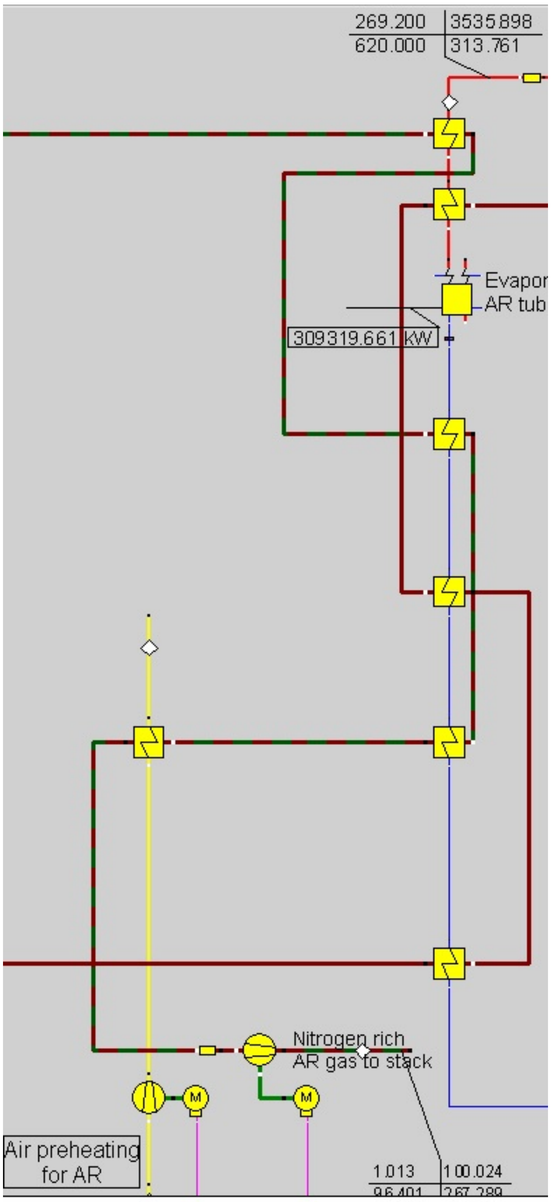


Figure 5.4: Flue gas path. Brown: FR flue gas stream, Brown/green: AR flue gas stream, Yellow: Air feed to the AR

temperature difference in the table are the minimum values used. Some of the specifications for some heat exchangers may be greater than the minimum values.

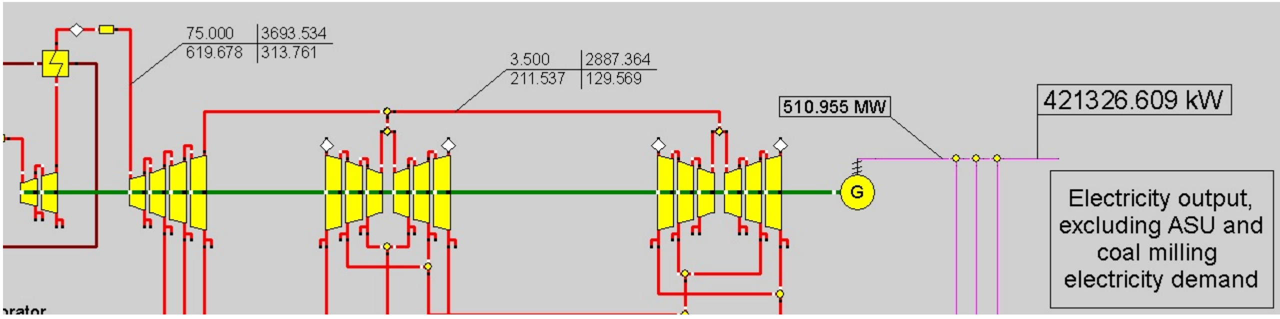


Figure 5.5: Steam turbine configuration

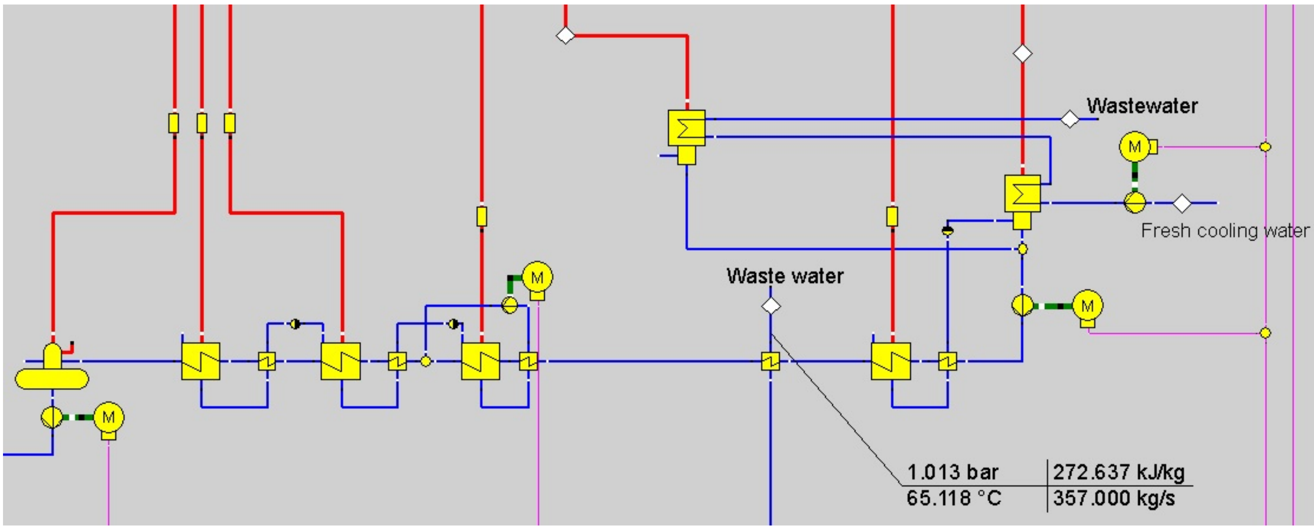


Figure 5.6: Feed water pre-heating system

Table 5.4: Guideline values for the heat exchanger specifications based on the information from [32]

Type of heat exchanger	**Minimum terminal temperature difference** $[°C]$	**Fluid on hot side / fluid on cold side**
Condensers	2	Condensating steam / liquid
Condensate cooler	7	Liquid / liquid
Superheaters	180	Gas / gas
Economizers	25	Gas / liquid
Gas coolers in the CO_2-compression unit	25	Gas / liquid
Air pre-heater	30	Gas / gas

The fluidization medium in the FR is in this model flue gas from the FR. This is taken from a split fraction after the last water pre-heater. The remaining flue gas is set off to the pipeline where the CO_2 is transported to the final storage. Before that, the CO_2 needs to be compressed and meet the purity conditions. The compression is performed with four compressors with subsequent coolers to separate water by condensation and to cool the compressed CO_2 to a

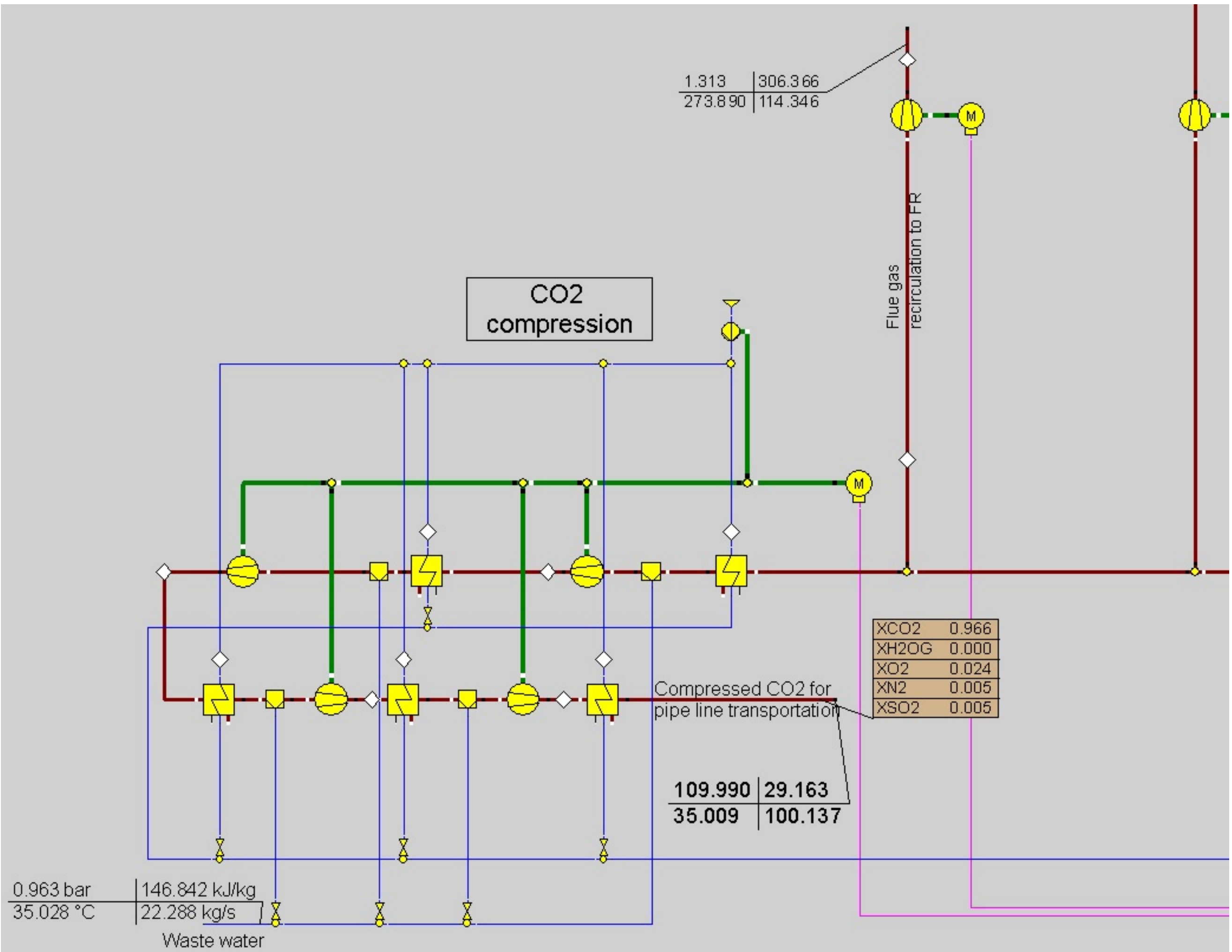


Figure 5.7: CO_2-compression section

sufficient transportation and storage temperature. Each compression stage has a compression ratio of four, according to recommendations written in Section 3.4. The CO_2-compression is an energy demanding process that is not present in a conventional coal fired power plant, and reduces the overall efficiency compared to the latter.

The two models together create the proposed plant flow sheet. In total, a number of 51 process units are included that have been taken into account when performing capital cost evaluations, shown in Table 5.5.

The most important process design parameters for the steam cycle are listed in Table 5.6. The live steam properties, i.e the properties of the steam that enters the first steam turbine, were decided after comparison with the already mentioned E.ON. power plant and consultation with employees at Institut Energiesysteme und Energietechnik (EST) of TU Darmstadt. The cooling water, assumed to be taken from a natural source as a river, has a temperature that naturally varies on a seasonal basis. Anyhow, $10°C$ has been chosen as a yearly average. The efficiencies of the turbines, generators, motors and pumps are chosen as the default values set by Ebsilon Professional.

Table 5.5: Main process units present in the proposed plant design

Process unit	**Number of units**
Air reactor	1
Fuel reactor	1
Post oxidation chamber	1
Air separation unit	1
Carbon stripper	1
Cyclone	2
Steam turbine	6
Compressor	8
Pump	5
Heat exchangers ‡	24
Feedwater tank	1

‡ Including superheaters, evaporators, feedwater pre-heaters, air pre-heater and condensers. More of a number of tube packs than individual heat exchangers.

Reasonable values for the pressure drops of the heat exchangers have been approximated individually for every heat exchanger side. As a very rough assumption, for liquids that are heated on the tube side, it is assumed that the pressure drop is proportional to 10 mbar bar per MW of heat transferred. Gas passing through the shell side is assumed to have a significantly lower pressure drop and is for simplicity set to 10 mbar for all heat exchangers. The pressure drop of condensing steam is assumed to be negligible. These assumptions are of course very rough, since there are many factors that affect the pressure drop; e.g. fluid properties as velocity and viscosity which are temperature dependent; design of the heat exchangers, as number of baffles, shape and size of the tubes; placement of the heat exchanger packages which affects the piping between them. Nevertheless, as the aim of the modelling is not to obtain an extraordinarily precise number of the predicted electricity output, these assumptions are seen as acceptable.
The results that design parameters together with the model provide, are presented in Section 5.3.

Table 5.6: Defined process design parameters

Process parameters		
Live steam pressure	270	bar
Live steam temperature	620	°C
Cooling water temperature	10	°C
Steam turbine isentropic efficiency ‡	0.88	-
Steam turbine mechanical efficiency ‡	0.998	-
Generator efficiency ‡	0.99	-
Motor electrical efficiency ‡	0.98	-
Motor mechanical efficiency ‡	0.98	-
Pump isentropic efficiency ‡	0.8	-
Pump mechanical efficiency ‡	0.998	-

‡ Default value from Ebsilon Professional

5.2.1 Modifications of the original Ebsilon Professional model

All in all, the system has been simplified and is more realistic with respect to equipment design specifications and also in the sense of realistic construction. The main changes are made to the heat exchangers, where the number of units has been reduced, and approximated pressure drops as well as realistic terminal temperature differences have been added. Naturally, by reducing the number of heat exchangers, the load of some heat exchangers are increased. As a consequence, the temperature of the CO_2-rich gas stream that enters the CO_2-compression section is higher, thus the cooling water demand increased by approximately 10 %. Additionally, an increased amount of heat is now lost through the waste water. Nevertheless, the advantages summoned by the modifications are considered to be greater than the drawbacks. A more detailed description of the modifications made is provided in the bullet list below. Component names that are used in this section refer to the abbreviations that are used in the original model [5].

• The recirculation of flue gas from the air reactor has been removed, as it was considered unnecessary.

• The temperature of the fuel reactor flue gas stream after the mixing point with the recirculated stream is fixed to 1200°C. 1200°C was considered to be a reasonable value for modelling purposes and to prevent the construction materials from being exposed to too high temperatures.

• The terminal temperature differences of the the heat exchangers were reviewed and changed to guideline values according to Table 5.4. Note that the stated values for the terminal temperature difference are the minimum values used. Some of the specifications made may be greater than the minimum values.

• A motor was missing for the work supply to a pump (Pumpe_2), and was therefore added and connected to the electricity grid.

• The cooling water mass flows for each heat exchanger in the the CO_2 compression unit are chosen so that the outlet water temperature is approximately 10°C below the boiling point, to avoid a two-phase flow. An exception is made in the last heat exchanger before the end output line (Wärmetauscher_15), where the outlet cooling water is 30°C below the boiling point, because of the need to cool down the CO_2-rich gas stream to a certain temperature for storage purposes [26].

• Pressure drops have been added to the heat exchangers where it is significant, according to what is written in

Section 5.2. Even though the pressure drop of a single heat exchanger isn't large, the total pressure loss in a whole heat exchanger network definitely is significant and affects the pump demand.

• A pump was added to the cooling water system in the CO_2-compression unit, to compensate for pressure drops on the cold side of the heat exchangers.

• The original steam outlet from the intermediate pressure steam turbine named IPT1 and the corresponding three heat exchangers it was connected to was removed, as it was found out that the overall efficiency was as high as possible when the mass flow of this stream was as low as possible. After careful consideration, even though steam cycles for some other coal fired power plants have a high pressure or intermediate pressure steam outlet to heat up the feed water, it was concluded that it did not have any essential role in this configuration, and was therefore removed. The reason to why it was not contributing to increase the net efficiency of the plant, could be that enough heat is recovered by the CO_2-compression and transferred to the water pre-heating, so that more of the steam instead can be dedicated to electricity generation.

• For the same reason as mentioned in the latter point, the first low pressure steam outlet, seen by the direction of water flow, from turbine LPT2_1A and LPT2_1B and the condenser and aftercooler connected to it was removed. Similarly as in the previous argument, the overall plant net efficiency peaked when those mass flows were as low as possible.

• Nachkühler_1 had a very small contribution to the heat recovery, and the equivalent heat could be recovered by increasing the load of surrounding heat exchangers, so also that component was removed. This particular component frequently caused errors when modifying the system, so another result was that the whole model could be modified easier after it was eliminated.

• The live steam temperature has been lowered from 650°C to 620°C, as this is a typical value for state-of-the-art steam cycles. Higher temperatures may cause damage for the materials that the process equipment is made of. This is also discussed in 5.2.

• The order of the last two economizers, by the direction of water flow, has been swapped. This because the temperature of the air reactor flue gas stream at that point is higher than the fuel reactor flue gas stream. Thermodynamically, it makes more sense to expose the stream to be heated to the hottest gas stream at the latest point. A result of this was that the following two heat exchangers that were heated by the fuel reactor flue gas stream could be merged into one bigger.

A selection of the most relevant results from the Aspen model are presented in Table 5.7. The fractions of heat generated by the respective reactors have been calculated by adding the heat recovered from the AR fluidized bed cooler with the heat extracted from the AR flue gas, and comparing with the heat recovered from the flue gas that origins from the FR and has passed the post oxidation chamber. The flue gas elemental compositions are separately shown in Table 5.8.

Table 5.7: Results from the Aspen model

Aspen model results		
Oxygen carrier make-up stream	34.6	$\frac{kg}{s}$
Oxygen carrier in circulation before losses	3640	$\frac{kg}{s}$
Air feed to AR	339	$\frac{kg}{s}$
Coal feed	41	$\frac{kg}{s}$
ASU pure oxygen demand	15	$\frac{kg}{s}$
Fraction of total heat generated by AR	60	%
Fraction of total heat generated by FR	40	%

Table 5.8: Flue gas compositions

(a) AR

Compound	**Fraction [mass-%]**
N_2	97.3
O_2	2.5
CO_2	0.2

(b) FR

Compound	**Fraction [mass-%]**
CO_2	79
H_2O	18.2
O_2	2
N_2	0.4
SO_2	0.4

The most important process data from the Ebsilon model is shown in Table 5.10. The gross electricity output is the output without any internal electricity demand withdrawn, as electricity consumed by compressors and the ASU unit etc. The properties of the compressed CO_2 that will be either transported or directly lead to a storage is also of high relevance, and is shown in Table 5.11. Note that a volumetric CO_2 flow of 0.54 $\frac{m^3}{s}$ at the stated temperature and pressure corresponds to a mass flow of approximately 100 $\frac{kg}{s}$.

Table 5.10: Results from the Ebsilon model

Ebsilon model results		
Live steam total flow	314	$\frac{kg}{s}$
Pre-heated air temperature	187	$°C$
AR flue gas to stack	96	$°C$
Waste water temperature	65	$°C$
Gross electricity output	511	MW_{el}
Total cooling water demand	12,4	$\frac{m^3}{s}$

Table 5.11: Compressed CO_2 properties

(a) Elemental composition

Compound	**Fraction [mass-%]**
CO_2	96.6
O_2	2.4
N_2	0.5
SO_2	0.5

(b) Other properties

Property	**Value**
Temperature	35 $°C$
Pressure	110 bar
Mass flow	0.54 $\frac{m^3}{s}$

By merging the results from the two models, an energy breakdown can be made to analyze how the input thermal energy from the fuel is distributed. A list of the energy related activities are listed in Table 5.13 and is illustrated graphically by Figure 5.8. The electricity demand for the air separation unit and the fuel milling, which is related to the operational expenses, is further described in Chapter 6. The electricity demand for the CO_2-compression which gives an overall plant efficiency decrease of 4% is in line with the predictions of 3-4% loss made by [3, 11]. The air separation unit gives an efficiency penalty of around 1%, which decreases the total plant electricity output down to 41%. The electricity demand of the fuel milling is excluded from Figure 5.8 due to its small contribution.

