Conventional Exergetic and Exergoeconomic Analyses of a Power Plant with Chemical Looping Combustion for CO_2 Capture*

Fontina Petrakopoulou**[1], George Tsatsaronis[2], Tatiana Morosuk[3]

[1,2] Institute for Energy Engineering, Technische Universität Berlin, Germany
[3] Institute of Marine Propulsion Plants Operation, Maritime Academy of Szczecin, Poland
E-mails: [1] f.petrakopoulou@iet.tu-berlin.de, [2] tsatsaronis@iet.tu-berlin.de, [3] morozyuk@iet.tu-berlin.de

Abstract

Exergy-based methods can be used as a tool for examining, comparing and assessing thermodynamic systems. In this paper, an exergoeconomic analysis is used to evaluate a power plant with *chemical looping combustion* (CLC) for CO_2 capture. This oxy-fuel plant is compared, from an exergetic and an economic perspective, to a conventional, reference power plant without CO_2 capture. The exergetic analysis shows decreased exergy destruction in the CLC reactors, compared to the exergy destruction in the conventional combustion chamber of the reference case; thus, the irreversibilities caused by combustion in the CLC are reduced. However, due to the addition of the CO_2 compression unit, the overall exergetic efficiency of the plant with CLC is lower than that of the reference plant by approximately 5 percentage points. The economic analysis confirms a significant increase in the investment cost of the CO_2 capture plant, due to the addition of the units for CO_2 compression and CLC. Thus, the cost of electricity is 24% higher for this plant in comparison to that of the reference case. Nevertheless, when compared to the reference plant with CO_2 capture with monoethanolamine, the plant with CLC was found to be a more economical option. Since CO_2 abatement must be realized in the future, given expected environmental or tax measures, CLC provides relatively low cost carbon dioxide capture and it, therefore, appears to be a promising option for reducing greenhouse gases emitted by power plants using fossil fuels.

Keywords: CO_2 capture, chemical looping combustion, exergetic analysis, exergoeconomic analysis.

1. Introduction

The mitigation of environmental pollution through CO_2 capture in power stations is an area that drew intense attention from a large group of scientists a little more than two decades ago (Herzog, 2001). Although several possible methods for capturing CO_2 have been developed in such a short time (Kakaras et al., 2005), few appear promising with respect to efficiency and cost.

In this paper, a combined cycle power plant that performs CO_2 capture using *chemical looping combustion* (CLC), is compared to a 411 MW power plant without CO_2 capture (reference plant). Previous studies (Hossain and de Lasa, 2008; Rubin and Rao, 2002) show that CLC has the potential to become an efficient and relatively economical technology for capturing produced CO_2 from power plants. The idea was first introduced in 1954 as a way to produce pure CO_2 from fossil fuels, using two interconnected fluidized-bed reactors (Lewis and Gilliland, 1954). In 1968, it was proposed by Knoche and Richter as an option for reducing irreversibilities in combustion processes (Knoche and Richter, 1968, 1983) and, in the 1990s, it was recognized as a way to capture CO_2 emitted through fossil fuel use to reduce the climate impact (Ishida and Jin, 1994). CLC has been identified as having inherent advantages for carbon dioxide separation with relatively low thermodynamic inefficiencies (Lewis and Gilliland, 1954; Brandvoll and Bolland, 2004).

To evaluate the operation and feasibility of CLC technology for CO_2 capture, exergy-based methods were applied in this study. The advantages of an exergetic analysis over a conventional energy analysis are well established (Bejan et al., 1996; Moran and Shapiro, 2000; Tsatsaronis and Cziesla, 2004-2007). Moreover, an economic analysis has been conducted to analyze the total cost of construction, operation and maintenance associated with the power plant.

The exergoeconomic analysis, an appropriate combination of an exergetic analysis with an economic analysis (Tsatsaronis et al., 1991; Bejan et al., 1996; Tsatsaronis, 1999; Tsatsaronis and Cziesla, 2002), can be considered as an exergy-aided cost reduction approach. In the exergoeconomic analysis, the principle of *exergy costing* is used to assign monetary costs to all energy streams, as well as to the exergy destruction incurred within each component of a plant. Important information about the trade-offs between exergy destruction and the investment cost of components can then be used for iterative design improvements of the plant.

The purpose of this paper is to present and evaluate the results of a detailed exergoeconomic analysis of a plant capturing CO_2 with CLC, as compared to the performance of a conventional plant, referred to as the *reference plant*.

This paper is part of a study analyzing different concepts of CO_2 capture from energy conversion systems exclusively generating electricity (Petrakopoulou et al., 2009a – 2009d, 2010).

2. Methodology
2.1 Exergetic analysis

The exergetic analysis is conducted with a system of balance equations, stated at the component-level, and a general equation for the overall system. The exergy of the

**Corresponding Author

product can be defined as the exergy of the desired output resulting from the operation of a component, while the exergy of the fuel is the expense in exergetic resources for the generation of this desired output. The exergy destruction within a component is then calculated as the difference between the exergy of the fuel and the exergy of the product ($\dot{E}_{D,k} = \dot{E}_{F,k} - \dot{E}_{P,k}$).

In dissipative components, such as condensers, intercoolers and throttling valves, exergy is destroyed without any useful product in the component itself; thus, no exergetic purpose for these components can be defined (Bejan et al., 1996; Lazzaretto and Tsatsaronis, 2006). The essential role of these components is to serve other components, or the overall system.

For the component-level analysis, all streams exiting a component are considered either as part of the product, or they are used in the definition of the component's fuel. Thereafter, exergy loss ($\dot{E}_{L,tot}$) is defined only for the overall system.

The exergetic efficiencies of the kth component and of the overall system consisting of NC-components are defined by Eqs. (1a) and (1b), respectively:

$$\varepsilon_k = \frac{\dot{E}_{P,k}}{\dot{E}_{F,k}} = 1 - \frac{\dot{E}_{D,k}}{\dot{E}_{F,k}} \tag{1a}$$

$$\varepsilon_{tot} = \frac{\dot{E}_{P,tot}}{\dot{E}_{F,tot}} = 1 - \frac{\sum_{k=1}^{NC} \dot{E}_{D,k} + \dot{E}_{L,tot}}{\dot{E}_{F,tot}} \tag{1b}$$

Here the system's exergy of the product ($\dot{E}_{P,tot}$) is the net power produced, whereas the system's exergy of the fuel ($\dot{E}_{F,tot}$) is the sum of the fuel and the air exergy provided to the plant. General guidelines for the definition of exergetic efficiencies have been proposed by Lazzaretto and Tsatsaronis (2006).

A useful variable of the exergetic analysis is the exergy destruction ratio defined by Eq. (2):

$$y_{D,k} = \frac{\dot{E}_{D,k}}{\dot{E}_{F,tot}} \tag{2}$$

This ratio is a measure of the contribution of the exergy destruction within each component to the reduction of the overall exergetic efficiency.

With the aid of an exergetic analysis, the main sources of irreversibilities within a plant are identified and are then linked to costs in the exergoeconomic evaluation.

2.2 Economic Analysis

For the economic analysis, the *total revenue requirement* (TRR) method has been implemented (Bejan et al., 1996). The first step of the analysis is to calculate the *fixed capital investment* (FCI) of the plant (Tsatsaronis and Winhold, 1984; Tsatsaronis et al., 1990; U.S. Department of Energy, 2000; Gas Turbine World Handbook, 2006). Costs are then escalated to the reference year 2008 using the *chemical engineering plant cost index* (CEPCI) as published in *Chemical Engineering Magazine*.

The next step that allows the implementation of the TRR method is to meet assumptions concerning market conditions, plant operation and plant construction. To estimate the *total capital investment* (TCI) of a project, the total expenses of the construction and the overall operation of the plant must be assessed. The economic life of the plant is assumed to be 20 years, while its life for tax purposes is 15 years. The date of commercial operation of the plant is set to 01.01.2012 with a two year period of design and construction preceding. The average capacity factor of the plant is assumed to be 85%, resulting in an annual operation of 7446 hours. The *total capital requirement* (TCR) of the plant comes from equity and debt financing with a 10% average annual rate of the cost of money. Lastly, all costs, except that of the fuel, increase annually with an inflation rate of 3%; the cost of natural gas is assumed to be 7€ per GJ of *lower heating value* (LHV: 50.015 MJ/kg) and its average annual increase amounts to 4%.

With these assumptions the levelized TRR is calculated for a levelization period of 10 years. After completion of the economic analysis, the exergy costing procedure and an exergoeconomic evaluation follow.

2.3 Exergoeconomic Analysis

In an exergoeconomic analysis a specific cost c is assigned to each exergy stream of the plant. The specific cost of stream i, c_i, multiplied by the exergy of the same stream, $\dot{E}_i$, calculated in the preceding exergetic analysis, provides the cost rate $\dot{C}_i$, associated with the ith exergy stream:

$$\dot{C}_i = c_i \dot{E}_i \tag{3}$$

The cost balance for the kth component is

$$\sum_{i=1}^{i=n} \dot{C}_{i,k} - \sum_{j=1}^{j=m} \dot{C}_{j,k} + \dot{Z}_k = 0 \tag{4}$$

where,

$\sum_{i=1}^{i=n} \dot{C}_{i,k}$ is the sum of the cost rates associated with the n steams entering component k,

$\sum_{j=1}^{j=m} \dot{C}_{j,k}$ is the sum of the cost rates associated with the m streams leaving component k, and

$\dot{Z}_k$ is the cost rate associated with the investment cost and the operating and maintenance costs of component k. $\dot{Z}_k$ is known from the preceding economic analysis.