Table 5.13: Breakdown of the gross electricity output

Activity	$[MW_{el}]$
Gross electricity output ‡	511
Steam cycle compressor and pump demand ‡	- 49.7
CO_2-compression ‡	- 40.0
Air separation unit *	- 11.1
Fuel milling *	- 1.50
Net electricity output	Sum: 408.8

‡ Data from the Ebsilon model
* Calculated in the Matlab script

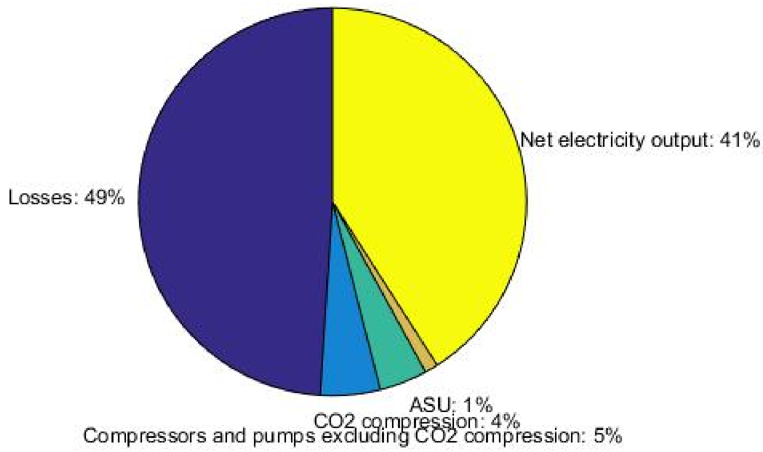


Figure 5.8: Breakdown of the $1000\ MW_{th}$ fuel input. The energy demand for fuel milling is excluded since its contribution is less than 0.15%

6 Economic evaluation of a chemical looping combustion plant

A cost estimation could be carried out in several different ways. One method is to do as Lyngfelt, A. does in [3] and estimate the additional equipment and operational costs of a CLC-plant compared to a conventional CFB coal fired power plant and then relate it to the amount of CO_2 that is captured. As an example, since the fuel reactor is adiabatic as written in Section 3.2, the additional costs of a CFB for CLC-purposes compared to a normal CFB can be approximated by estimating the costs of the insulation. This can be added up together with the cost of a post oxidation chamber, CO_2 compression units and other costs that are not present for a normal coal fired CFB power plant and then be related to the amount of CO_2 that is captured, to estimate a CO_2-avoidance cost. This is an accurate method in the sense that the uncertainty of the costs for base components as the steam cycle and indirect costs related to building a plant do not need to be considered.

However, another approach is chosen in this work. From the process charts developed in the models in Aspen Plus and in Ebsilon Professional, information is available about the number and size of the process units in the proposed plant. Cost functions that depend on dimensions, materials and design parameters have been defined for each process unit. These costs can then be summed up to get a number for the total equipment cost. From the total equipment cost, the total capital investment cost can be evaluated, which includes both direct and indirect costs, according to the method "Percentage of Delivered-Equipment Cost" in [31]. On top of that, interest payments, which is one of the big cost contributors seen over the whole plant lifetime, and asset retirement costs need to be added.
The reason to why this method was chosen, was partly because of learning objectives about plant economics, but also to be able to do a more detailed cost breakdown. By using this method, it is possible to compare the costs for each part of the plant with each other, and also compare the capital expenses to the operational expenses that were added later. Furthermore, the profitability can be investigated, and key parameters as levelized cost of electricity and payback-periods can be determined. Additionally, when having a function that calculates all the key performance parameters, trends throughout the plant lifetime can be analyzed. Attention must be paid when comparing annual costs from year to another. Since the value of one € today is not the same as in the future, due to inflation rates, price changes of commodities and expected profit for the investor, all the annual costs should be converted to net present value in order to be comparable. By summing up the plant lifetime costs, the levelized cost of electricity and CO_2-avoidance cost may be evaluated. The whole financial evaluation procedure is illustrated by Figure 6.1, and is described in detail in Section 6.3 and 6.4.

By following this procedure, it's also possible to conduct sensitivity analyses to see which design parameters, operating conditions and economic factors that influences the total costs the most. The whole economic evaluation, sensitivity analysis and additional calculations related to the plant operation is conducted by the Matlab files listed in Table 6.1.

Naturally, when doing a cost estimation like this, a number of assumptions need to be made. The most important assumptions are listed in Table 6.2. Some of the assumptions have been set as fixed, while some have been varied in the sensitivity analysis. Table 6.2 also indicates which ones that were fixed respectively variable.
The values depicted in Table 6.2 are denoted as the *base case*. All the key performance indicators are calculated with the base case values as input values. In the sensitivity analysis, the variables indicated as "varied" in Table 6.2 are adjusted upwards and downwards. Some of the data is extracted from the process models, and this is also shown in the table.

Plant lifetime, construction time and annual operating time
The plant lifetime is assumed to be 25 years with an annual operating time of 7500 hours, excluding the construction time of four years. Naturally, a power plant should be run as many hours as possible each year, to increase the revenue made by the sales of electricity if the electricity price is favourable. However, due to technical issues and improvements that are to be made, a certain time of the available operating time needs to be dedicated to maintenance. The annual operating hours may also be decreased due to political reasons. In some literature, it is suggested to include an annual decrease of operating hours because of a globally decreased usage of coal as an energy source [11]. In this work however, it has not been taken into account. The reason to this decision is that if the usage of coal for energy production decreases in the future [57], it is it is likely that firstly old plants and plants that release big amounts of CO_2 will be shut down, rather than uniformly decreasing the operational hours of all coal power plants, including the "clean" plants that release small amounts of CO_2. [11].

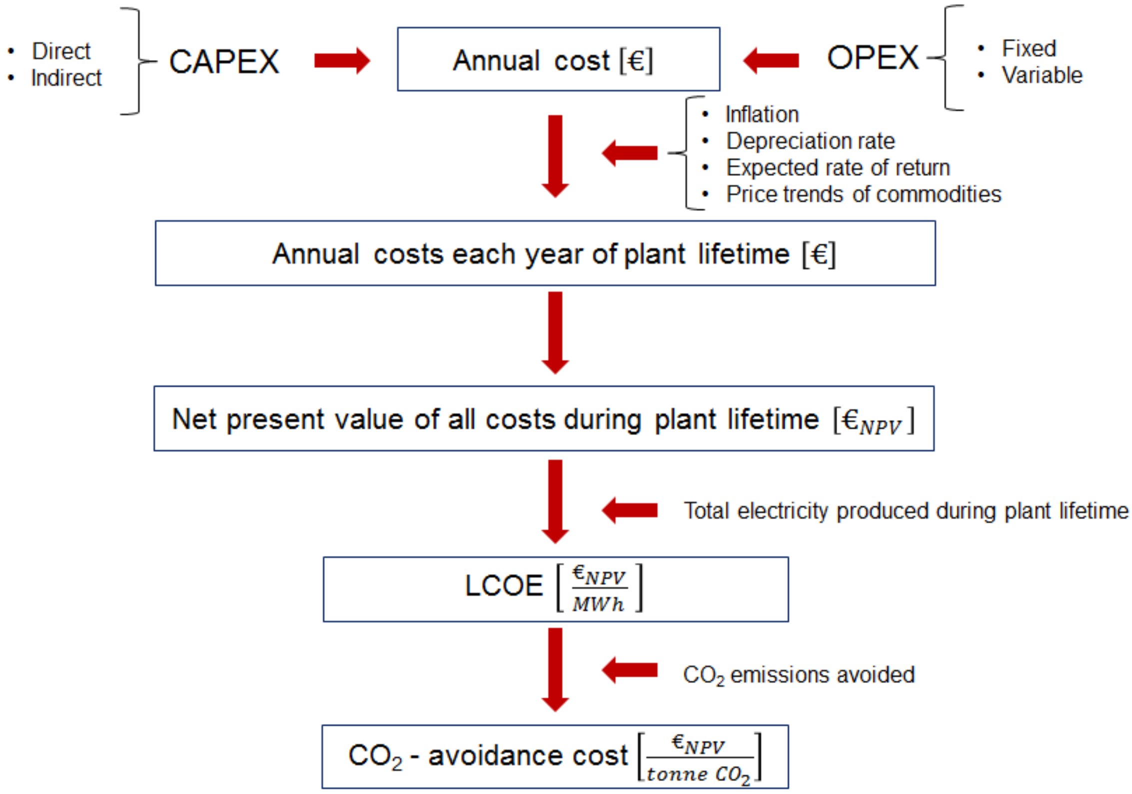


Figure 6.1: Cost evaluation procedure

Economic parameters
Currency exchange rates and prices of commodities fluctuate on a daily basis, so it's difficult to make accurate assumptions. The exchange rate between € and \$ has been selected as an average value from the beginning of 2017. The initial coal price has been chosen according to the forecast for 2018 by The World bank [55]. It is in the range of the coal price checked on 8 June, 2017 [52]. With time, the coal price is expected to increase. How this is modeled and why an annual increase of 0,8% is chosen is further described in Section 6.2.2.
The rate of return, also called imputed rate or discount rate, is a factor that a company assigns to reflect how big return it expects from an investment and is defined in Equation 7. The debt-to-equity ratio may also affect the magnitude of the assigned rate of return. In the base case, 8% has been set to the rate of return, and is a common value for economic evaluations [11]. The assigned rate of return has a big influence on the Levelized Cost of Electricity (LCOE). The assigned rate of return for asset retirement is related to the money that will be spent on retiring the plant components at the end of the plant lifetime, and has been set to 6%.
An important tool when estimating the investment costs of process equipment, is the Marshall & Swift cost index. It has been utilized to compare current prices of process equipment to historical reference objects or future values. In this work, a Marshall & Swift index for the steam power industry from the second quarter of 2011 is used [47]. More about Marshall & Swift cost indexes can be read about in Section 6.1.1.
The annual inflation rate has been assumed to be 2% in the base case. It is the inflation predicted for US economy until 2060 [51].
The electricity price is also a tricky value to assume, since it varies on a daily basis. However, the initial value has been chosen as 19,3% of the electricity price sold to end-user households in Germany in February 2017, which represents the selling price from power plants to the grid supplier. The remaining part of the end-consumer price is built up by grid charges (25,7%), renewable energy surcharge (23,6%) and a number of other surcharges [53]. The electricity price is assumed to increase with the same rate as the inflation rate.
The combined corporate tax rate for profits has been selected according to the tax level in Germany [56].

Table 6.1: Matlab files created for the economic evaluation, sensitivity analysis and additional calculations

File name	**Description**
BaseCase.m	Main script for the base case study. It estimates the plant net electricity output, capital expenses, operational expenses, annual costs, annual revenues, LCOE, CO_2-avoidance costs and provides illustrative figures
SensitivityAnalysis.m	Main script for the sensitivity analysis. Operational and economic parameters are varied to see how the LCOE, CO_2-avoidance costs and pay-back times are affected
Cost_evaluation_function.m	Function file used by SensitivityAnalysis.m
AR.m	Function file used by BaseCase.m and Cost_evaluation_function.m to evaluate capital costs of a AR
FR.m	Function file used by BaseCase.m and Cost_evaluation_function.m to evaluate capital costs of a FR
Compressors.m	Function file used by BaseCase.m and Cost_evaluation_function.m to evaluate capital costs of compressors
HeatExchangers.m	Function file used by BaseCase.m and Cost_evaluation_function.m to evaluate capital costs of stand-alone heat exchangers
Pump.m	Function file used by BaseCase.m and Cost_evaluation_function.m to evaluate capital costs of pumps

$$ir = \frac{earned\ capital}{invested\ capital} \cdot 100 \tag{7}$$

Where,

ir : Assigned rate of return (interest rate) [%]

Oxygen carrier properties and other
A capture efficiency of 98% is thought to be a reasonable assumption [6, 4], and the expected oxygen carrier lifetime, oxygen carrier price as well as the energy demand for fuel grinding has been taken from [3]. The extra cost for fuel grinding is due to the fact that a CLC process uses smaller coal particle sizes than in normal fluidized bed combustion of coal [3].
The CO_2 emissions factor has been calculated from the gas composition in the gasification stream in the Aspen Plus model.

Not used in this modeling but worth mentioning due to the relevance in this field
Industrial plants that emit above a certain amount of CO_2 are obliged to purchase a corresponding amount of CO_2 certificates. This is a tool to control the greenhouse gases released by the industry, and is supposed to create an economic driving force for "clean technology". CO_2 certificates can be be bought and sold in a market, and the price consequently varies [54].
There is a tax in Germany under the name *Verbrauchsteuern* that taxes the usage of coal and fossil fuels. For coal, the tax rate amounts to to 9.55 $\frac{€}{tonne}$. However, coal used for electricity generation is exempt from this excise duty, as the electricity already is taxed. There is always a risk that the taxes increase or that the legislation changes in a way that disfavours usage of coal and therefore the coal price was a crucial variable to vary in the sensitivity analysis, which indirectly could include tax changes. Other countries also have similar taxes for the usage of coal [58].

Table 6.2: Assumptions made for the cost evaluation

Process parameters			
Thermal energy output	1000	MW_{th}	Fixed
Gross electricity output ‡	511	$MW_{el,gross}$	Fixed
Net electricity output ‡	409	$MW_{el,net}$	Fixed
Plant lifetime	25	years	Varied
Construction time	4	years	Fixed
Annual operating time	7500	hours	Varied
Fuel demand (coal) *	40.6	$\frac{kg}{sec}$	Fixed
CO_2 capture efficiency	98	%	Varied
Oxygen carrier lifetime	300	hours	Varied
Oxygen demand for POC *	15.4	$\frac{kg}{sec}$	Fixed
FR exhaust gas flow *	221	$\frac{kg}{sec}$	Fixed
AR exhaust gas flow *	268	$\frac{kg}{sec}$	Fixed
Pressure drop of the AR *	30	kPa	Fixed
Pressure drop of the FR *	20	kPa	Fixed
Diamater of the AR *	18	m	Fixed
Diamater of the FR *	15	m	Fixed
Fuel grinding energy demand	10	$\frac{kWh}{tonne\ fuel}$	Fixed
Economic parameters			
Exchange rate between € and $	0,9	$\frac{€}{\$}$	Fixed
Coal price	60	$\frac{\$}{tonne}$	Varied
Annual coal price increase	0,8	%	Fixed
Oxygen carrier price (ilmenite)	175	$\frac{\$}{tonne}$	Varied
Assigned rate of return	8	%	Varied
Assigned rate of return for asset retirement costs	6	%	Fixed
Marshall & Swift cost index	1546.5	-	Fixed
Inflation rate	2	%	Varied
Selling electricity price to grid	0.056	$\frac{€}{kWh}$	Varied
Corporate tax	32	%	Varied
Other input data			
LHV bituminous coal from Cerrejón *	24.64	$\frac{GJ}{tonne}$	Fixed
CO_2 emission factor for coal	350	$\frac{kg\ CO_2}{MWh_{th}}$	Fixed

‡ Data from the Ebsilon model
* Data from the Aspen Plus model

6.1 Capital expenses

As mentioned early in Chapter 6, the CAPEX is evaluated from the the total equipment cost. The following subsections describe how this is done, and a number of important tool when defining cost functions for process units.

6.1.1 Cost indexes

A cost index is a number that indicates how the cost of industrial equipment at a given time has changed relative to a certain reference time. Cost indexes are used to perform estimates of equipment cost if the corresponding values are known from the past. The present cost of a piece of equipment can be estimated by multiplying the equivalent cost from the past by the ratio of the present index to the index value applicable when the reference value was taken, as shown in Equation 8. [31]

$$Present\ cost = Original\ cost \cdot \left(\frac{index\ value\ at\ present}{index\ value\ at\ time\ original\ cost\ was\ obtained} \right) \tag{8}$$

Cost indexes naturally only provide estimates of the price range, but are fairly accurate estimations if the time comparison period is less than ten years. Cost indexes are also used to extrapolate costs into the near future, to predict expenses at a planning phase, but it is not recommended to do this for a future period of more than 2-3 years [39].

There are several different cost indexes published regularly. All with the objective to estimate costs of various parameters as equipment cost, labor cost, construction cost, material price etc. A commonly used cost index and the one that will be used in this work is the Marshall & Swift index. The Marshall & Swift has however now stopped being published since April 2012. In this work, a Marshall & Swift index for the steam power industry from the second quarter of 2011 is used, which values to 1546.5 [47].

6.1.2 Guthrie's formulae

Cost indices have a vital role in Guthrie's formulae, which are set of non-linear regressed formulae to estimate equipment and installation costs for industrial equipment. Guthrie's formulae also take into account the type of equipment, materials, dimensions, operating pressure and difficulty of installation, and hence can have many different forms. The general Guthrie's formula looks like (9). The installation costs may in some cases be included in d_2, while in some other cases it may be represented by an additional factor. The error when using Guthrie's method for cost estimation may be up to 15-25% [44, 39, 13].

$$IC = a \cdot \frac{M\&S}{280} \cdot L_1^b L_2^b d_1 d_2 \tag{9}$$

Where,

IC :	Investment cost of equipment
a:	Specific constant for each industrial equipment
280:	Marshall & Swift cost idex at year 1969
$M\&S$:	Present Marshall & Swift cost idex
L_1, L_2:	Equipment dimension factors
d_1:	Operating pressure factor
d_2:	Material factor

6.1.3 Six-tenths factor rule

If the cost data for a piece of equipment with a certain size or capacity is not available, predictions can be made with the *six-tenths factor rule* (10). [31]

$$IC_a = IC_b \cdot \left(\frac{size_a}{size_b}\right)^m \tag{10}$$

Where,

IC_a :	Investment cost of unit a
IC_b:	Investment cost of unit b
$size_a$:	Size of unit a
$size_b$:	Size of unit b
m:	Exponent specific for each piece of equipment

0.6 is a general number for the exponent m of the capacity quotient, but it varies for different pieces of equipment. It is in the range of 0.3-1.2, but in most cases it is under 1.0 and close to 0.6. In absence of other information, 0.6 can be used as exponent. It is not recommended to utilize the six-tenths factor rule beyond a 10-fold range of capacity, because of increased uncertainty. Furthermore, it is important that the two pieces of equipment being compared are similar with regard to type of material, construction, operating range of temperature and pressure. Nevertheless, the six-tenths factor rule is widely used in approximations of industrial equipment and even total process costs.

6.1.4 Equipment and installation costs functions for the proposed plant

Table 5.5 provides a list of the main process units that costs need to be estimated for. How the cost for each unit is estimated is described in detail in this section. To be consistent, the equipment installation costs are included in all these cost functions.