An important variable of the exergoeconomic evaluation is the relative cost difference. For a given component k, the difference between the specific cost of the product, $c_{P,k}$, and the specific cost of the fuel, $c_{F,k}$, depends on the cost of exergy destruction, $\dot{C}_{D,k}$, and the investment-based cost $\dot{Z}_k$. The relative cost difference, r_k is defined by

$$r_k = \left(\frac{c_{P,k} - c_{F,k}}{c_{F,k}} \right) \tag{5}$$

The contribution of the capital cost to the total sum of costs associated with capital investment and exergy

destruction is expressed by the exergoeconomic factor f_k that is defined by Eq. (6):

$$f_k = \frac{\dot{Z}_k}{\dot{Z}_k + \dot{C}_{D,k}} \tag{6}$$

The relationship between the monetary impact of each component's exergy destruction and investment is then examined. If necessary, design changes to improve the cost effectiveness of the plant are proposed. The objective is to reduce the cost associated with a unit of the product of the overall plant.

3. The plants
3.1 The reference plant

The reference plant, which includes no CO_2 capture and is used as a basis for the evaluation of the plant including CLC, is a combined cycle with a three-pressure-level *heat recovery steam generator* (HRSG) and one reheat stage. The plant has only one product – electricity – and works with natural gas that was assumed here to be pure methane. The specific CO_2 emissions of the plant are 339 g/kWh. A simplified diagram of the process is shown in Figure 1.

3.2 Chemical looping combustion (CLC)

The process including CLC is a combined cycle with close to 100% CO_2 capture. The fuel mass flow is the same as in the reference case. However, the net power output of the plant (375 MW) is lower by almost 9%, due to the additional CO_2 treatment. The process includes two HRSGs and its main configuration is similar to that of the reference case. The main HRSG has three pressure levels and one reheat stage, whereas the second one is a one-pressure-level HRSG. No cooling of the *gas turbine* (GT) was taken into account, because of the lower inlet temperature in comparison to the cooled expander of the reference case. A diagram of the process is shown in Figure 2.

In this plant, the compressed air and the preheated methane at 300°C are sent to the CLC unit (black box). The two streams exiting the CLC unit are the combustion products, consisting of CO_2 and water vapor, and the oxygen depleted air, consisting of 15% v/v O_2. The oxygen depleted air, that replaces the flue gas of the conventional combustion chamber of the reference case, exits the GT at 659 kg/sec and is led to the main HRSG of the plant. The high-pressure steam produced in this cycle has a temperature of 500°C that is lower than the one of the reference case, due to the lower temperature at the exit of the GT.

On the other side, the combustion products are expanded in the CO_2/H_2O expander, and they are then sent to the secondary HRSG, where they provide a large part of their thermal energy for producing steam. It has been suggested that the inlet temperature of the CO_2/H_2O expander should be as low as 900°C, to increase the conversion of the fuel in the fuel reactor and the energy available for oxidation of the metal in the oxidation reactor (Brandvoll and Bolland, 2004; Bolhàr-Nordenkampf et al., 2008). With this lower temperature, a lower cost for the expander is also achieved. A higher temperature, equal to that of the oxygen-depleted air at the outlet of the oxidation reactor could bring an increase in the energetic efficiency of about 1 percentage point (to 54.2%), if the CH_4 conversion is assumed to remain constant. After the expansion and the

secondary HRSG, the CO_2-rich gas is cooled down further in a flue gas condenser to 30°C, in order to condense a large part of the included water vapor. The almost pure CO_2 (94% v/v) is then compressed in four intercooled stages with similar pressure ratios to a final pressure of 103 bar. Carbon dioxide at that pressure and 30°C is in liquid phase and is ready for transport and sequestration. This CO_2 compression itself, without inclusion of the intermediate coolers, causes an approximately 2 percentage point decrease in the overall efficiency of the plant.

In previous publications, CLC has been examined using natural gas (Abad et al., 2006, 2007; Bolhàr-Nordenkampf et al., 2008; Brandvoll and Bolland, 2004; Kolbitsch et al., 2008; Lyngfelt and Thunman, 2005; Naqvi and Bolland, 2007), synthetic gas (Abad et al., 2006, 2007; Jin and Ishida, 2004; Klara, 2007) or hydrogen (Brandvoll et al., 2003). In this plant, the conventional combustion chamber is replaced by two reactors, an *air* or *oxidizing reactor* (AR) and a *fuel reactor* (FR). A metal oxide is used as a solid *oxygen carrier* (OC) between these two reactors. The design of the reactors is based on two interconnected fluidized beds and should have advantages over alternative designs, since good contact between the gas and the solid material is required. In recent years, various arrangements of alternative designs of the fluidized beds (Abad et al., 2006; Lyngfelt et al., 2001; Mattison and Lyngfelt, 2001), as well as different gas turbine configurations (Brandvoll and Bolland, 2004; Naqvi and Bolland, 2007; Naqvi et al., 2005) for integrating CLC in power plants, have been proposed and studied.

Figure 3 shows the CLC process in more detail. Atmospheric air is introduced into the AR, where the metal (or metal oxide) is oxidized. The metal oxide then exits the AR and is fed to the FR, where the transported oxygen reacts with the fuel, to produce CO_2 and H_2O. At the same time, the metal oxide is reduced and led back to the AR continuing its loop between the two reactors. No direct contact between air and fuel takes place, and the carbon dioxide produced can be easily separated after water condensation, without costly energy consumption.

The main two reactions that take place in the chemical looping reactors are shown in Eqs. (7) and (8) and constitute the net reaction described in Eq. (9).

- FR, reduction

$$C_nH_{2m} + (2n+m)Me_xO_y \rightarrow$$
$$(2n+m)Me_xO_{y-1} + mH_2O + nCO_2 \tag{7}$$

- AR, oxidation

$$(2n+m)Me_xO_{y-1} + \left(n+\frac{m}{2}\right)O_2 \rightarrow (2n+m)Me_xO_y \tag{8}$$

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Net reaction

$$C_nH_{2m} + \left(n+\frac{m}{2}\right)O_2 \rightarrow mH_2O + nCO_2 \tag{9}$$
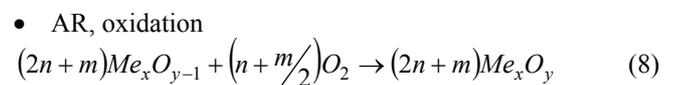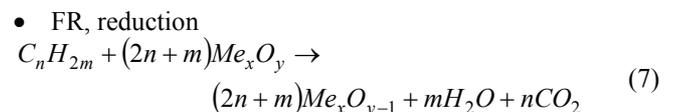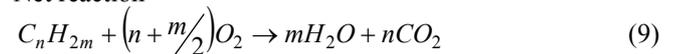
The net reaction of the CLC and the associated net heat release are equal to those of the conventional fuel combustion. The oxidation is an exothermic reaction, whereas the reduction can be either exothermic or endothermic, depending on the fuel and the metal oxide. The OC reduction with CH_4 is endothermic for most of the oxides examined in the literature (Lyngfelt et al., 2008). Conversely, when syngas is used, the metal reduction is always exothermic. This can be considered as an advantage