Fuel Reactor

The investment cost of the fuel reactor has been approximated through comparison with a reference CFB available in [12]. In this case, a 445 $MW_{el,brutto}$ Oxyfuel CFB has been selected as a reference. Information about this reference plant is provided in [12, 11]. The cost of the reference CFB is 115 million €. As mentioned in Section 3.2, the fuel reactor in a CLC plant will be designed to be adiabatic without fluidized bed coolers, to maintain a high operating temperature. In the reference CFB, a fluidized bed cooler is included in the price, so a correction factor, f_{FR} has been introduced to disregard the cost contribution of the fluidized bed cooler. It has initially been assumed to be 0.85, as done in [12]. The total cost of the fuel reactor is built up by the cost contributions of the combustion chamber, the integrated heat exchangers and the solid inventory. The cost of the combustion chamber is estimated by Equation 11, which is the six-tenths factor rule (Equation 10) with an additional factor to disregard the cost of the fluidized bed cooler.

$$IC_{FR,chamber} = \left(\frac{\dot{m}_{flue,FR}}{\dot{m}_{flue,ref}}\right)^{0.6} \cdot IC_{ref} \cdot f_{FR} \tag{11}$$

Where,

$IC_{FR,chamber}$: Investment cost of the fuel reactor chamber

IC_{ref}: Investment cost of the reference CFB

$\dot{m}_{flue,FR}$: Flue gas mass flow of the fuel reactor

$\dot{m}_{flue,ref}$: Flue gas mass flow of the reference CFB

f_{FR}: Fuel reactor correction factor

Estimating the costs for the superheater and economizer tubes integrated with the fuel reactor flue gas path is trickier. A first assumption is that radiative heat transfer can be neglected in comparison to the convective heat transfer since the only heat exchanger tubes exposed to direct radiation are the ones in the front row. This is not a perfect assumption because radiative heat transfer may have a non-negligible contribution to heat transfer at the high temperatures that are present in a combustion chamber, and the front row tubes that are exposed to radiation may heat up the second row tubes by radiation and so on. Anyhow, this assumption is made to simplify the cost evaluation and the heat transfer tubes don't have a significant impact on the total cost, which is shown in Section 6.6.

Superheater tubes

Since the fluids on both heat transfer sides are gases, the heat transfer properties are similar. Therefore unfinned tubes have been selected as a suitable tube construction [16]. Further assumptions are displayed in Table 6.3. The material of the superheater tubes has been selected as SA213-T91 according to information given in [46] about suitable materials for heat exchanger tubes. The outer diameter of the tubes has been assumed to be 5cm, and the wall thickness 1cm according to [16]. By taking use of the coefficient $U \cdot A$ calculated by Ebsilon Professional, and with an overall heat transfer coefficient, U, the total necessary outer surface area, A, can be calculated. U is assumed to be $7 \frac{W}{m^2 K}$. Furthermore, when the total necessary outer surface area is known, the total length of the tubes can be estimated by Equation 12. An additional factor, f_{pipes} is added to take into the curvature of the pipes.

Table 6.3: Superheater and economizer tube design

	Outer surface type	**Material**	**Outer diameter [mm]**	**Wall thickness [mm]**	**U** $\left[\frac{W}{m^2 K}\right]$
Superheater tubes	Unfinned	SA213-T91	50	5	7
Economizer tubes	Finned	SA213-T11	50	5	25

$$l_{superheaters} = \frac{A_{superheaters}}{d_{outer\ wall} \cdot \pi} \cdot f_{pipes} \tag{12}$$

Where,

$l_{superheaters}$: Total length of all superheater tubes

$A_{superheaters}$: Total heat exchange area of the superheaters

$d_{outer\ wall}$: Outer diameter of the superheater tubes

f_{pipes}: Factor to take into account the pipe curvatures

With the total length, diameter and thickness of the tubes known, the total weight can be determined by Equation 13.

$$w_{superheaters} = \rho_{superheaters} \cdot l_{superheaters} \cdot \left(\left(\frac{d_{outer\ wall}}{2} \right)^2 - \left(\frac{d_{inner\ wall}}{2} \right)^2 \right) \cdot \pi \tag{13}$$

Where,

$w_{superheaters}$: Total weight of all superheater tubes

$d_{inner\ wall}$: Inner diameter of the superheater tubes

Finally, the total cost contribution by the superheater tubes can then be evaluated by multiplication with the current price of superheater tubes of the chosen material SA213-T91, which is approximated to be 1000 $\frac{\$}{ton}$ from an active supplier, which equals to 907.2 $\frac{\$}{tonne}$. [50].

Economizer tubes
The costs for the economizer tubes are estimated in an equivalent way as the superheaters. The assumed tube characteristics are listed in Table 6.3 and the total weight is calculated similarly by Equations (14,15). From that, the cost can be evaluated by multiplication with the material cost of 907.2 $\frac{\$}{tonne}$ from an active supplier [49]. A suitable material has been found to be SA-210-C [46], but since reliable price data was easier found about SA-213-11, which is assumed to be of the same price range, it was selected as the material price. A factor named f_{fins} is multiplied with the weight to take into account the increased material costs by the usage of finned tubes, assumed to be used for economizers.

$$l_{economizer} = \frac{A_{economizer}}{d_{outer\ wall} \cdot \pi} \cdot f_{pipes} \tag{14}$$

$$w_{economizer} = \rho_{economizer} \cdot l_{economizer} \cdot \left(\left(\frac{d_{outer\ wall}}{2} \right)^2 - \left(\frac{d_{inner\ wall}}{2} \right)^2 \right) \cdot \pi \cdot f_{fins} \tag{15}$$

Solid inventory

The filling of the fuel reactor with oxygen carrier is added as an investment cost. The solid inventory, i.e the amount of oxygen carrier that the fuel reactor should be filled with, is calculated by Equation (16). The pressure drop and diameter of the fuel reactor, through which the cross sectional area is calculated, is available in the Aspen Plus model and displayed in Table 6.2. The assumed pressure drops are of the same magnitude as what is predicted in [3]. 20% of m_{si} is added to take into account the oxygen carriers in circulation between the two reactors. An equivalent calculation is made to estimate the solid inventory in the air reactor.

$$\Delta P = \frac{F}{A} = \frac{m_{si} g}{A} \quad \longrightarrow \quad m_{si} = \frac{\Delta P \cdot A}{g} \tag{16}$$

Where,

ΔP : Pressure drop of the fuel reactor

F: Force

A: Cross sectional area of the fuel reactor

m_{si}: Solid inventory

g: Acceleration of gravity

Total
The total cost of the fuel reactor is given by Equation (17).

$$IC_{FR} = IC_{FR,chamber} + IC_{FR,sh} + IC_{FR,eco} + IC_{FR,si} \tag{17}$$

Where,

$IC_{FR,tot}$:	Total investment cost of the fuel reactor
$IC_{FR,chamber}$:	Investment cost of the fuel reactor chamber
$IC_{FR,sh}$:	Investment cost of the superheater tubes
$IC_{FR,eco}$:	Investment cost of the economizer tubes
$IC_{FR,si}$:	Investment cost of the air reactor solid inventory

Air Reactor

In a similar way as for the fuel reactor, the investment cost of the air reactor chamber has been approximated by comparison with a reference coal fired circulating fluidized bed available in [12]. The investment cost of the reference CFB is 160 million € and includes a fluidized bed cooler, similar to an air reactor designated for CLC. Furthermore, the oxygen source for the reference CFB is air, and thus the dimensions of the chamber are larger than for a CLC-CFB. It is therefore more suitable to estimate the investment cost by comparing the magnitude of the flue gas streams, rather than the power outputs. The investment cost for the reference CFB is simply multiplied by the ratio of the flue gas streams of the modelled AR and the reference CFB to estimate the cost of the chamber with the built-in fluidized bed cooler, according to what is shown by Equation 18. In other words, it is the six-tenths factor rule (10) with an exponent of 0.6. The fluidized bed cooler works as the boiler and partly a superheater for the steam cycle. The boiling is expected to occur in a temperature range present in the fluidized bed cooler.

$$IC_{AR_{chamber}} = \left(\frac{\dot{m}_{flue,AR}}{\dot{m}_{flue,ref}}\right)^{0.6} \cdot IC_{ref} \tag{18}$$

Where,

$IC_{AR,chamber}$:	Investment cost of the air reactor chamber, including fluidized bed cooler
IC_{ref}:	Investment cost of the reference CFB
$\dot{m}_{flue,AR}$:	Flue gas mass flow of the air reactor
$\dot{m}_{flue,ref}$:	Flue gas mass flow of the reference CFB

In the same way as for the fuel reactor, the costs for the superheater tubes, economizer tubes and the solid inventory have been estimated by Equations (12,13,14,15,16), with the corresponding data for the fuel reactor. The same assumptions about the heat transfer tubes as for the fuel reactor are made.

Air pre-heater
The air pre-heater is supposed to be of a Ljungström type, but due to lack of price information, the cost has been approximated in a similar way as for the superheater and economizer tubes. It is seen as a conservative approach since the price of a Ljungström heat exchanger should be equal or less to an equivalent capacity tube pack for air pre-heating. Otherwise there would not be an economic driving force to select a Ljungström type instead of simple tube packs in the flue gas path. An overall heat transfer coefficient of 20 $\left[\frac{W}{m^2K}\right]$ has been chosen.

Total
The total cost of the air reactor is given by Equation (19).

$$IC_{AR} = IC_{AR,chamber} + IC_{AR,sh} + IC_{AR,eco} + IC_{AR,ap} + IC_{AR,si} \tag{19}$$

Where,

$IC_{AR,tot}$:	Total investment cost of the air reactor
$IC_{AR,chamber}$:	Investment cost of the air reactor chamber, including fluidized bed cooler
$IC_{AR,sh}$:	Investment cost of the superheater tubes
$IC_{AR,eco}$:	Investment cost of the economizer tubes
$IC_{AR,ap}$:	Investment cost of the air pre-heater
$IC_{AR,si}$:	Investment cost of the air solid inventory

Post oxidation chamber and additional insulation

Costs for a post oxidation chamber and insulation for cyclones, ducts and the fuel reactor walls are additional costs compared to a normal coal fired CFB power plant [3]. The cost for this has been approximated to 3.7 million € for a $1MW_{th}$ CLC power plant [3]. A factor of 1.4 has been multiplied with this number to add installation costs.

Air separation unit

An estimation of the investment cost has been made by the sixth-tenths factor rule (10), with an air separation unit in [12] as reference which has a capacity of producing 76.3 $\frac{kg\ O_2}{s}$ at an investment cost of 83.3 M€.

$$IC_{ASU} = IC_{ASU,ref} \cdot \left(\frac{capacity_{ASU}}{capacity_{ASU,ref}} \right)^{0,6} \tag{20}$$

Where,

IC_{ASU}:	Investment cost of the air separation unit
$IC_{ASU,ref}$:	Investment cost of the reference air separation unit
$capacity_{ASU}$:	Production capacity of the air separation unit
$capacity_{ASU,ref}$:	Production capacity of the reference ir separation unit

Carbon Stripper

Difficulties arise when estimating a cost for a carbon stripper, since it's a specific component only used in CLC and not much research has been done about the design nor the costs of such a unit. However, since it is indirectly connected to the fuel reactor, it has been approximated as 20% of the fuel reactor combustion chamber cost.

Cyclones

According to [45], the investment costs for cyclones are low, but since the flue gas streams from the two respective reactors in the proposed CLC plant are relatively big, the investment costs for the cyclones have not been neglected. Due to lack of information found about cyclone costs, the two cyclones have been assumed to cost 10% of the air reactor and 10% of the fuel reactor chamber respectively.

Steam turbines and generators

The costs for steam turbines and generators have been approximated by the six-tenths factor rule (21) with a coal plant provided by E.ON engineering in 2008 [11] as a reference. $m = 0.9$ has been selected as a conservative assumption. Relevant data for the reference plant is shown in Table 6.4.

$$IC_{sg,tot} = IC_{sg,ref} \cdot \left(\frac{P_{el,gross}}{P_{el,gross,ref}} \right)^{0.9} \tag{21}$$

Where,

$IC_{sg,tot}$:	Investment cost for the a steam turbines and generators [€]
$IC_{sg,ref}$:	Investment cost of steam turbines and generators in the reference plant
$P_{el,gross}$:	Gross electricity output of proposed CLC plant
$P_{el,gross,ref}$:	Gross electricity output of reference CLC plant
m:	0.9

Table 6.4: E.ON reference plant [11]

E.ON reference plant		
Electricity output	600	$MW_{el,gross}$
Gross electrical efficiency	47.7	%
Net electrical efficiency	45.6	%
EPC cost for steam turbines and generators	102	million €
Total EPC cost	510	million €
Total EPC cost per thermal power	0.41	$\frac{million\ €}{MW_{th}}$

Compressors

The investment cost for the compressors has been calculated with a version of Guthrie's formula (22). F_p varies between 1-2 depending on the pressure drop. For pressure drops below 3.45 bars which is the case on most compressors in the proposed plant layout, F_p is equal to 1. F_m is a material factor that is chosen to 1.15 for carbon steel with thermal insulation. More about Guthrie's formulae is written in Section 6.1.2. [40]

$$IC_{comp} = 517.5 \cdot \frac{M\&S}{280} \cdot (bhp)^{0.82} \cdot (2.11 + F_p \cdot F_m) \quad (22)$$

Where,

IC_{comp}: Investment cost for a compressor [€]

$M\&S$: Marshall & Swift cost index

bhp: Power of the motor shaft [hp]

F_p: Operating pressure factor

F_m: Material factor

Pumps

For simplicity, all pumps are assumed to be centrifugal pumps. This assumption has been made by considering the pump types and tables of application suitability available in [31], due to its high capacity limit and cost efficiency. The costs for each individual pump has been estimated by graphical inter/extrapolation from capacity-cost graphs in [31].

Heat exchangers

To evaluate the investment cost for condensers and the heat exchangers in the CO_2-compression section and water preheating, general overall heat transfer coefficients, U, must be assumed. This to calculate the sizes of the heat exchangers, since the parameter $U \cdot A$ is given in the Ebsilon model. Reasonable values for U have been assigned by comparison with guideline values in literature [31, 36] and are shown in Table 6.5.

Table 6.5: Assigned overall heat transfer coefficients

Fluid on hot side	**Fluid on cold side**	$\boldsymbol{U}\ \left[\frac{\boldsymbol{W}}{\boldsymbol{m^2K}}\right]$
Forced convection gas	Forced convection water	30
Condensing steam	Forced convection water	2500
Forced convection water	Forced convection water	2000

With U and the heat transfer area known for each heat exchanger, a version of Guthrie's formula (23) can be utilized to estimate the cost for each unit. F_p is a factor depending on the operating pressure and has been chosen as 0.1. This is valid for pressures below 20 bars. F_m has been selected as 1 because carbon steel is assumed to be acceptable on both heat transfer sides. It might be the case that another material must be used, and F_m can then be significantly higher. However, it is shown in the results that the cost contribution by heat exchangers is relatively small. Lastly, the heat exchanger type is assumed to be floating head and F_d is therefore 1. [42]

$$IC_{HX} = \frac{M\&S}{280} \cdot 101.3 \cdot A^{0.65} \cdot (2.29 + F_m \cdot (F_d + F_p)) \tag{23}$$

Where,

IC_{HX}:	Investment cost for a heat exchanger [€]
$M\&S$:	Marshall & Swift cost index
A:	Heat exchanger surface area $[ft^2]$
F_p:	Operating pressure factor
F_m:	Material factor
F_d:	Heat exchanger type factor

Feedwater tank and deaeration unit

The investment cost of the feedwater tank and deaeration unit has been estimated by Guthrie's formula for a pressurized tank [40, 41]. This formula provides similar results as the cost graphs for pressurized vessels in [31]. A cylindrical pressurized tank with a radius of 2m and a length of 10m has been assumed. Carbon steel has been chosen as material and the operating pressure is supposed to be 17.5 bar as indicated by the Ebsilon model, resulting in an F_p value of 1.2 and F_m equal to 1. The Guthrie's formula for pressurized tanks has been multiplied by a factor of 2, to include additional costs related to stirring mechanisms, pump, valves, instrumentation and control systems that do not apply for a normal pressurized tank, as shown in Equation (24).

$$IC_{fwt} = 101.9 \cdot \frac{M\&S}{280} \cdot d^{1.066} \cdot h^{0.802} \cdot (2.18 + F_p \cdot F_m) \tag{24}$$

Where,

IC_{fwt}:	Investment cost for the feedwater tank [€]
d:	Diamater [ft]
h:	Height or length [ft]
F_p:	Operating pressure factor
F_m:	Material factor

CO2 compression

The investment cost for the heat exchangers, compressors and pump in the CO_2 compression section is calculated according to the cost functions for the respective units described in this chapter.

Other capital expenses

The equipment cost has a major contribution to the total capital expenses, but there are other direct costs and indirect costs that also need to be considered. These are listed in Table 6.6.

Table 6.6: Direct and indirect costs included in the capital expenses

Direct capital costs	**Indirect capital costs**
Equipment cost	Engineering and supervision
Equipment installation	Construction expenses
Electrical systems	Legal expenses
Piping	Contractor's fee
Buildings	Contingencies
Yard improvements	Start-up costs
Spare parts	(Interest payments)
Service facilities	
Land	
Instrumentation and control systems	

Equipment installation
Installation of purchased equipment usually costs in the range of 25-55% of the purchased equipment cost. In the cost functions described in Section 6.1.4, the installation costs are included. Expenses for insulation of equipment are often included in the installation costs. [31]

Electrical systems
Major electrical systems as power wiring, lighting and transformation are included in the electrical system costs. [31]

Piping
The piping costs covers pipes, valves, labor, supports and fittings connected to the pipes. This cost is higher for processes involving fluids than for solids. Expenses for piping usually has a significant contribution to the total capital costs. The material costs for the piping vary widely, depending on the operating temperatures and pressures. [31]

Buildings
Material costs, services, plumbing, heating, ventilation and all expenses related to the erection of all buildings of the plant are included here. Differently from the piping expenses, the cost for buildings of solid process plants is higher than for liquid process plants. [31]

Yard improvements
Another infrastructural group of costs are yard improvements. This includes landscaping, roads, sidewalks, fencing and grading. [31]

Service facilities
Under the general heading of service facilities, utilities for power, steam, water and fuel are included. Also waste disposals, fire protections items, warehouses and service facilities for the working force as shops and canteens are part of this heading. [31]

Land
Cost of land for industrial plants can by law not be depreciated and must usually be seen as a one-time investment at the beginning of construction. Therefore, it is sometimes not included in the list of fixed capital expenses. [31]

Instrumentation and control systems
Most process equipment require instrumentation and control systems. Instrument costs, installation labor costs and expenses of auxiliary equipment contributes to this cost. These expenses naturally depend on how big amount of control is required and varies with a big magnitude. [31]

Engineering and supervision
The heading is straightforward, and covers costs for construction design and engineering.

Construction expenses
An additional indirect plant cost is the one for construction and includes temporary construction; rental for construction tools; temporary offices, taxes and insurance for the workers.

Legal expenses
This is a cost necessary to prove compliance with government, environmental and safety requirements. Also expenses related to contracting are under this category.

Contractor's fee
The fee of contractors varies from situation to situation, but is a relatively small fraction of the total cost.

Contingencies
There is always a risk that unexpected events occur, such as floods, storms strikes, transportation accidents, design changes and other unforeseen incidents. Therefore, a portion of the total invested money needs to be dedicated to this.

Start-up costs
It always takes some time and modifications in a start-up phase of a new plant before reaching full capacity and finding the right operating conditions. This means additional costs, or perhaps rather a loss of profit. The cost for this has been estimated as 25% of the fuel capacity for one month to cover inefficient operation during the start-up phase and three months of operating and maintenance labor costs to include staff training. [18]

Interest payments
The magnitude of the interest payments depends on the rate of return that the investor assigns, explained in the beginning of Chapter 6. It represents a significant portion of the total plant costs. However, it is in many cases not listed among capital costs but rather as a stand-alone expense.