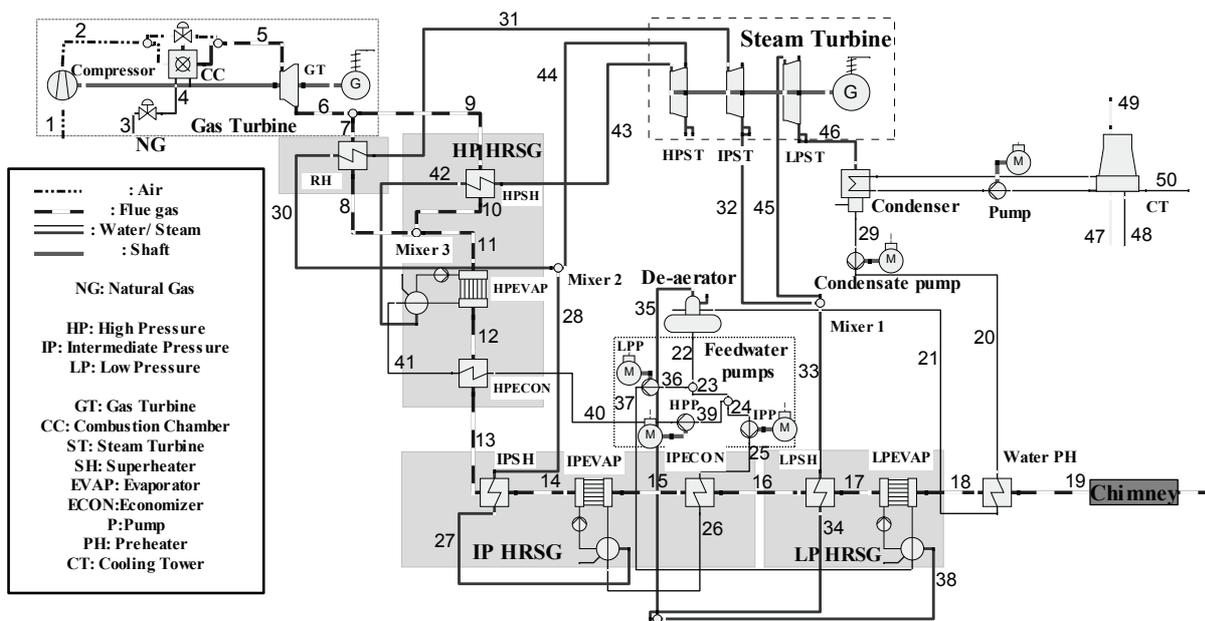
Figure 1. Structure of the Reference Plant.
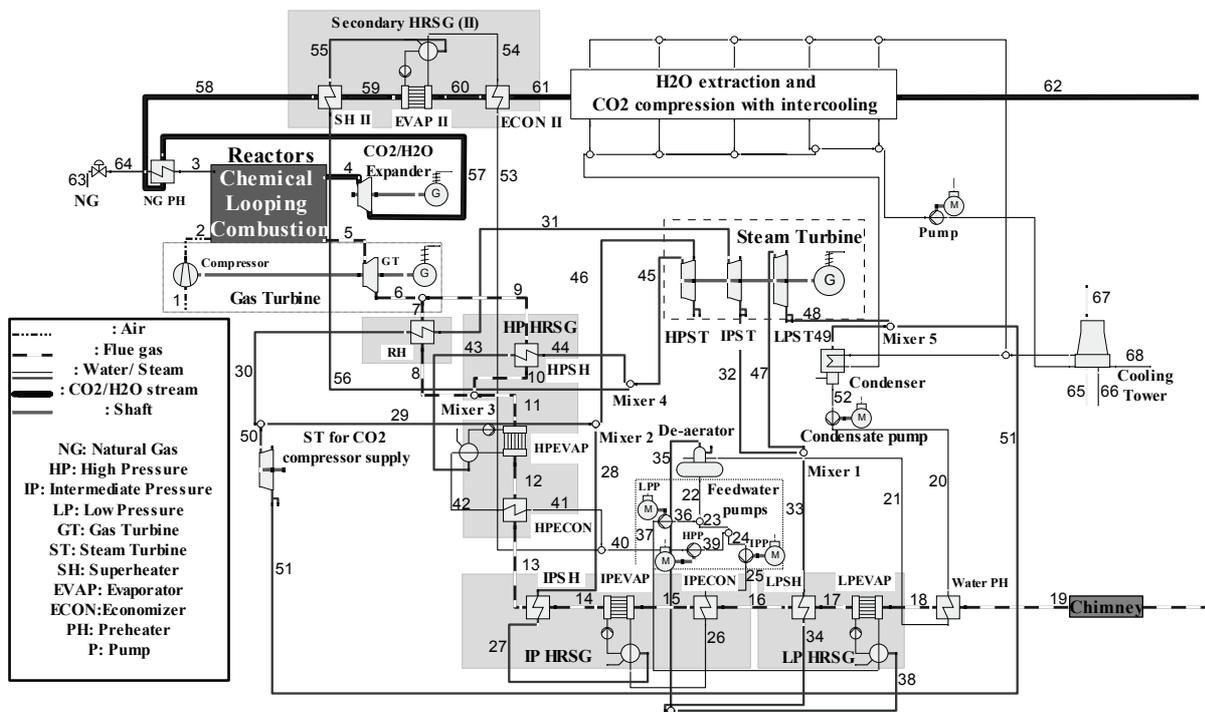
Figure 2. Structure of the plant with Chemical Looping Combustion.
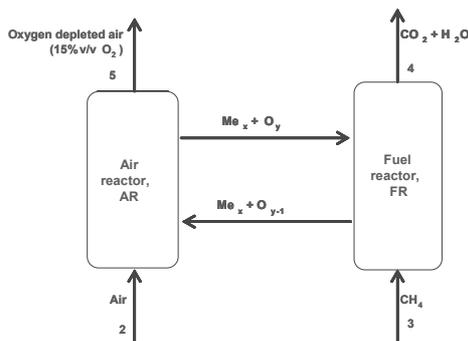
Figure 3. Configuration of Chemical Looping Combustion (the numbering of the streams agrees with Figure 2).

of coal-gas use, since the coal-gas reaction is driven with stronger intensity due to its exothermic character. A part of the produced thermal energy during the oxidation is used in the reduction, if the latter is endothermic.

Many different metals have been suggested for oxygen carriers, mainly based on nickel, Ni (Brandvoll and et., 2003; Jin and Ishida, 2004; Lyngfelt and Thunman, 2005; Johansson et al., 2006; Bolhàr-Nordenkampf et al., 2008; Kolbitsch et al., 2008), iron, Fe (Johansson et al., 2006; Abad et al., 2007; Klara, 2007) and manganese, Mn (Abad et al., 2006; Johansson et al., 2006). Important factors are the reduction and oxidation rate, the chemical and mechanical stability, as well as the price and the environmental characteristics of the oxidizer. Generally Ni and its corresponding oxides show higher oxidation and reduction rates compared to Fe and Mn, as well as greater durability after many repeated cycles. More specifically, experiments with natural gas have shown that Ni reactivity is considerably higher than that of Fe-based oxygen carriers (Johansson et al., 2006). A detailed status of development with respect to oxygen carrier alternatives is presented in Lyngfelt et al. (2008).

In this paper, the CLC reactors are simulated as a black box. In the fuel reactor, 98% of the methane provided is assumed to react with oxygen transferred from the AR. The remaining non-reacting 2% of the fuel is not recycled back to the fuel reactor, but is regarded as a loss. The air ratio - the ratio between the oxygen included in the air and the oxygen needed for stoichiometric combustion - is set to 2.9. This ratio is set to achieve outlet temperatures of the air and the fuel reactors of 1200°C and 930°C, respectively (Lewis and Gilliland, 1954; Abad et al., 2007; Naqvi and Bolland, 2007). In this work, a Ni-based oxygen carrier is considered, in order to achieve close to 100% CH_4 conversion to CO_2.

For the calculation of the reactors costs, data from the publications of Klara (2007), Wolf et al. (2005), as well as Lyngfeld and Thunman (2005) were used. In Wolf et al. (2005), an 800 MW_{th} CLC is studied. Residence times in the fuel and air reactor with a Ni-based oxygen carrier were estimated at 60 and 4.8 sec, respectively. In this paper, these residence times were adjusted to the 700 MW_{th} used and considered to be 53 and 4.2 sec, respectively. The volumetric flow rate of the gases was calculated using the simulation software *EbsilonProfessional 6.0* and the respective volumes of the air and the fuel reactors assumed here were then estimated to be 718 m^3 and 834 m^3, respectively. With a reference unit cost 16.12 million euro for a volume of 180 m^3 in the year 2000 and an exponent of 0.6 (Turton et al., 2002; Klara, 2007), the equipment costs of the AR and the FR were found to be 37.0 and 40.5 million euro, respectively. The installation together with the metal oxide costs were considered to be 20% of the delivered equipment cost. The total FCI of the CLC unit was then found to be 128 million with adjustment to the year 2008, using the CEPSI index. The cost of the metal oxide was not calculated in detail, because of high uncertainty (Lyngfelt and Thunman, 2005). However, even at the highest suggested prices and quantities it could be considered as negligible in comparison to the total cost of the unit.

4. Results and Discussion

Tables 1 and 2 show important variables for selected streams of the reference plant and the plant with CLC, respectively. In both cases, the cost of air and water

Table 1. Calculated variables for selected streams of the reference case without CO₂ capture.

Stream, j	$\dot{m}_j$ [kg/s]	T_j [°C]	p_j [bar]	$\dot{E}_{tot,j}$ [MW]	c_j [€/GJ]	$\dot{C}_j$ [€/h]	Stream, j	$\dot{m}_j$ [kg/s]	T_j [°C]	p_j [bar]	$\dot{E}_{tot,j}$ [MW]	c_j [€/GJ]	$\dot{C}_j$ [€/h]
1	614.5	15.0	1.01	0.96	0.0	0	25	7.2	140.5	25.13	0.68	33.8	83
2	614.5	392.9	17.00	232.25	19.0	15,860	26	7.2	216.6	24.38	1.56	27.2	153
3	14.0	15.0	50.00	729.62	9.2	24,037	27	7.2	222.6	24.38	7.23	21.8	568
5	628.5	1264.0	16.49	741.01	15.3	40,824	28	7.2	237.9	23.16	7.35	22.0	583
6	628.5	580.6	1.06	189.87	15.3	10,460	29	94.6	32.9	0.05	0.44	21.2	33
7	268.5	580.6	1.06	81.11	15.3	4,469	30	72.4	305.1	23.16	79.53	20.3	5,814
8	268.5	447.6	1.05	54.64	15.3	3,010	31	72.4	560.6	22.00	103.42	20.0	7,459
9	360.0	580.6	1.06	108.75	15.3	5,991	32	72.4	317.2	4.10	66.03	20.0	4,762
10	360.0	449.3	1.05	73.68	15.3	4,059	33	22.1	214.1	4.10	18.01	25.0	1,623
11	628.5	448.6	1.05	128.33	15.3	7,070	34	22.1	146.4	4.32	16.96	24.8	1,514
12	628.5	341.2	1.04	84.69	15.3	4,666	35	0.8	146.4	4.32	0.63	24.8	56
13	628.5	257.9	1.04	55.77	15.3	3,073	36	23.0	140.0	3.62	2.12	30.7	234
14	628.5	257.3	1.04	55.59	15.3	3,063	37	23.0	140.0	4.32	2.12	31.1	237
15	628.5	237.6	1.04	49.49	15.3	2,727	38	23.0	146.4	4.32	17.60	24.8	1,570
16	628.5	234.1	1.04	48.43	15.3	2,668	39	65.2	140.0	3.62	6.01	30.7	665
17	628.5	229.3	1.04	47.01	15.3	2,590	40	65.2	141.8	134.56	6.96	31.4	788
18	628.5	156.4	1.03	27.98	15.3	1,542	41	65.2	325.2	130.53	31.88	22.6	2,596
19	628.5	95.3	1.03	16.49	0.0	0	42	65.2	331.2	130.53	71.79	20.5	5,302
20	94.6	32.9	3.73	0.47	25.6	44	43	65.2	560.6	124.00	103.51	20.1	7,489
21	94.6	135.6	3.62	8.18	30.2	889	44	65.2	313.2	23.16	72.22	20.1	5,226
22	95.4	140.0	3.62	8.79	30.7	973	45	94.6	293.0	4.10	83.86	21.2	6,386
23	72.4	140.0	3.62	6.67	30.7	739	46	94.6	32.9	0.05	12.87	21.2	980
24	7.2	140.0	3.62	0.67	30.7	74							