Total investment cost

Performing an estimation of these costs is not an easy thing. As mentioned in Section 4, the "Percentage of Delivered-Equipment Cost" [31] has been utilized, with guidelines from the Global CCS Institute about how to perform cost estimations of CCS technologies [18]. The step-by-step procedure in Chapter 4 has been followed. Firstly, the total equipment and installation cost is calculated by adding the contribution of all process components according to Equation (25).

$$IC_{eq,inst} = + IC_{AR} + IC_{FR} + IC_{POCplus} + IC_{ASU} + IC_{CS} + IC_{cyc,tot} + IC_{sg,tot} + IC_{comp,tot} + IC_{pump,tot} + IC_{HX,tot} + IC_{fwt} \quad (25)$$

Where,

$IC_{eq,inst}$:	Total equipment and installation cost
IC_{AR}:	Investment cost for the air reactor
IC_{FR}:	Investment cost for the fuel reactor
$IC_{POCplus}$:	Investment cost for the post oxidation chamber and additional insulation
IC_{ASU}:	Investment cost for the air separation unit
IC_{CS}:	Investment cost for the carbon stripper
$IC_{cyc,tot}$:	Investment cost for the cyclones
$IC_{sg,tot}$:	Investment cost for the steam turbines and generators
$IC_{comp,tot}$:	Investment cost for the compressors
$IC_{pump,tot}$:	Investment cost for the pumps
$IC_{HX,tot}$:	Investment cost for the heat exchangers
IC_{fwt}:	Investment cost for the feedwater tank

For CCS technologies, the contingency costs are expected to be in the range of 10% of the installed equipment cost according to IEAGHG's recommendations [18]. The owner costs and expenses for spare parts are then derived from the sum of the installed equipment and contingency costs, according to (26-29). The owner costs cover most of the expenses in Table 6.6. In this study, the owner costs are assumed to be 7% of the sum of equipment, installation and contingency costs according to IEAGHG's recommendations [18]. However, there are indications that this number should be significantly higher. Therefore, it is strongly varied upward in the sensitivity analysis.

$$IC_{cont} = 0.1 \cdot IC_{eq,inst} \quad (26)$$

$$IC_{oc} = 0.07 \cdot (IC_{eq,inst} + IC_{cont}) \quad (27)$$

$$IC_{sp} = 0.005 \cdot (IC_{eq,inst} + IC_{cont}) \quad (28)$$

$$IC_{su} = 0.01 \cdot (IC_{eq,inst} + IC_{cont}) + 0.25 \cdot fco + 3 \cdot op \quad (29)$$

Where,

IC_{cont}:	Expected contingency costs
IC_{oc}:	Owner costs
IC_{sp}:	Costs for spare parts
IC_{su}:	Start up costs
fco:	Fuel cost for one month
op:	Monthly operating and maintenance labor costs, to cover training costs

Interest during construction is also something that must be considered in the investment cost. Naturally, the construction expenses are not equally distributed during the construction time. As mentioned in Table 6.2, it is assumed that the construction time is four years and the distribution of the expenses during that period is assumed to be equivalent as in [11] for the construction of a carbonate looping process, according to Table 6.7. The total cost for interest during construction is then estimated by Equation 30.

Table 6.7: Distribution of the capital expenses during the construction time

Year of construction	1	2	3	4
Percentage of payment	15%	30%	35%	20%

$$\begin{aligned} IC_{idc} = & x_1 \cdot (IC_{eq,inst} + IC_{cont}) \cdot (1+ir)^{ct} + x_2 \cdot (IC_{eq,inst} + IC_{cont}) \cdot (1+ir)^{ct-1} + \\ & x_3 \cdot (IC_{eq,inst} + IC_{cont}) \cdot (1+ir)^{ct-2} + x_4 \cdot (IC_{eq,inst} + IC_{cont}) \cdot (1+ir)^{ct-3} \\ & - (IC_{eq,inst} + IC_{cont}) \end{aligned} \tag{30}$$

Where,

x_n: Percentage of payment at year n

ir: Rate of return (interest rate)

ct: Construction time [year]

Finally, the total investment cost is calculated by summing the equipment costs, installation costs, owner costs, interest payments during construction, costs for spare-parts and costs related to the start-up [18] as in Equation 31. The start-up costs are described earlier in this section. The interest payments during the plant lifetime are added in a later step, since it varies from year to year.

$$IC_{tot} = IC_{eq,inst} + IC_{cont} + IC_{oc} + IC_{idc} + IC_{sp} + IC_{su} \tag{31}$$

In some cost estimations, working capital is added as an investment cost. The working capital is the money invested in raw material and supplies in stock; products in stock and semi-finished products being manufactured; cash available for payment of salaries, taxes, operating expenses and purchase of raw materials. Normally, the raw material in inventory equals to the consumption of one month. [31]
It is true that some capital needs to be dedicated to this to be able to have an inventory of fuel and chemicals. However, since this study considers the whole plant lifetime, it is assumed to be recovered at the end of the lifetime, since nearly all the raw material and supplies are consumed. [18]

6.2 Operating expenses

The operating expenses (OPEX) are divided into two categories; fixed operating expenses and variable operating expenses. Fuel costs are in some economic evaluations seen as another category, but is here included in the variable costs. All operating expenses except the fuel costs have been assumed to increase in the same rate as the inflation rate.

6.2.1 Fixed operating expenses

Fixed operating expenses are operating expenses that are close to independent on the operating conditions and production rate. The contributions to the fixed OPEX are listed in Table 6.8. Note that labor and maintenance costs are slightly dependent on the operation of the plant, and are sometimes categorized as variable OPEX. Anyhow, they are relatively constant through the plant lifetime and are thus placed here.

Table 6.8: Fixed operating expenses

Fixed OPEX
Maintenance materials and equipment
Maintenance labour
Permanent labour
Support labour
Administration
Insurance
Property taxes

The fixed OPEX terms are estimated as follows [11, 18]:

- Permanent labor: 50 employees
- Average annual cost per employee: 60000 $[\frac{€}{year}]$
- Total maintenance cost, out of maintenance labour stands for 40%: 1.5% of the total equipment, installation and contingency costs per year
- Administrative and support labour: 30% of permanent labor and 12% of maintenance labor
- Insurance and property taxes: 2% of the total equipment, installation and contingency costs per year

6.2.2 Variable operating expenses

In contrast to the fixed OPEX, the variable OPEX are directly associated with the manufacturing conditions. The terms that build up the variable OPEX are present when a plant operates, and are listed in Table 6.9.

Table 6.9: Variable operating expenses

Variable OPEX
Fuel
Oxygen carrier make-up
Disposal of ash and used oxygen carrier

A list of assumptions has been done when the variable OPEX has been estimated. Important ones are:

- Large amounts of cooling water is needed in a the proposed CLC plant. Therefore, it is assumed that the plant would be built next to a natural water source, as a river, so that the water supply practically is free of charge. It is also assumed that the cooling water stream leaving the plant is free from contamination and of a temperature accepted by environmental regulations so that no penalty fees need to be paid.

- As already mentioned in this chapter, it is assumed that coal fired CLC combustion plants are not obliged to purchase CO_2 certificates for the small amounts of CO_2 that is released to the atmosphere.

- Costs for transport and storage of compressed CO_2 are not included in these calculations. IEAGHG expects these costs to be in the magnitude of 10 $\frac{€}{tonne}$ [18].

Fuel

The coal price fluctuates on a daily basis and is difficult to predict. To do a reasonable assumption of how it will change in the future, a forecast made by The World Bank has been used. Data from [55] provides information about how the coal price is predicted to change during a selection of years between the years 2018-2030. This data is used to interpolate the coal price between the missing years within the interval and is also used to extrapolate the subsequent years, leading to an annual price growth of approximately 0,8%. The current coal price of today matches the price for year 2018 predicted by The World Bank, so it is considered to be a valid source to use. The annual fuel cost is simply calculated by multiplying the coal price with the yearly coal demand.
As mentioned in Section 3.3, it probably is necessary to grind the coal to smaller particles before being used as fuel. The coal grinding is assumed to consume 10 $\frac{kWh_{el}}{tonne\ coal}$ and is withdrawn from the gross electricity output.

Oxygen carrier make-up

The cost for the oxygen carrier make-up stream is estimated by Equation 33. The total solid inventory is approximated to be the sum of the solid inventory in the air reactor and the fuel reactor with an additional 20% added to take into account the oxygen carrier present in the rest of the process streams between the reactors, shown by Equation 32. An oxygen carrier lifetime of 300 hours is assumed, as suggested by [3]. An assumption that has been made is that there is no oxygen carrier loss through the ash stream. Depending on the coal particle size, the OC slipped through the carbon stripper to the ash outflow, can potentially be separated with a filter, leading to practically no losses. The oxygen carrier stream will hereby be continuously filled, and will consist of a mix of fresh and old oxygen carriers.

$$SI_{tot} = 1.2 \cdot (SI_{AR} + SI_{FR}) \tag{32}$$

$$OC_{oxc} = \frac{SI_{tot}}{OC_{lt}} \cdot aot \cdot p_{OC} \tag{33}$$

Where,

SI_{tot}:	Total solid inventory
SI_{AR}:	Solid inventory of the air reactor
SI_{FR}:	Solid inventory of the fuel reactor
OC_{oxc}:	Cost of the oxygen carrier make-up stream
OC_{lt}:	Oxygen carrier lifetime
p_{OC}:	Oxygen carrier price

Operation of the air separation unit

Operational costs of an air separation unit (ASU) may be as low as $\frac{160\ kWh_{el}}{ton\ O_2}$ kWh/ton (or 176 $\frac{176\ kWh_{el}}{tonne\ O_2}$) according to Air Liquide [8]. In literature, the corresponding number is 0,3 $\frac{kWh_{el}}{Nm^3\ O_2}$ which equals to 210 $\frac{kWh_{el}}{tonne\ O_2}$ [33]. This is in line with what is concluded in a feasibility study about carbonate looping [12]. Operating costs and investment costs for an air separation unit have been estimated in that study, since it has an essential role in a standard directly heated carbonate looping configuration [11]. In the report, it is stated that an electrical energy demand of 53 MW_{el} is required to produce 76,3 $\frac{kg}{s}$ O_2. This corresponds to 193 $\frac{kWh_{el}}{tonne\ O_2}$. For this study, an average value for the ASU electricity demand of 200 $\frac{kWh_{el}}{tonne\ O_2}$ has been chosen. This electricity demand is multiplied with the pure oxygen demand for the post oxidation chamber to get a number for the total electricity demand of the ASU unit. That electricity demand is thereafter subtracted from the plant's gross electricity output. Data for the pure oxygen demand is take from the Aspen model.

Other variable operating expenses

Compressors and pumps
Compressors and pumps have a non-negligible electricity demand. This electricity demand is subtracted from the plant's gross electricity output.

Cyclones
Operational costs of cyclones are low and only related to added compressor demand due to the pressure drop of the gas through the cyclones [45]. Operational expenses apart from maintenance costs of the cyclones are therefore neglected.

Ash disposal
The net cost of ash disposal is expected to be negligible according to IEAGHG [18].

6.3 Levelized cost of electricity

Levelized cost of energy (LCOE) is a useful tool to compare the average lifetime costs of electricity production between different plants and energy sources. It is defined in Equation 34, and is the ratio between the total lifetime costs and the total amount of electricity produced over the plant lifetime. It does not take into account market or technology risks and is therefore more suitable for mature, regulated and monopoly markets than for markets with high uncertainties. Despite this weakness, it is widely used for comparing the costs of power generation technologies [59] . As the electricity market often has the characteristics just mentioned, it gives relatively accurate predictions.
The calculation of LCOE in this study is carried out in a similar way as by The International Energy Agency (IEA) in [59] and by Junk, M. in [11].

$$LCOE = \frac{\sum(plant\ lifetime\ costs)}{\sum(plant\ lifetime\ electricity\ production)} \tag{34}$$

All cost contributions are calculated on an annual basis and are then levelized by means of an annuity factor that is based on the plant lifetime and the assigned rate of return. When the LCOE is determined, it can be used to estimate the CO_2-avoidance cost, which is the cost to avoid releasing CO_2 to the atmosphere as in a reference plant. A first step to calculate the LCOE, is to determine the total investment cost, according to Equation 31.
A common way to pay back the investment costs of an investment is to depreciate it during the whole lifetime due to accounting reasons. Also here, for the base case, the plant lifetime of 25 years is assumed to be the depreciation time. A linear depreciation model is used, giving an annual depreciation according to Equation 35. In the sensitivity analysis, the depreciation time is varied to see what effect it has on the total cost.

$$AD = \frac{IC_{tot}}{dp} \tag{35}$$

Where,

AD:	Annual depreciation $[\frac{€}{year}]$
dp:	Depreciation period $[years]$

The annual interest payment decreases as the invested money is depreciated and is calculated by Equation 36. Since the depreciation time is chosen to be the plant lifetime, the whole investment is "paid back" by the end of the assumed lifetime.

$$I_n = (IC_{tot} - (n-1) \cdot AD) \cdot ir \tag{36}$$

Where,

I_n:	Interest payment at year n [€]
IC_{tot}:	Total investment cost [€]
n:	Year of operation, where $n = 1$ is the first year of construction
ir:	Rate of return (ir) [%]

Thereafter, the annual capital expenses (CAPEX) for year n can be evaluated by summing the depreciation and interest payment for the corresponding year, according to Equation 37.

$$CAPEX_n = AD + I_n \tag{37}$$

Where,

$CAPEX_n$:	Capital expenses at year n [€]

The CAPEX and the operating expenses (OPEX) together constitute the annual costs. The total OPEX is the sum of the variable operating expenses, in which fuel costs are included and fixed operating expenses, as shown in Equation 38. In Section 6.2.2, a description of what is included in the two OPEX terms is available.

$$OPEX_n = OPEX_{n,fixed} + OPEX_{n,var} \tag{38}$$

Where,

$OPEX_n$:	Operating expenses at year n [€]
$OPEX_{n,fixed}$:	Fixed operating expenses at year n [€]
$OPEX_{n,var}$:	Variable operating expenses at year n [€]

Since the goal is to levelize all the costs through the whole plant lifetime, all costs must be transformed to the net present value in order to be comparable. The net present value of the total costs related to the plant is evaluated by Equation 39.

$$NPV = \sum_{n=1}^{ct+lt} \frac{CAPEX_n + OPEX_n}{(1+ir)^{n-1}} \tag{39}$$

Where,

NPV:	Net present value [€]
ct:	Construction time $[years]$
lt:	Plant lifetime $[years]$

The total cost in net present value is then multiplied with an annuity factor in Equation 41. The annuity factor, r, gives information about how large share of the capital cost that should be paid in interest and depreciation each year over the specified period [38].

$$r = \frac{ir \cdot (1+ir)^{(ct+lt)}}{(1+ir)^{(ct+lt)} - 1} \tag{40}$$

$$LC = NPV \cdot r \tag{41}$$

Where,

LC:	Levelized cost $\left[\frac{€}{year}\right]$
r:	Annuity factor $\left[year^{-1}\right]$

An additional cost that has to be added is the asset retirement cost, i.e costs related to the deconstruction of the plant. It has a separate interest rate, IR_{ar}, shown in Table 6.2. The specific price, sAR, has been chosen as 75 $\frac{DM}{kW_{el}}$ according to [20]. Equation 42 shows how the levelized asset retirement cost is estimated [11] by multiplication with an annuity factor so that it can be added up with the other levelized costs..

$$LAR = \frac{sAR \cdot P_{el,gross}}{(1+ir_{ar})^{ct+lt}} \cdot \underbrace{\frac{ir_{ar} \cdot (1+ir_{ar})^{ct+lt}}{(1+ir_{ar})^{ct+lt} - 1}}_{r_{ar}} \cdot (lt + ct) \tag{42}$$

Where,

LAR:	Levelized cost of asset retirement $\left[\frac{€}{year}\right]$
sAR:	Specific asset retirement price $\left[\frac{€}{kW_{el}}\right]$
$P_{el,gross}$:	Gross electricity output of the plant $[kW_{el}]$
r_{ar}:	Annuity factor for asset retirement $[year^{-1}]$
ir_{ar}:	Rate of return for the money assigned for asset retirement (ir) [%]

Finally, the LCOE can be calculated by Equation 43. Junk, M. [11] has another approach and divides the $CAPEX_n$ and $OPEX_n$ terms in Equation 39 by the electricity produced at year n, since he has included an annual decrease of the electricity production. As described in the beginning of Chapter 6, an annual decrease is not included in this work and the electricity produced is assumed to be constant over the plant lifetime. This allows the division of the annual electricity production to be done in a later step, Equation 43, for easier calculations.

$$LCOE = \frac{LC + LAR}{AEP} \tag{43}$$

Where,

$LCOE$: Levelized cost of electricity $\left[\frac{€}{MWh_{el}}\right]$

AEP: Annual electricity production $\left[\frac{MWh_{el}}{year}\right]$

6.4 CO2 avoidance cost

The CO_2 avoidance cost can be evaluated by comparing the difference in LCOE between the proposed CLC plant and a conventional coal fired reference plant, and relating it to the CO_2 that would not be captured and instead be released to the atmosphere. The difference in LCOE is calculated by Equation 44. $LCOE_{ref}$, the LCOE for the reference plant, is calculated in a similar way as for LCOE by Equation 43.

$$\Delta LCOE = LCOE - LCOE_{ref} \tag{44}$$

The difference between the amount of CO_2 that is captured by the CLC plant and what would be released by the reference plant, is approximated by Equation 45. It is important to include the plant electrical efficiency factors, since a modern reference coal fired power plant produces more electricity than a CLC-plant with the same fuel input, due to lower internal electricity consumption. The electrical efficiency of the reference plant has been assumed to be 46%. The value for the CO_2 emissions from the fuel that is burnt, $\dot{m}_{CO_2}$, is based on results from the Aspen model.

$$\Delta \dot{m}_{CO_2} = \underbrace{\dot{m}_{CO_2} \cdot \frac{\eta_{el}}{\eta_{el,ref}}}_{CO_2 \text{ emissions from reference plant}} - \underbrace{\dot{m}_{CO_2} \cdot (1 - \eta_{CO_2})}_{CO_2 \text{ emissions from CLC plant}} \tag{45}$$

Where,

$\Delta \dot{m}_{CO_2}$: Difference in CO_2 emissions $\left[\frac{tonnes\ CO_2}{MWh_{th}}\right]$

$\dot{m}_{CO_2}$: CO_2 emissions from the fuel burnt in the CLC plant $\left[\frac{tonnes\ CO_2}{MWh_{th}}\right]$

η_{el}: Electrical efficiency of the CLC plant$[-]$

$\eta_{el,ref}$: Electrical efficiency of the reference plant$[-]$

η_{CO_2}: CO_2 capture efficiency $[-]$

The electrical efficiency is defined by Equation 46 and the CO_2 capture efficiency by Equation 47.

$$\eta_{el} = \frac{P_{el,net}}{P_{fuel}} \tag{46}$$

$$\eta_{CO_2} = \frac{amount\ CO_2\ captured}{amount\ CO_2\ released\ from\ combustion} \tag{47}$$

Where,

$P_{el,net}$: Net electricity production $[MW_{el}]$

P_{fuel}: Thermal energy from consumed fuel $[MW_{th}]$

Lastly, the CO_2 avoidance cost can be calculated by Equation 48, and is given in the unit $\left(\frac{€}{tonne\ CO_2}\right)$.

$$CO_2AC = \frac{\Delta LCOE}{\Delta \dot{m}_{CO_2}} \tag{48}$$

6.5 Payback-period through cumulative cash flow approach

The payback-period approach is an alternative method to evaluate an investment, and is more straightforward than calculating the LCOE. It simply is the time required to recover the cost of an investment and does not take into account the time value of money as in net present value calculations [31].