provided are considered to be zero. High values of specific costs are observed for pumped water streams produced using costly mechanical power. The highest values of the cost rate, $\dot{C}_j$, are reached in both plants in streams that have high physical and/or chemical exergy.

The results of the exergetic and exergoeconomic analyses at the component-level for the reference plant and the plant with CLC are presented in Tables 3 and 4, respectively. As expected, the main exergy destruction in both plants occurs within the GT system and it is caused by the chemical reaction taking place in the CC. The components that follow the GT system in exergy destruction are the HRSG and the *low-pressure steam turbine* (LPST). Through the use of CLC, the variable $y_{D,k}$, defined in Eq. 2, is decreased from 30.23% in the combustion chamber of the reference case to 26.56% in the reactors of the plant with CLC. The amount of fuel provided is the same for both plants and the small difference in the exergy of the fuel of the overall plants is due to the higher amount of air needed in the plant with CLC. The absolute amount of the total exergy destruction is

similar for the two plants, with a relatively small difference of about 8 MW. There is, however, a significant difference in the exergy of the resulting product. These differences can be explained by the differences in exergy loss between the two concepts: The exergy loss associated with the CO_2 stream in the plant with CLC corresponds to over 5% of the overall exergy of the fuel, whereas the total exergy loss from the reference plant is about 2% of the overall exergy of the fuel. A comparison of the total y_D in the heat recovery steam generators per produced amount of power in the steam turbines shows that the HRSG of the plant with CLC works less efficiently. This is mainly due to the lower operational efficiency of the added steam turbine supplying the CO_2 compressors and to the lower corresponding steam temperature.

From an economic point of view, when CO_2 capture is considered, there is a considerable increase in the investment cost from 215 € million needed for the reference plant, to 367 € million. Out of the latter, almost 13% is due to the added equipment of the CO_2 compression unit, i.e. the

Table 2. Calculated variables for selected streams of the plant with CLC.