6.5.1 Incomes

The revenue of an electricity production plant is simply made up by the sales of electricity, calculated by Equation 49. The electricity price is assumed to vary according to the assumptions on page 25 in Chapter 6.
An additional potential income is the rest value of equipment and utilities that can be sold at the end of the plant lifetime. However, due to estimation difficulties and uncertainty, it has been neglected in this study.

$$rev_n = AEP \cdot p_{el} \tag{49}$$

Where,

rev_n: Revenue at year n [€]

AEP: Annual electricity production $[MWh_{el}]$

p_{el}: Price of electricity sold $[\frac{€}{MWh_{el}}]$

6.5.2 Expenses

In this payback-period approach, all the capital expenses are paid during construction and the interest rate is ignored. In other words, no depreciation is taken into account and all capital costs are paid at the time they appear. The operating expenses are equal to what is estimated in the LCOE calculations in Section 6.3.

6.5.3 Cumulative cash flow

In order to determine the payback-period, the net cash flow of each year needs to be predicted. This is done by Equation 50. The cash flow will naturally be negative during all years of construction before becoming positive at the start of operation for a profitable plant. The magnitude of taxation on profits varies from country to country, but the current German taxation rate has been selected according to Table 6.2.
A term that commonly is used is the break-even point. It is defined as the year when the cumulative cash flow reaches above 0 [43].

$$ccf = \sum_{n=1}^{n} (rev_n - expenses_n) \cdot \underbrace{(1 - tax)}_{\text{if positive cash flow}} \tag{50}$$

Where,

ccf:	Cumulative cash flow [€]
$expenses_n$:	Operating expenses at year n [€]
tax:	Corporate tax rate for profits [€]

A comparison of the equipment and installation costs for all process units is done in Figure 6.2. The total installed equipment cost equals to €580 million. Table 6.10 provides information about the costs for each group of process units.

In Figure 6.3 and Table 6.11, the total cost of the CO_2-compression is compared with the rest of the process units. The CO_2-compression configuration has not been optimized, but will anyhow constitute a significant part of the total EPC cost.

In Figure 6.4 and Table 6.12, the additional costs for contingencies, owner costs, interest during construction and costs related to the start-up phase have been added. The sum of these terms is the total investment cost. The costs related to equipment installation, contingency, owner costs, start-up and spare parts together sum up to approximately 70% of the equipment cost.

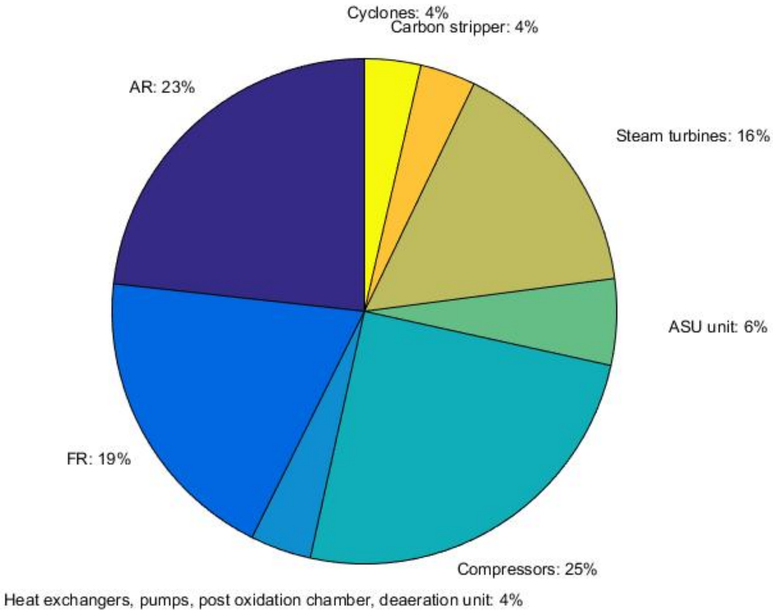


Figure 6.2: Breakdown of the engineering, procurement and construction (EPC) cost of equipment

Table 6.10: Breakdown of the engineering, procurement and construction (EPC) cost of equipment

Group of process units	[€ *million*]
Compressors	144
Air reactor ‡	134
Fuel reactor *	112
Steam turbines	91
Air separation unit	32
Cyclones	23
Carbon stripper	21
Pumps	10
Post oxidation chamber	5
Heat exchangers §	7
Feedwater tank	0.1
	Sum: 578 $\left(0.578\ \frac{€ million}{MW_{th}}\right)$

‡ Fluidized bed cooler, superheater, economizer and air pre-heater included
* Superheaters and economizers included
§ Includes heat exchangers for CO_2-compression and heat exchanger for feedwater pre-heating by process steam

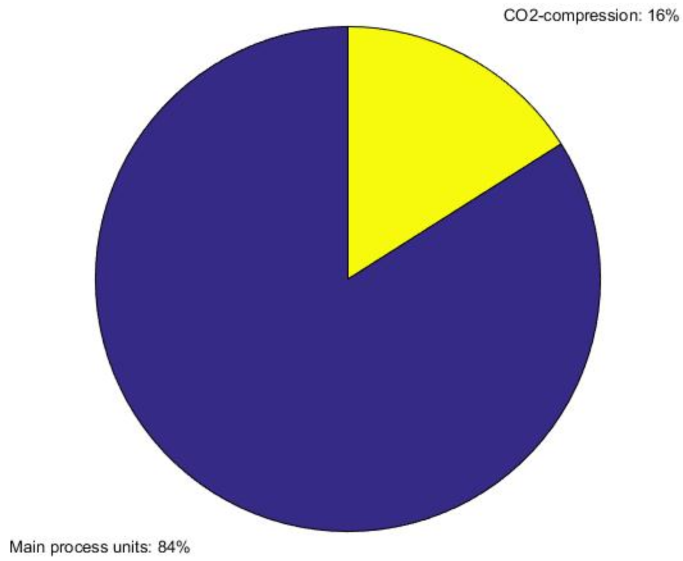


Figure 6.3: CO_2-compression EPC cost in comparison with the rest of the process equipment

Table 6.11: CO_2-compression EPC cost in comparison with the rest of the process equipment

Group of process units	[€ ***million***]
Main process units	486
CO_2-compression ‡	93
	Sum: 578

‡ Includes heat exchangers, compressors, pumps and motors in the CO_2-compression section

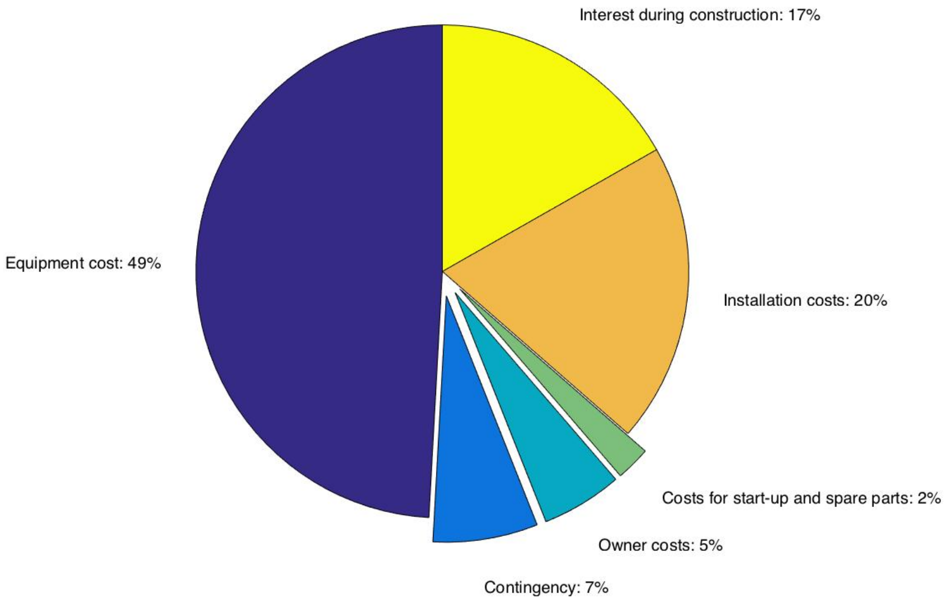


Figure 6.4: Breakdown of the total investment cost

Table 6.12: Breakdown of the total investment cost

Group of process units	[€ $\boldsymbol{million}$]
Equipment	413
Installation	165
Interest during construction	141
Contingency	58
Owner costs	45
Start-up	16
Spare parts	3
	Sum: 840
	$\left(0.84 \frac{€\ million}{MW_{th}}\right)$

A breakdown of the annual operating costs of the first operating year is shown in Figure 6.5. The fuel costs and the oxygen carrier costs together represent the variable operating costs while the remaining parts are the fixed operating costs. Internal electricity demand is not counted as an operating cost. The reason to why Figure 6.5 only is representative for the first year is because the cost fractions do not evolve uniformly with time. Table 6.13 shows the operating costs of year one. A graphical illustration of how the total operating cost varies with time is available in Figure 6.7 and 6.8.

In order to make a comparison between the group of costs seen over the whole plant lifetime, Figure 6.6 and Table 6.14 have been created. Important to note is that to enable comparable numbers, all the values are transformed into NPV.

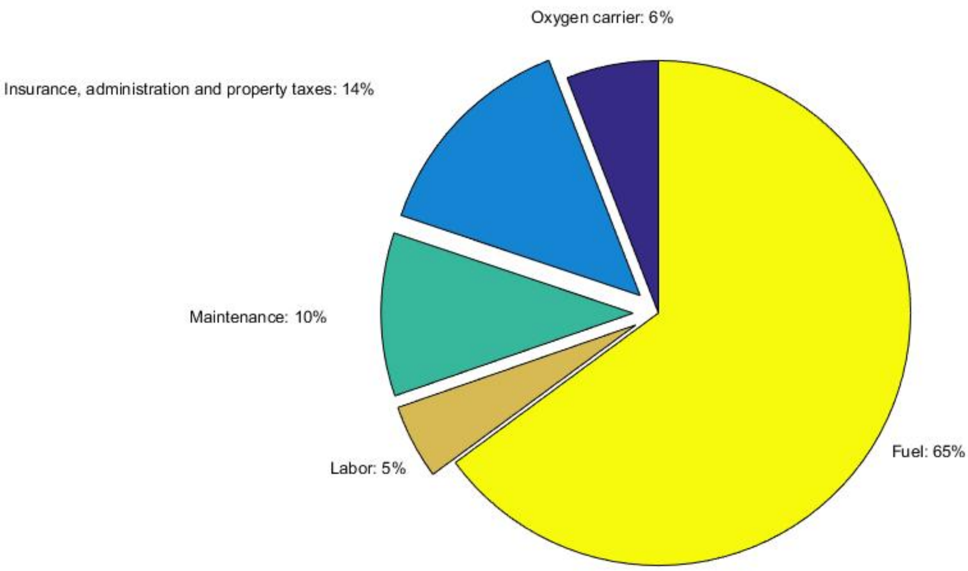


Figure 6.5: Breakdown of the plant operating costs for the first year

Table 6.13: Breakdown of the plant operating costs for the first year

Operating cost	[€ *million*]
Fuel	59
Insurance, administration and taxes	13
Maintenance ‡	10
Oxygen carrier	5
Labor	4
	Sum: 91
	$\left(0.091\ \frac{million\ €}{MW_{th}}\right)$

‡ 40% material cost and 60% maintenance labor cost according to [18]

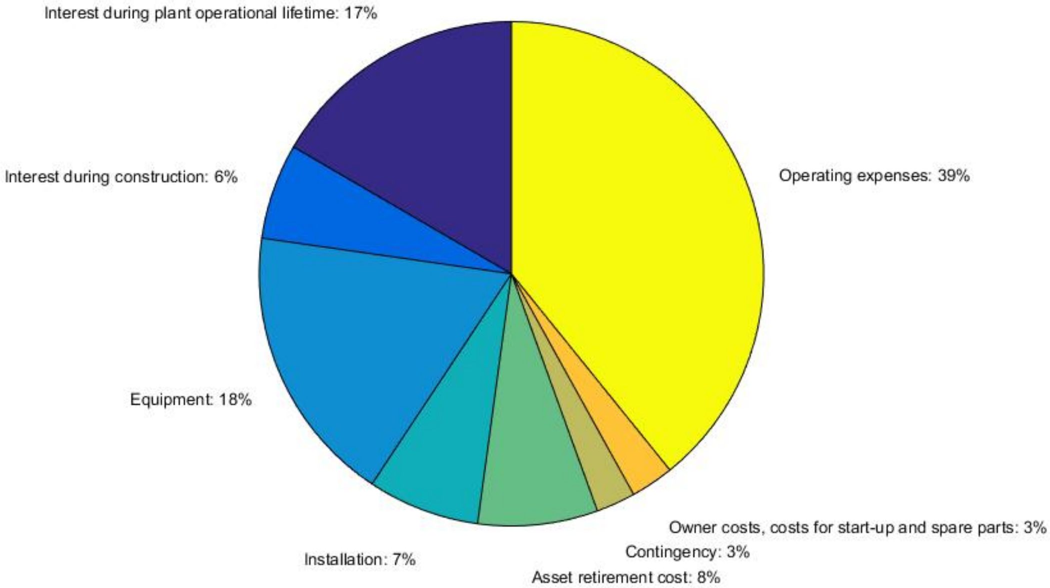


Figure 6.6: Breakdown of the plant lifetime cost (NPV)

Table 6.14: Breakdown of the plant lifetime cost (NPV)

Cost contribution	[€ ***million***]
Operating cost	902
Equipment cost	413
Interest during plant operation	382
Asset retirement cost	177
Installation cost	165
Interest during construction	141
Owner costs, spare parts and start-up costs	64
Contingency costs	58
	Sum: 2301
	$\left(2.3\ \frac{million\ €}{MW_{th}}\right)$

There is a difference between when costs related to a power plant appears and when the debts will be paid back. Construction costs are present during the years of construction and operating costs during the years of operation, but the costs will only be paid back during the operating years, when there is an income. To display the difference, Figure 6.7 illustrates when the costs appear and how they vary from year to year, while Figure 6.8 illustrates a payment plan. The costs related to CAPEX after the years of construction are due to interest payments. The annual interest payment becomes lower and lower in the same speed as the invested money linearly is paid off, until the whole investment is paid back at the end of the depreciation period. The operating costs on the other hand, increases due to inflation and an increasing fuel cost. The values in Figure 6.7 and 6.8 are actual values and not NPV.

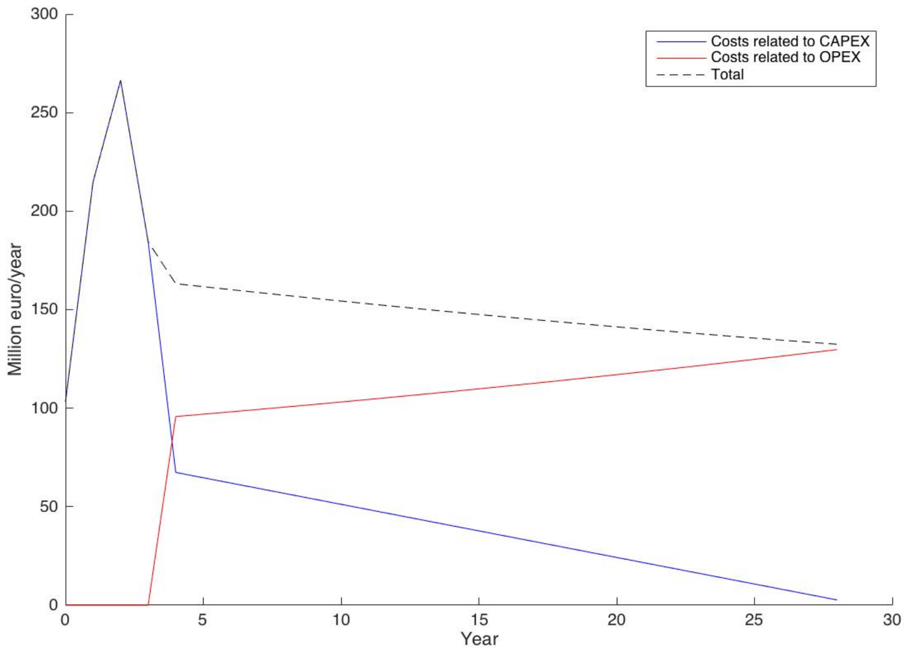


Figure 6.7: Costs during construction and plant lifetime

The cumulative cash flow over the whole plant lifetime is shown in Figure 6.9. In this graph, year 0 represents the start of plant operation, in order to make an easy visual representation of the break-even point, which is the payback period. As can be seen, it is ten years for the base case. The income during the last year is uncertain due to the asset retirement cost, which could be evened out by the spare value of some of the equipment that could be sold.

In Table 6.15, the most important economic results are listed, including the key performance indicators that were sought and part of the project objectives, mentioned in Chapter 2.

Table 6.15: Key results from the economic evaluation of the base case

Economic key results		
Investment cost	840	€ *million*
Total installed equipment cost	578	€ *million*
Initial annual operating cost	91	€ *million*
Levelized cost of electricity	48	$\frac{€}{MWh}$
CO_2-avoidance cost	23	$\frac{€}{tonne\ CO_2}$
Payback period	10	*years*

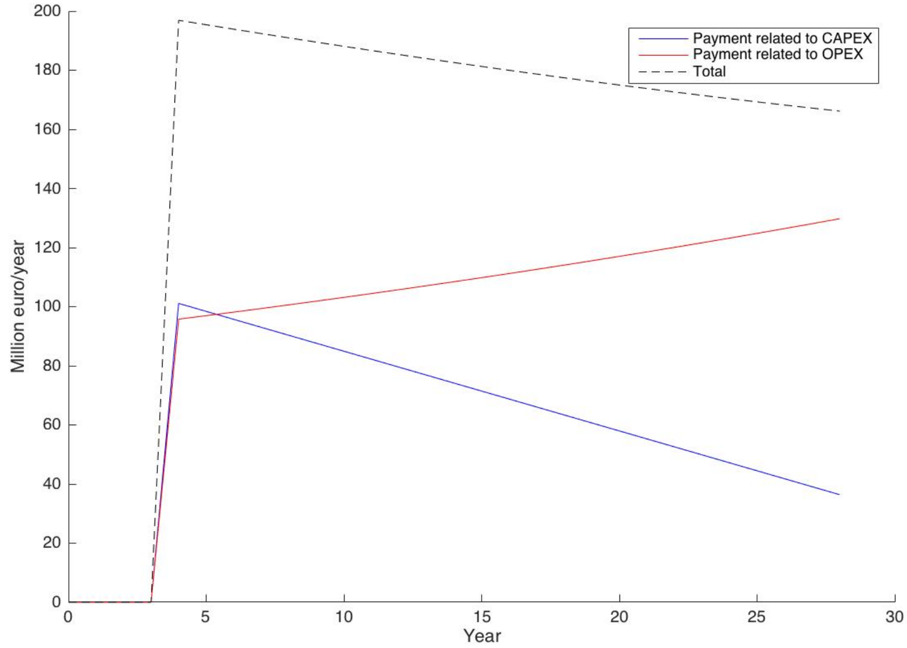


Figure 6.8: Payment plan

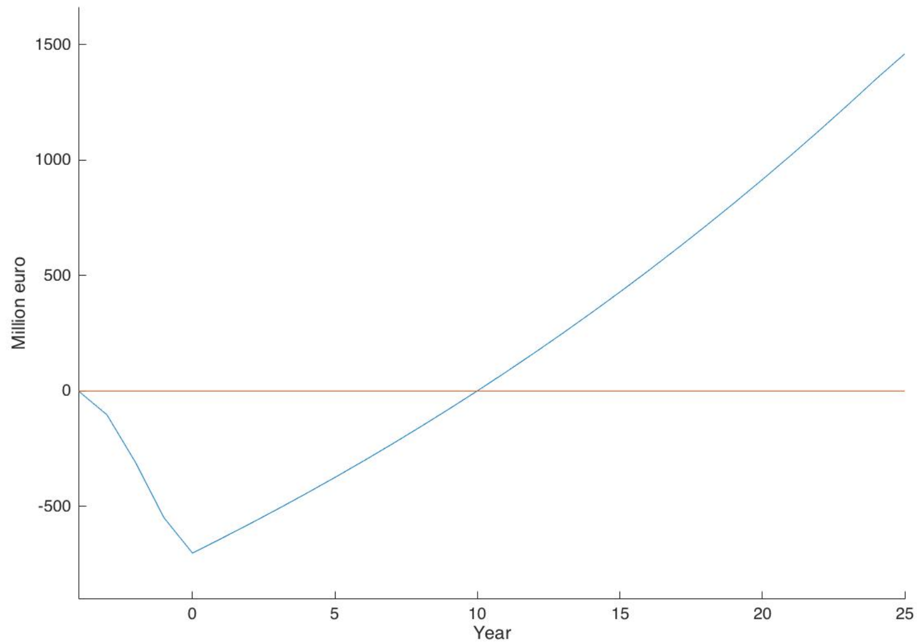


Figure 6.9: Cumulative cash flow

7 Sensitivity analysis

A sensitivity analysis has been performed to gain knowledge about which changes in which parameters the economic output is most sensitive against. A number of parameters have been varied both upwards and downwards from the base case, that was described in Chapter 6. The parameters have not been varied uniformly positively and negatively. Instead, the ranges have rather been chosen as realistic intervals in where the parameters could lie within, and can are listed in Table 7.1. Due to the time constraint, only a one-dimensional sensitivity analysis has been performed, which means that only one parameter at a time has been varied. A drawback from this method is that interconnected effects cannot be analyzed.

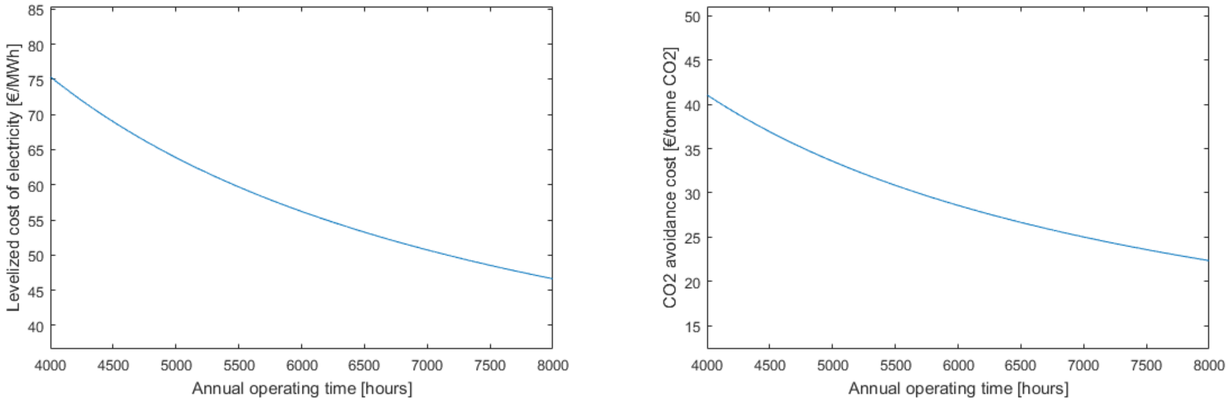
Table 7.1: Parameters that are varied in the sensitivity analysis, and their respective intervals

Parameter	**Lower boundary**	**Base case**	**Upper boundary**	**Unit**
Plant lifetime	15	25	40	$years$
Depreciation time	15	25	40	$years$
Annual operating hours	4000	7500	8000	$hours$
Rate of return	4	8	12	%
Inflation rate	0	2	5	%
Normalized electricity price to grid supplier ‡	0.25	1	2	–
Corporate tax rate on profits	10	32	50	%
CO_2-capture efficiency	80	98	100	%
Normalized oxygen carrier performance ratio †	0.05	1	20	$\frac{OC\ price}{OC\ average\ lifetime}$
Normalized equipment and installation cost ‖	0.5	1	2	–
Investment cost factor *	1.5	1.69	4	–
Coal price §	0.5	1.0	1.5	–

‡ At the beginning of the plant lifetime. Normalized value 1 = 0.0563 $\frac{€}{kWh}$
§ At the beginning of the plant lifetime. Normalized value 1 = 60 $\frac{\$}{tonne}$
† Normalized value 1 equals to an oxygen carrier performance ratio of 175 $\frac{\$}{tonne}$ and lifetime of 300 hours
‖ Normalized value 1 = 578 € million
* Factor multiplied with the equipment cost to estimate the total investment cost. The base case number corresponds to the results indicated by the base case calculations. E.g in the base case, the investment cost excluding interest expenses equals to 169% of the total equipment cost

7.1 Sensitivity analysis results

How a variation of the annual operating hours affects the LCOE and CO_2-avoidance cost is shown in Figure 7.1. As expected, the LCOE increases with decreased annual operating hours for the CLC plant. The CO_2-avoidance cost shows the same trend because even though a decrease of the annual operating hours affects the reference conventional coal fired power plant negatively, the LCOE for the CLC-plant has a higher dependency of the operating hours, and therefore $\Delta LCOE$ changes in Equation 48, resulting in a change in CO_2-avoidance cost.

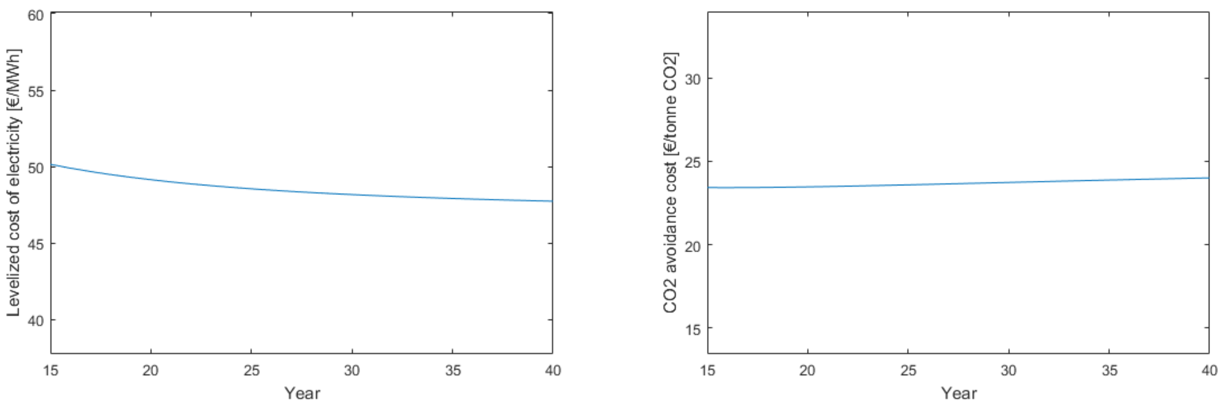


(a) LCOE (b) CO_2-avoidance cost

Figure 7.1: Annual operating hours varied

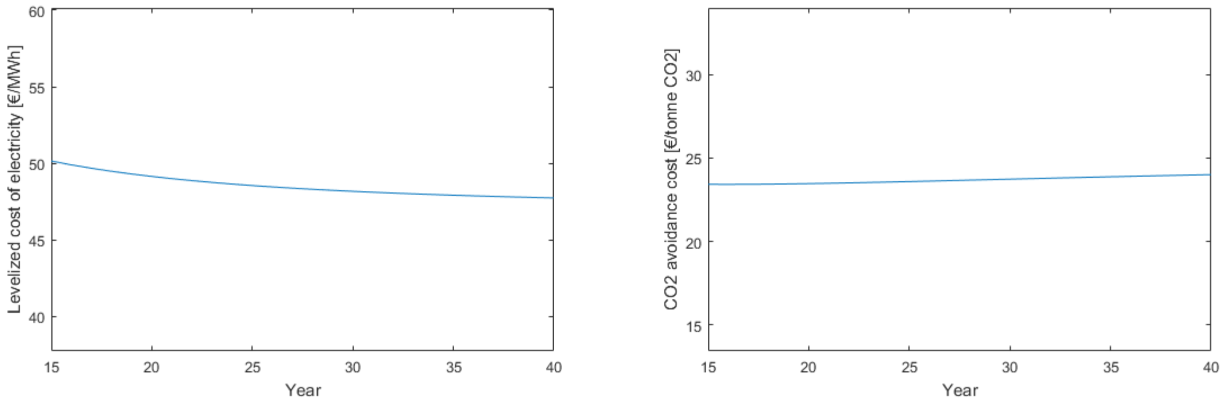
The effect that the plant lifetime has on the LCOE and CO_2-avoidance cost, is displayed in Figure 7.2. The LCOE decreases with plant lifetime while the CO_2-avoidance cost is nearly constant but with a slight increase. This is because the LCOE for the reference plant decreases marginally more than the LCOE for the CLC plant. The difference is due to the diverse variable operating cost functions for the reference plant LCOE compared to the corresponding one for the CLC plant.

Moreover, in Figure 7.2, the depreciation time is set to the whole plant lifetime. This is a somewhat questionable assumption because of the uncertainty when predicting a plant lifetime. The question can be asked; what happens if the plant lifetime is predicted to 35 years but then must be shut down after 20 years due to political reasons? There would be a remaining undepreciated amount of the investment that needs to be paid off earlier than planned. Therefore, another test was made where the depreciation time was set to maximum 25 years, e.g. if the plant would remain active for 30 years, it is fully depreciated already after 25 years and no depreciation occurs during the last five years. The result is shown in Figure 7.3. It can be concluded that since all depreciation and interest payments are converted to net present values, the graphs in Figure 7.2 are equivalent to the ones in 7.3. Seen in net present values, the total depreciated capital expense is independent on the depreciation time.



(a) LCOE (b) CO_2-avoidance cost

Figure 7.2: Plant lifetime varied, depreciation during whole plant lifetime

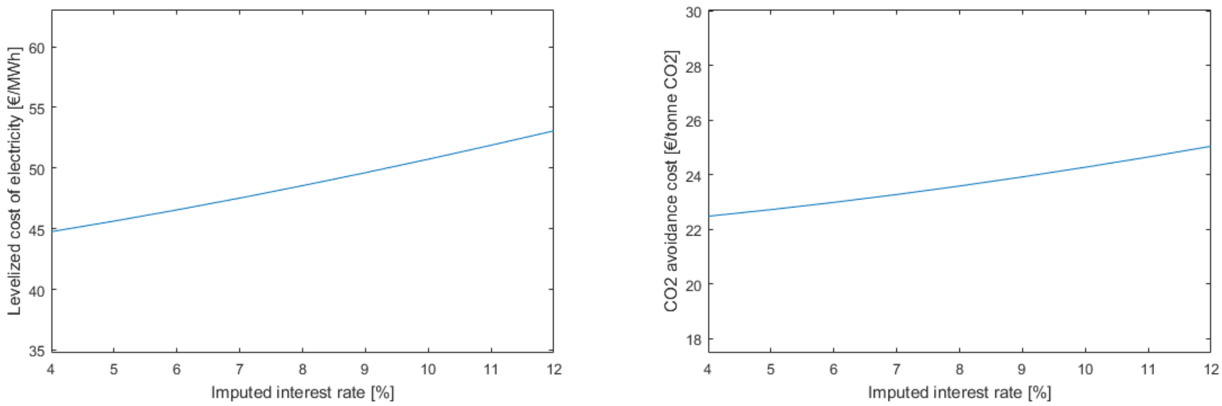


(a) LCOE

(b) CO_2-avoidance cost

Figure 7.3: Plant lifetime varied, depreciation during 25 years if lifetime is greater than 25 years

The discount rate, or assigned rate of return, is varied in Figure 7.4. As can be see in Equation 41, a change in the assigned rate of return affects both the NPV and the annuity factor. If the assigned rate of return is increased, the annuity factor increases by percentage more than the NPV decreases, resulting in an increased LCOE. The CO_2-avoidance cost likewise increases, because the higher LCOE for the CLC plant increases more compared to the LCOE for the reference plant.

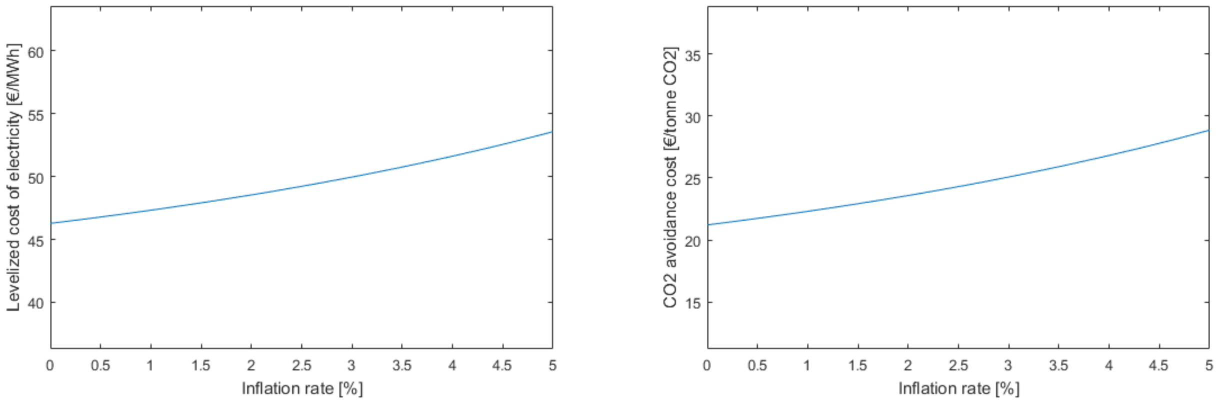


(a) LCOE

(b) CO_2-avoidance cost

Figure 7.4: Assigned rate of return (imputed interest rate)

Figure 7.5 shows that both the LCOE and the CO_2-avoidance cost are positively correlated to the inflation rate. Naturally, since big part of the operating cost, including the oxygen carrier cost is modelled to be directly dependent on the inflation rate, the LCOE increases with an increased inflation rate. Why the CO_2-avoidance cost also shows an increasing trend is because the operating costs for the CLC plant is higher than for the reference coal fired power plant, and therefore increases more in magnitude, leading to a higher CO_2-avoidance cost.

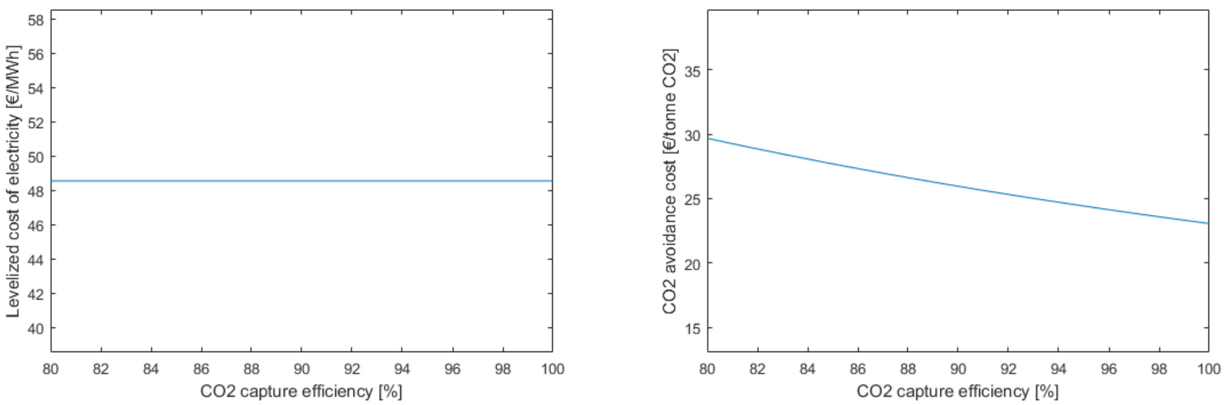


(a) LCOE

(b) CO_2-avoidance cost

Figure 7.5: Inflation rate varied

According to the definition of CO_2 capture efficiency in Equation 47, it affects how much CO_2 that is released to the atmosphere. The operating conditions would slightly vary due to some changed mass flows, leading to a changed load of the corresponding compressors. However, when the CO_2 capture efficiency only is varied within the small stated range, these changes would have a negligible effect on the total output. Therefore, in this sensibility analysis, it only changes the CO_2 released to the atmosphere. For a conventional coal fired power plant, it plays a significant role because the number of CO_2-emission certificates that must be purchased is altered. That is the case for the reference plant. For the proposed CLC plant, the LCOE is not changed since it is assumed that a plant with this small amount of CO_2 is excluded from the law to purchase CO_2 emission rights. This is the explanation to the constant curve in Figure 7.6a. The CO_2-avoidance cost on the other hand, in Figure 7.6b, has a negative correlation with the CO_2-capture efficiency. This because $\Delta LCOE$ in Equation 48 decreases due to the increased amount of CO_2-certificates that must be purchased by the reference plant, at the same time as $\Delta \dot{m}_{CO_2}$ increases because more CO_2 is captured.

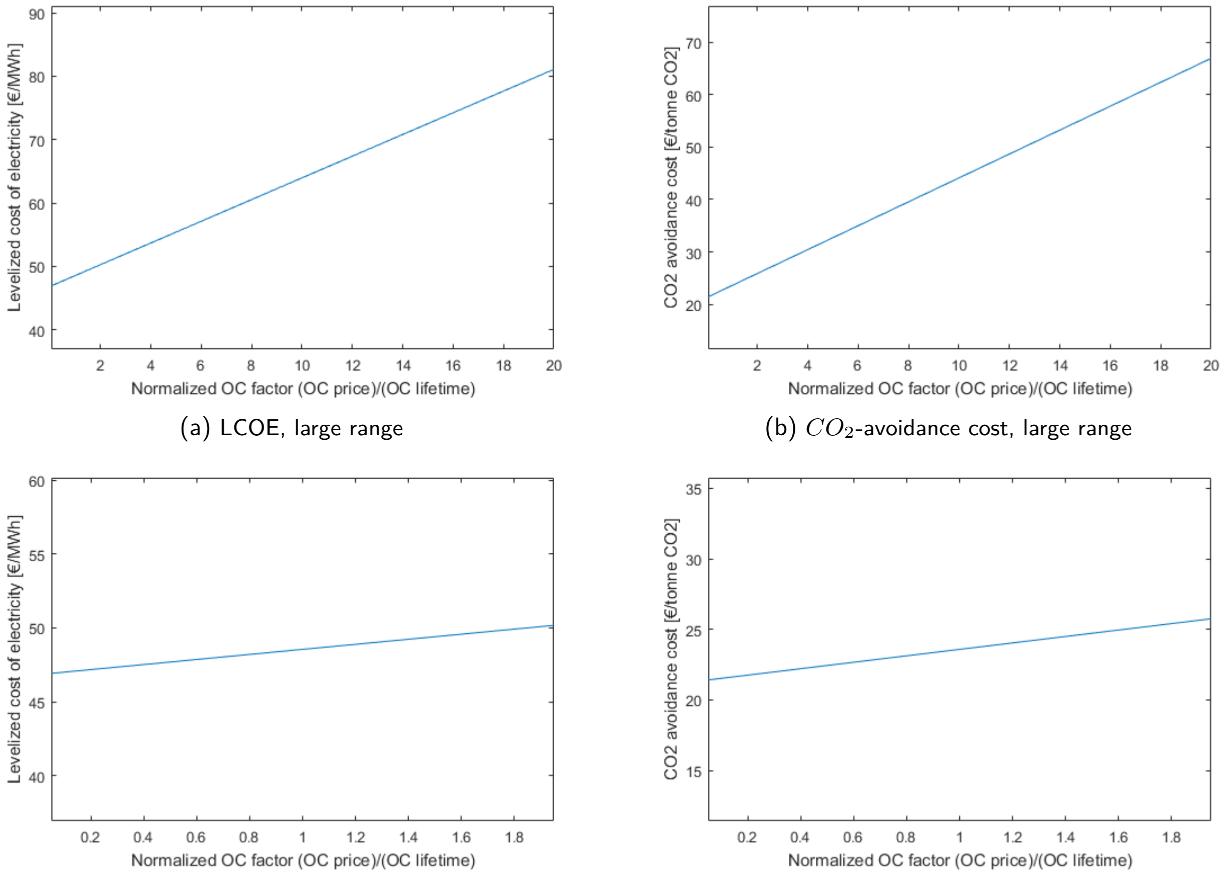


(a) LCOE

(b) CO_2-avoidance cost

Figure 7.6: CO_2-capture efficiency varied

As seen in Table 7.1, the normalized oxygen carrier factor (OC factor) has been varied in a large range. This because the performance and price of OC still is relatively uncertain. Because of this, two graphs each for the LCOE and the CO_2-avoidance cost with different axis is provided to offer a sufficient visual understanding of the trends. A high OC factor means a high price and/or a low lifetime, which consequently increases the LCOE, seen in Figure 7.7a and 7.7c. Since a high OC factor only affects the CLC plant and not the reference coal fired power plant, a high LCOE thereupon gives a high CO_2-avoidance cost as can be concluded from Figure 7.7b and 7.7d. All graphs in Figure 7.7 are linear.



(a) LCOE, large range

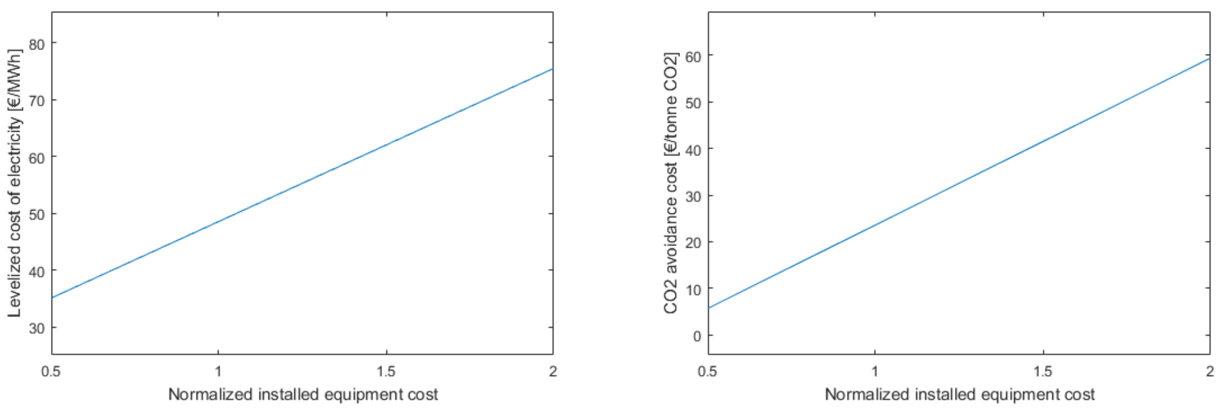
(b) CO_2-avoidance cost, large range

(c) LCOE, small range

(d) CO_2-avoidance cost, small range

Figure 7.7: Normalized oxygen carrier ratio $\left(\frac{OC\ price}{OC\ lifetime}\right)$ varied

The installed equipment cost has a big contribution to the total cost, seen in Figure 6.6, and therefore has a great impact on the LCOE. This is confirmed by Figure 7.8. The LCOE as well as the CO_2-avoidance cost increases linearly with an increased installed equipment cost factor. Both the reference plant and the CLC plant are dependent on the installed equipment cost in the same way.

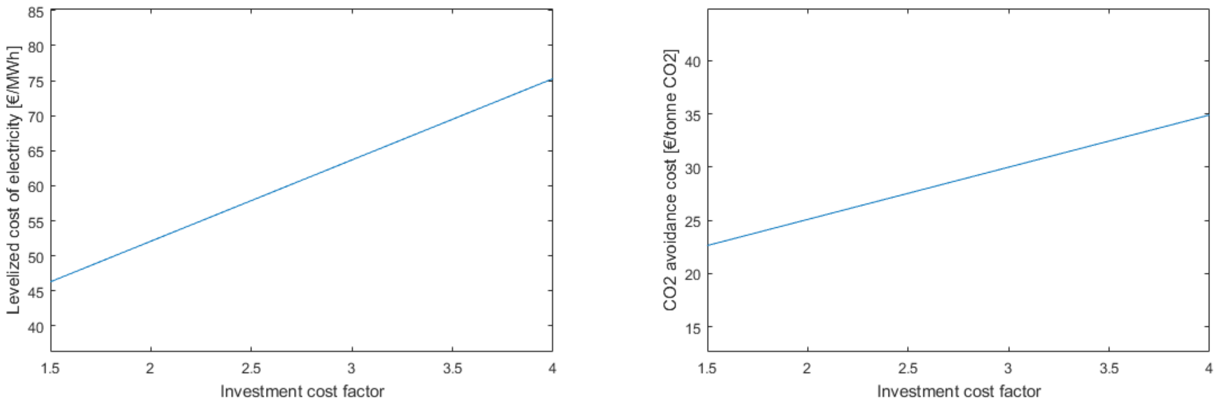


(a) LCOE

(b) CO_2-avoidance cost

Figure 7.8: Normalized installed equipment cost varied

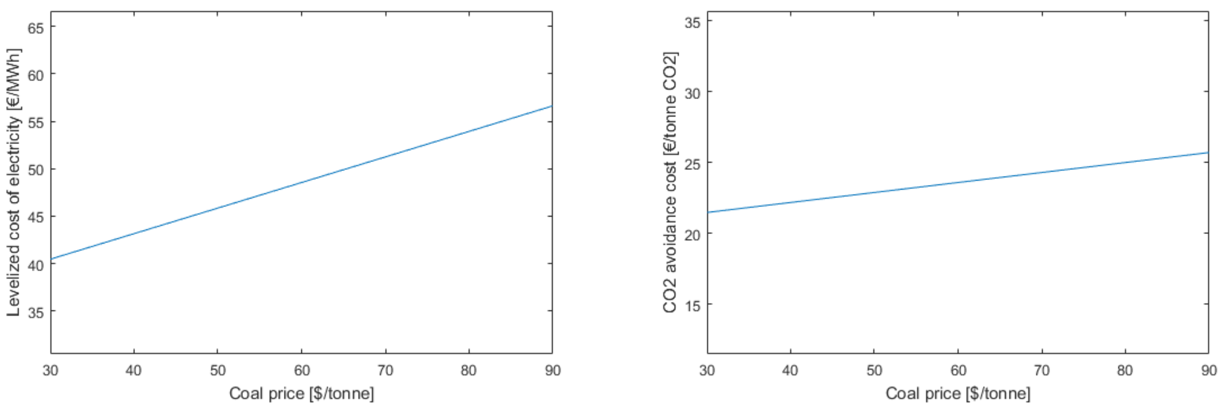
A similar trend as for when the installed equipment cost is varied, appears when the investment cost factor is varied in Figure 7.9. When the investment cost factor, which is the factor multiplied with the equipment cost to predict the total investment cost, is varied, a linear trend for both the LCOE and the CO_2-avoidance cost is visible.



(a) LCOE (b) CO_2-avoidance cost

Figure 7.9: Investment cost factor varied

The initial coal price is varied in Figure 7.10. Note that the discussion in this section is about the coal price at the beginning of plant construction, since the coal price variation during the plant lifetime is based on the price at the first year. The fuel cost has a major contribution to the variable operating cost, and consequently raises the LCOE significantly. A high coal price also affects the reference coal fired power plant negatively, but since a CLC plant consumes a bigger amount of fuel due to a higher internal electricity demand, the latter is slightly more sensitive towards an increase of coal price. That is why a lightly ascending trend is seen for the CO_2-avoidance cost curve. Both graphs are linear.



(a) LCOE (b) CO_2-avoidance cost

Figure 7.10: Coal price at start of construction varied

The relation between the payback-period and the initial selling electricity price at the start of construction is presented by Figure 7.11. A discrete curve is obtained because the payback-period is considered as full years. For the base case, a payback-period of ten years is expected according to the results presented in Table 6.15. The higher the selling electricity price to the grid supplier is, the higher the annual profit is, and a shorter pay-back period is expected. If the initial selling price of electricity is as low as 65% of the one in the base case, corresponding to 0.0366 $\frac{€}{kWh}$, the investment of the plant is not expected to pay-off itself during a lifetime of 25 years.

The corporate tax rate for profits also has an impact on the payback-period, even though it is not as significant as the selling electricity price. Figure 7.12 shows the impact of the corporate tax rate on the payback-period. Corporate tax rates above 32% are expected to give payback-periods longer than ten years.

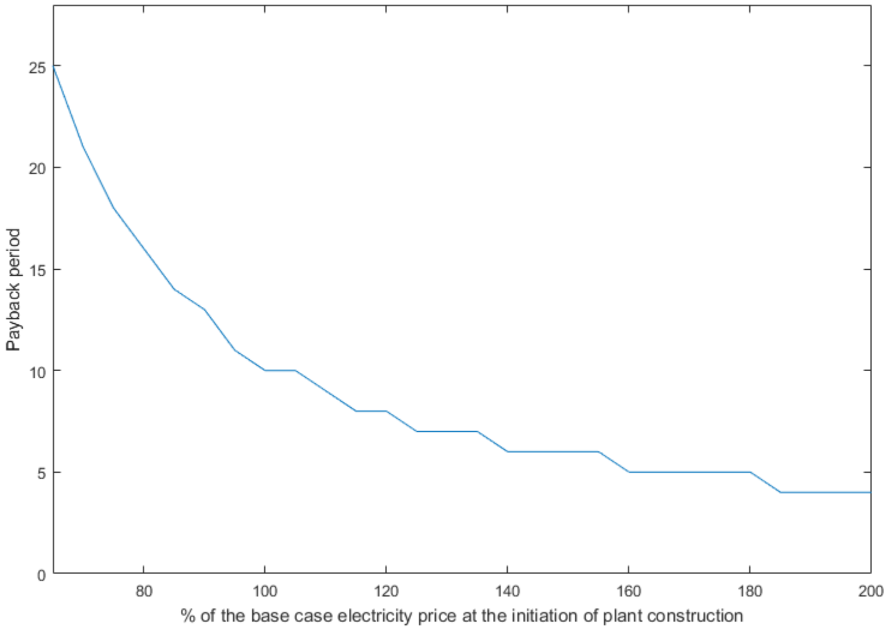


Figure 7.11: Payback-period, selling electricity price varied

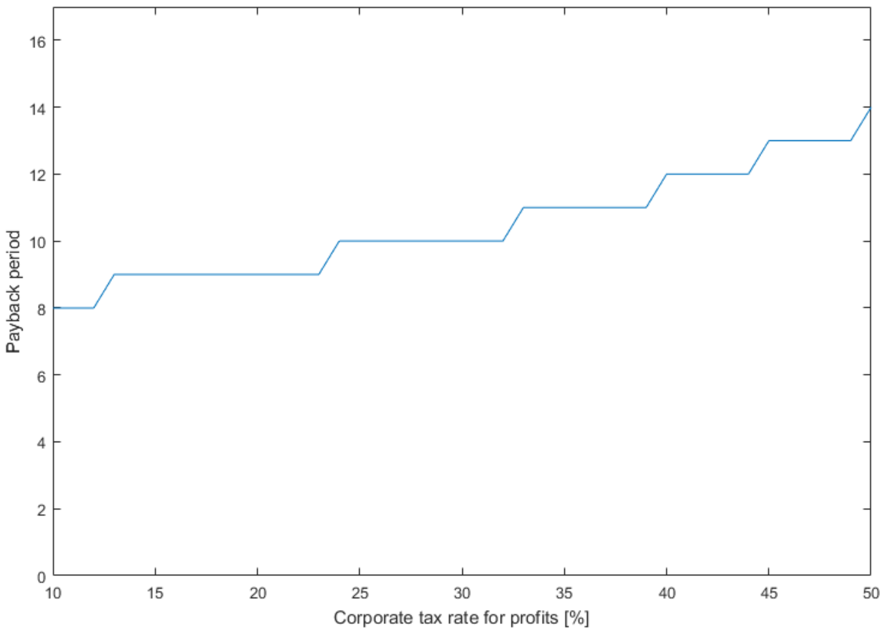


Figure 7.12: Payback-period, corporate tax rate varied

A best case scenario and a worst case scenario has been analyzed to know in which span the LCOE and CO_2-avoidance cost will lie within. Due to a lack of knowledge in the beginning, leading to operational inefficiencies, the LCOE and CO_2-avoidance cost is thought to be closer to the worst case scenario than the best case scenario. Parameters as the coal price can not be much influenced, but operational inefficiencies and lack of experience, leading to decreased operating hours, increased energy losses and increased plant construction costs will push the levels towards the worst case. After experience has been gained, the costs are assumed to get closer to the best case scenario. When choosing best and worst case scenarios, realistic conditions have been considered. The chosen scenarios are presented in Table 7.2. The plant lifetime and depreciation time have been fixed to 25 years since the lifetime is closely related to the annual operating hours, and varying both would cause a double effect. The rate of return has also been fixed to 8%, since investors assign a fixed rate of return when comparing investments. It should also be pointed out that values close to the worst case and the best case are unlikely, since it means that all variables are pushed to the positive or negative limits at the same time

Table 7.2: Parameters that are varied in the best/worst case analysis, and their best, base and worst case values

Parameter	**Worst case**	**Base case**	**Best case**	**Unit**
Annual operating hours	7000	7500	8000	$hours$
Inflation rate	3	2	1	%
Normalized electricity price to grid supplier‡	0.7	1	1.1	–
Corporate tax rate on profits	40	32	20	%
CO_2-capture efficiency	96	98	99	%
Normalized oxygen carrier performance ratio	2	1	0.6	$\frac{OC\ price}{OC\ average\ lifetime}$
Normalized equipment and installation cost	1.2	1	0.8	–
Investment cost factor *	3.5	1.69	1.6	–
Coal price §	1.2	1	0.8	–

‡ At the beginning of the plant lifetime. Normalized value 1 = 0.0563 $\frac{€}{kWh}$

§ At the beginning of the plant lifetime. Normalized value 1 = 60 $\frac{\$}{tonne}$

* Factor multiplied with the equipment cost to estimate the total investment cost. The base case number corresponds to the results indicated by the base case calculations

The results from the best and worst case scenario analysis is presented in Table 7.3, with the base case as a reference point. How well a real plant would perform in the range between the worst and best case depends on both controllable factors as knowledge and experience to reduce contingency costs, engineering costs and costs related to plant operational inefficiencies; but also on uncontrollable factors such as the coal price.

Table 7.3: Results from the best and worst case scenario analysis

Key performance indicator	**Worst case**	**Base case**	**Best case**	**Unit**
Levelized cost of electricity	90	48	36	$\frac{€}{MWh}$
CO_2**-avoidance cost**	55	23	13	$\frac{€}{tonne\ CO_2}$
Payback-period	25+ ‡	10	5	$years$

‡ The payback-period is greater than the plant lifetime of 25 years

8 Comparison with other techno-economic studies about industrial scale chemical looping combustion

A few number of techno-economic evaluations have been performed for chemical looping combustion power plants in industrial scale as well as for other CCS-technologies. This chapter compares the results.

The paper written by Lyngfelt, A. [3] about design and costs of a 1000 MW_{th} CLC power plant has already been mentioned and used frequently as a reference in this study. Instead of considering costs of every component of the plant, the additional costs compared to a conventional coal fired power plant are estimated. As mentioned on page 24 in Chapter 6, this is a method with a low uncertainty, but disables the ability to estimate the whole plant cost, to look at the trends during the plant lifetime and to estimate a LCOE. Lyngfelt predicts an efficiency penalty, which is the efficiency reduction compared to a conventional coal fired power plant, to 3.9%. The results from the process models developed in this work predict the efficiency penalty to be around 5%. The difference is due to a higher CO_2-compression cost that has been modeled and calculated in this study, while it was approximated from another study in Lyngfelt's paper. The CO_2-avoidance cost has in this work been predicted to be slightly higher than in Lyngfelt's paper, but have the same range of magnitude.

In [27], fluidized bed models in Aspen plus of a 10 MW_{th} CLC power plant with gaseous fuel was upscaled to 500 MW_{th} and an economic evaluation was carried out. Moreover, a brief sensitivity analysis was conducted. Overall, the equipment cost was estimated to be far lower than in this work, but the additional direct and indirect capital costs were approximated significantly higher, resulting in a total investment cost of a similar magnitude. When extracting a comparable result from the sensibility analysis done in that study for a similar fuel price and OC lifetime, a value of approximately 80 $\frac{€}{MWh}$. That is a higher number than what is estimated in this study.

A Chinese research group has made its contribution to the techno-economic evaluation of an industrial CLC power plant through Xixian's paper [28]. Notable is that the electrical efficiency is in the same range as what is predicted in this work and the paper by Lyngfelt [3]. The LCOE is somewhat higher, while the CO_2-avoidance cost is lower. This must be due to a higher estimation of the costs related to the reference coal fired power plant. However, a detailed comparison is not possible since the corresponding paper is only available in Chinese.

Table 8.1: Comparison with other techno-economic studies about industrial scale CLC power plants

Article	**Plant size** $[MW_{th}]$	**Fuel**	η_{el} * [%]	**LCOE** $\left[\frac{€}{MWh_{el}}\right]$	CO_2**-avoidance cost** $\left[\frac{€}{tonne\ CO_2}\right]$
This study	1000	Coal	41	48	23
Lyngfelt, Chalmers University of Technology [3]	1000	Coal	- ‡	-	20
Porrazzo, Heriot-Watt University [27]	500	Gas §	52	80	-
Xixian, Huazhong University of Science and Technology [28]	1319	Coal	40	63	9

‡ Suggests instead an efficiency penalty of 3.9%

§ Syngas or natural gas

9 Discussion and conclusions

When it comes to introducing new technology to the market, there are three main obstacles. The first is to overcome technical issues. Secondly, this must be done in an economically feasible way so that it is competitive with other available technologies, which in this case are other CCS technologies. Lastly, when it comes to an industry with strict regulations related to environmental issues as the electricity production industry, it must win political support and support by the society. Positive results of the two first points may then consequently favour the third one. Hopefully, with my work, I have contributed to overcome the two first mentioned obstacles.

The electrical efficiency of the proposed plant of 41% is in the same range as in the other comparable studies in Table 8.1. Furthermore, it is also in line with the 40-41% that is predicted by the ENCAP project in [21]. This is an indication that the developed models perform according to expectations. With the results from this work together with the other mentioned in Table 8.1. It can be concluded that the efficiency penalty compared to a conventional coal fired plant is approximately 5 % including CO_2-compression.

As can be seen in Table 5.10, the waste water has a volumetric flow of 12.4 $\frac{m^3}{s}$ and a temperature at the outlet of 65°C. That is energy that is lost to the environment and is a potential heat source to a district heating system. Even if only heat corresponding to a 5°C decrease of temperature of the waste water, 7.5 MW would be recovered. If the double heat could be extracted, equal to a 10°C decrease, 15 MW would be recovered.

An important fact to consider regarding the CO_2-avoidance cost, is if the final CO_2-storage is not located next to the plant, transportation costs of CO_2 will be added to the global CO_2-avoidance cost. This expense is not taken into account in this work. It is expected that the CO_2-transportation cost would be around 10 $ per 100km and costs related to the storage itself and monitoring would add up to another 5 $ per tonne [29].
Another point that might be added to the overall cost is the cost for removal of sulphuric components in the fuel reactor flue gas stream as well as equipment for dust removal, if the content is higher than the allowed limit for the CO_2-storage.

As expected, the most expensive process units are the air reactor, fuel reactor, compressors and steam turbines. These units together value up to 83% of the equipment cost. The cost of the compressors constitutes such a big portion due to the big number of compressors used in the CO_2-compression. The costs of the carbon stripper and post oxidation chamber are uncertain though, due to a lack of information since they are not standard components in a conventional coal fired power plant. From Figure 6.3, it can be concluded that the CO_2-compression has a significant portion of the equipment cost, as much as 16% for a CLC power plant. That whole cost fraction is not present in a conventional coal fired power plant.

The total investment cost that is calculated from the total equipment cost is a point of discussion. Recommendations from IEAGHG in [18] have been followed for this part of the cost estimation and the result is that including the interest payments during construction which is a non-negligible cost, the total investment cost sums up to 203% of the equipment cost. Without interest payments during construction, the same number equals 169%. These are both numbers far lower than what is suggested in [31], even though what is suggested in [31] is thought to be more suitable for plants of smaller scale and includes other factors as working capital. Nevertheless, the value of 203% could be significantly higher in reality. Thus, the upper limit was varied to a large extent in the sensitivity analysis, and it can be concluded that it has a great impact on both the LCOE and CO_2-avoidance cost, seen in Figure 7.9 in the sensitivity analysis. The investment cost factor is strongly dependent on if the plant is totally built as a grassroot design or partially retrofitted from an old location of power plant. In the case of partial retrofitting or reconstruction of an old power plant, costs for land, permission and buildings may significantly be reduced, leading to a lower investment cost factor.

The single greatest source of expense to the operating cost and also the total plant lifetime cost is the fuel cost. Naturally, a variation of the coal price has a heavy impact on the LCOE, seen in Figure 7.10 in the sensitivity analysis. The same was concluded in [28], where the fuel cost is identified as the greatest cost contribution. A variation in coal price also affects the CO_2-avoidance cost, but in a smaller extent because the reference plant also is negatively affected by a high coal price.

In addition to the already mentioned influence of the coal price, the sensitivity analysis reveals that the parameters that have a big influence on the LCOE and CO_2-avoidance cost are the annual operating hours, assigned rate of return, equipment cost and the magnitude of the costs related to the investment except for the equipment. A high OC price or a short OC lifetime could also have a negative influence on the LCOE and CO_2-avoidance cost, but has a lower impact than expected. The OC price is difficult to predict, since a commercialization of the CLC-technique could strongly increase the demand of materials used for OC, which influences the price. The results from a varied plant lifetime is also interesting. Even though the LCOE is not strongly affected by the plant lifetime, it affects the total profit of the plant to the owner, which is related to the difference between the selling price of electricity and the LCOE. The longer the plant is active and is operated profitable, the more profit can be made. In other words, the cumulative cash flow curve shown in Figure 6.9 would rise higher the longer the plant lifetime is, provided that the plant operated profitably. How the selling price of electricity changes from year to year is another difficult parameter to predict. Its price has a big impact on the payback-period, which can be concluded by Figure 7.11.

One conclusion that can be made from the sensitivity analysis is that a CLC power plant is more sensitive toward changes that affect the plant economics negatively. This conclusion is made by the fact that when parameters that are of importance both for the CLC plant and the reference plant such as annual operating hours, assigned rate of return, inflation rate and coal price are altered in a negative way, the CO_2-avoidance cost increases.

The biggest uncertainty with a cost evaluation of a CLC power plant is concluded to be the cost of direct and indirect capital expenses added on top of the equipment cost and the price development of oxygen carriers, fuel and selling electricity price. The first point due to lack of experience when it comes to construction of CLC power plants. The price developments are seen as uncertainties because it is in general problematic to forecast price trends, and all three mentioned utilities have a potential to affect the LCOE and CO_2-avoidance cost. It is also affected by if the plant is is a complete grassroot construction or if it partly can be retrofitted from an old power plant, to save costs related to land, permissions, buildings and so on.

The comparison between different studies made in Table 8.1 together with the best and worst case scenario study, gives relatively certain indications about in which ranges an industrial scale CLC plant performance would be. This conclusion is presented in Table 9.1. These results show that CLC is a competitive CO_2-capturing technology in comparison with other CO_2-capturing technologies. In the future, when the technology has become more mature so that more experience has been gained about how to build and operate a full scale plant, and the performance of oxygen carriers has been improved, there is a chance to approach the lower boundary of the LCOE and CO_2-avoidance cost in the best/worst-case scenario and perhaps even surpass it. It is unlikely in the present time, but the future offers new possibilities.

Table 9.1: Expected values of a 1000 MW_{th} CFB-CLC coal fired power plant key performance indicators

$\boldsymbol{\eta_{el}}$ [%]		**LCOE** $\left[\frac{€}{MWh_{el}}\right]$		$\boldsymbol{CO_2}$**-avoidance cost** $\left[\frac{€}{tonne\ CO_2}\right]$	
Lower bound	Upper bound	Lower bound	Upper bound	Lower bound	Upper bound
40	41	45	65	20	30

As final words, a general comment and opinion is that the value that CO_2-certificates traded on at the public market are too low. At the moment of writing, the price of a CO_2-certificate is 5 $\frac{€}{tonne\ CO_2}$, which makes the economic driving force for investments in CO_2 abatement technologies too low. Regulations should be implemented to raise the price.

10 Further work

The right time for CCS technologies to be introduced to the market is right now, during the transition phase when dirty, heavy greenhouse gas emitting energy production technology is being replaced by cleaner environmentally friendly technology. In a far perspective, it is likely that all energy production will be made by renewable energy sources, so the time for CCS technology is now, before that stage. Therefore, it is important to speed up the development of the CCS technologies, including CLC. The scientific society is invited to contribute with further research about CLC, and all contributions are welcomed. Some points of further work are suggested below.

With a global perspective, the development of the CLC technology for large scale applications needs to be speeded up. As several small scale pilot plants today exist, the next step should to plan the construction of an intermediate scale plant with a capacity of 20 MW_{th} or even 100 MW_{th} as soon as possible to meet the demand of a cost effective CO_2-capturing technology for power plants.

In a local point of view, the process models may be further developed. Process blocks representing an air separation unit could be implemented in the Aspen model. The flowsheet configuration in the Ebsilon model could perhaps be further improved to increase the net efficiency. Each process unit could also be looked into more detail to determine the most economically favourable design specifications, as heat exchanger terminal temperature differences, number of heat exchangers, steam turbine design and how to take use of the waste water heat in an efficient way. It would also be interesting to adapt the models for use of other fuels, as biomass.
Something else that could be improved in the Ebsilon model are the properties of the CO_2 that is to be stored or transported. In a late phase of the project, it was discovered that the temperature of the CO_2 to storage is slightly to high and concentration of SO_2 is too high compared to recommendations by [26]. The rest of the concentrations of impurities and the pressure is on an acceptable level. The temperature can easily be decreased by increasing the capacity of the heat exchangers in the CO_2-compression section and increase the cooling water flow if needed. The excess of SO_2 is an issue of higher concern, but could either be solved by using a fuel with a low sulphur content, be separated together with the condensing water or in the worst case, an external separation equipment is necessary. This however should not lead to a major additional cost since the concentration of SO_2 in the CO_2-storage stream already is relatively low and flue gas streams are smaller compared to conventional power plants.
Regarding the economic evaluation, a deeper look should be done about the magnitude of cost contributions to the investment cost for expenses except for the installed equipment costs. Cost contributions as piping, land, buildings, engineering and instrumentation and control systems could be re-estimated with more industry-related experience, to improve the accuracy of the investment cost prediction. A point of improvement of the sensitivity part is to make a multi-variable analysis and vary several parameters at a time, and not only one-dimensional changes as in this study. Furthermore, it would increase the user-friendlyness if the base case calculation script could be merged with the sensitivity script, so that the necessity of changes in both scripts can be avoided when making changes of the cost functions.
Moreover, it would be helpful to improve the accuracy of the results by performing a statistical evaluation of the economic results.
It would also be of interest to make a thorough comparison about advantages and disadvantages in relation to other CCS technologies, in both a technical and economic way.

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A Appendix

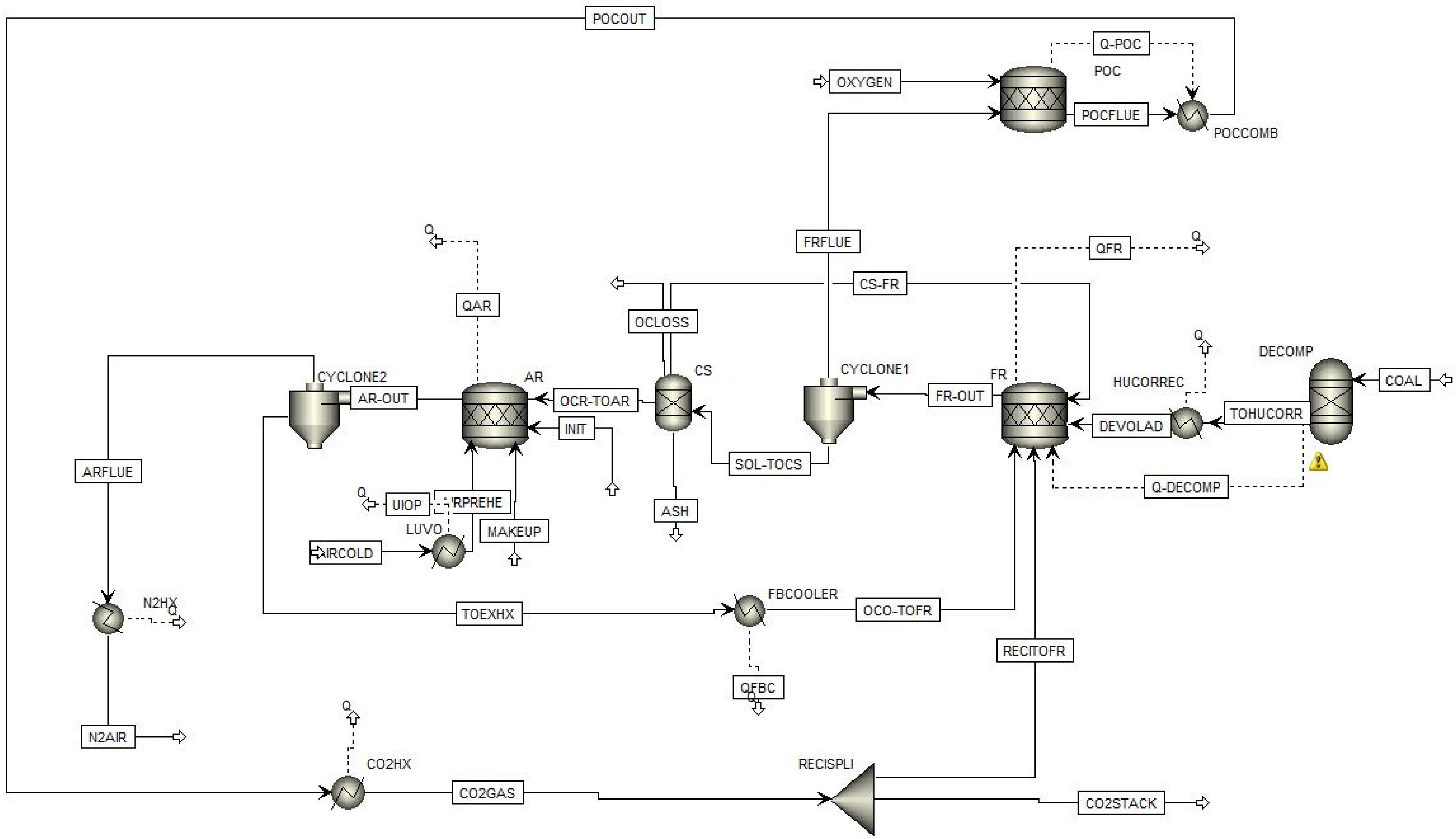

Figure A.1: Process flow diagram of Aspen model

Table A.1: Blocks in the Aspen Plus model

Process unit blocks	
N2HX	Represents the heat extracted from the air reactor flue gas to superheat steam, pre-heat feedwater and pre-heat the AIRCOLD stream
CYCLONE2	Cyclone to separate oxygen carrier from the gas
CO2HX	Represents the heat extracted from the fuel reactor flue gas to superheat steam and pre-heat feedwater
LUVO	Luvo air pre-heater
AR	Air reactor
CS	Carbon stripper
FBCOOLER	Fluidized bed cooler
CYCLONE1	Cyclone to separate solids from gas
RECISPLI	Splitter to adjust the recirculated FR flue gas stream for fluidization
POC	Post oxidation chamber
FR	Fuel reactor
HUCORREC	Corrects the lower heating value
DECOMP	Represents the coal gasification

Table A.2: Streams in the Aspen Plus model

Material streams		Energy streams	
FRFLUE	Fuel reactor flue gas	QN2	Heat recovered from the air reactor flue gas stream
POCOUT	Post oxidation flue gas	QCO2	Heat recovered from the fuel reactor flue gas stream
OXYGEN	Pure oxygen from air separation unit	QAR	Heat absorbed by the air reactor walls
POCFLUE	Ficticious post oxidation flue gas	QFBC	Heat recovered from the fluidized bed cooler
CO2GAS	Cooled flue gas stream	Q-POC	Heat produced by the adiabatic post oxidation process
RECITOFR	Recirculated stream to flue gas reactor for fluidization	QFR	Heat produced in the fuel reactor
CO2STACK	Flue gas stream that is lead to CO2-compression and water separation	QHU	Energy removed to correct for the lower heating value of coal defined by Aspen Plus
TOHUCORR	Gasified coal stream	UIOP	Heat provided to the air pre-heater
DEVOLAD	Gasified coal after enthalpy correction		
FRFLUE	Flue gas from fuel reactor with solids separated		
FR-OUT	Outlet from fuel reactor		
SOL-TOCS	Oxygen carrier stream containing coal and ash		
CS-FR	Unburnt coal		
OCLOSS	Oxygen carrier loss		
ASH	Ash output		
OCR-TOAR	Oxygen carrier to air reactor		
INIT	Initial oxygen carrier supply for start-up		
MAKEUP	Oxygen carrier make-up stream		
AIRPREHE	Preheated air feed		
AIRCOLD	Fresh air feed		
AR-OUT	Outlet stream of the air reactor		
ARFLUE	Gas outlet from CYCLONE2		
N2AIR	Cooled down flue gas from the air reactor that is released to the atmosphere		

B Appendix

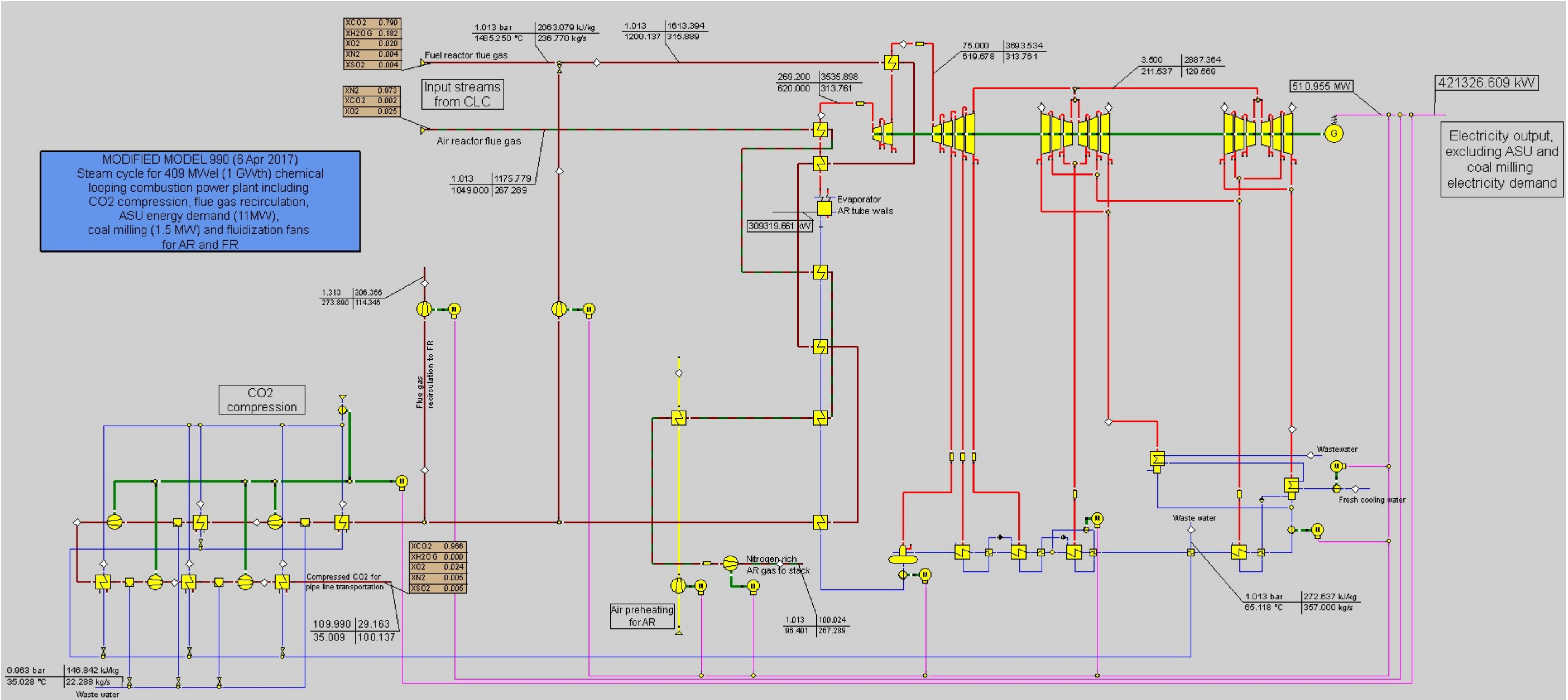

Figure B.1: Process flow diagram of steam cycle in Ebsilon Professional