Stream, j	$\dot{m}_j$ [kg/s]	T_j [°C]	p_j [bar]	$\dot{E}_{tot,j}$ [MW]	c_j [€/GJ]	$\dot{C}_j$ [€/h]	Stream, j	$\dot{m}_j$ [kg/s]	T_j [°C]	p_j [bar]	$\dot{E}_{tot,j}$ [MW]	c_j [€/GJ]	$\dot{C}_j$ [€/h]
1	713.5	15.0	1.01	1.11	0.0	0	33	20.1	205.8	4.10	16.21	26.7	1,559
2	713.5	392.9	17.00	269.68	20.1	19,489	34	20.1	146.4	4.32	15.40	26.4	1,466
3	14.0	300.0	17.00	730.74	9.3	24,377	35	0.7	146.4	4.32	0.50	26.4	48
4	68.9	932.2	16.49	113.32	14.6	5,962	36	20.8	140.0	3.62	1.91	33.0	227
5	658.6	1200.0	16.49	693.03	17.1	42,727	37	20.8	140.0	4.32	1.91	33.3	229
6	658.6	517.4	1.06	152.04	17.1	9,373	38	20.8	146.4	4.32	15.91	26.4	1,514
7	313.6	517.4	1.06	72.40	17.1	4,464	39	57.0	140.0	3.62	5.25	33.0	623
8	313.6	439.9	1.05	55.80	17.1	3,440	40	57.0	141.7	134.56	6.09	33.2	728
9	345.0	517.4	1.06	79.64	17.1	4,910	41	53.9	141.7	134.56	5.75	33.2	688
10	345.0	421.1	1.05	57.21	17.1	3,527	42	53.9	325.2	130.53	26.33	24.9	2,358
11	658.6	430.1	1.05	113.00	17.1	6,968	43	53.9	331.2	130.53	59.30	22.5	4,793
12	658.6	341.2	1.05	77.26	17.1	4,765	44	53.9	497.4	124.00	79.73	22.1	6,342
13	658.6	272.7	1.04	53.02	17.1	3,269	45	57.0	489.0	124.00	83.57	22.0	6,623
14	658.6	270.7	1.04	52.39	17.1	3,230	46	57.0	258.4	23.16	59.43	22.0	4,710
15	658.6	232.6	1.04	40.41	17.1	2,492	47	68.5	250.2	4.10	57.90	23.5	4,908
16	658.6	225.8	1.04	38.38	17.1	2,367	48	68.5	32.9	0.05	9.10	23.5	772
17	658.6	221.9	1.04	37.26	17.1	2,298	49	91.1	32.9	0.05	11.95	23.2	1,000
18	658.6	156.4	1.03	20.23	17.1	1,248	50	22.6	257.3	23.16	23.57	22.3	1,891
19	658.6	97.7	1.03	8.95	0.0	0	51	22.6	32.9	0.05	2.85	22.3	228
20	91.1	32.9	3.73	0.46	26.9	44	52	91.1	32.9	0.05	0.42	23.2	35
21	91.1	136.4	3.62	7.97	32.6	934	53	3.2	141.7	134.56	0.34	33.2	40
22	91.8	140.0	3.62	8.46	33.0	1,004	54	3.2	325.2	130.53	1.54	23.2	129
23	71.0	140.0	3.62	6.54	33.0	777	55	3.2	331.2	130.53	3.47	20.0	249
24	14.0	140.0	3.62	1.29	33.0	153	56	3.2	368.9	124.00	3.89	20.0	280
25	14.0	140.4	25.13	1.33	34.7	166	57	68.9	486.7	1.04	59.17	14.6	3,113
26	14.0	216.6	24.38	3.03	28.5	310	58	68.9	388.9	1.04	52.85	14.6	2,780
27	14.0	222.6	24.38	14.03	23.2	1,172	59	68.9	381.2	1.03	52.37	14.6	2,755
28	14.0	252.7	23.16	14.52	23.4	1,223	60	68.9	346.2	1.03	50.34	14.6	2,649
29	71.0	257.3	23.16	73.95	22.3	5,934	61	68.9	319.3	1.03	48.87	14.6	2,571
30	48.4	257.3	23.16	50.38	22.3	4,042	62	38.0	30.0	103.09	36.08	1.9	280
31	48.4	497.4	22.00	64.74	22.3	5,190	63	14.0	15.0	50.00	729.62	9.2	24,037
32	48.4	268.9	4.10	41.76	22.3	3,347							

intercooled CO_2 compression unit and the steam turbine used to support it with power. Moreover, almost 35% of the total fixed capital investment of the plant is due to the reactors of the chemical looping unit. Comparing the total FCI per produced kW of the two plants, a large difference is observed. The resulting investment costs for the reference plant and the case with CLC are 522 €/kW and 980 €/kW, respectively.

The specific cost of the product for the reference case is found to be 20.5 €/GJ, while that of the plant with CLC is 25.5 €/GJ. The resulting levelized costs of electricity for the reference plant and the plant with CLC are 73.9 and 91.7 €/MWh, respectively; thus, CO_2 capture causes an increase in the cost of electricity (COE) of about 24%.

In order to examine the uncertainties related to the reactors' costs, a sensitivity analysis was conducted. In this analysis the cost of the reactors was varied from -50% to +100% of the base cost calculated. The results show that with a 50% lower cost, the levelized cost of electricity produced in the plant is 87.2 €/MWh, representing a small decrease of 5% in comparison to the base case cost estimation of the plant with CLC and furthermore an 18% increase in comparison to the reference case with no CO_2 capture. On the other hand, a 100% underestimation in the price of the reactors, could result in an increase of the cost of electricity by about 10%, with respect to the base-price (100.5 €/MWh).

The cost of the CO_2 avoided can be calculated with Eq. (10) (Rubin and Rao, 2002):

$$\frac{\left(€ / kWh\right)_{capture} - \left(€ / kWh\right)_{reference}}{\left(t_{CO_2} / kWh\right)_{reference} - \left(t_{CO_2} / kWh\right)_{capture}} \tag{10}$$

Here, the levelized cost of CO_2 avoided in the plant with CLC was found to be 53.1 €/t. However, to examine whether CLC is a viable solution for CO_2 capture from an economic viewpoint, this cost is compared to the cost associated with the simplest alternative method: Chemical absorption with *monoethanolamine* (MEA). The plant with post-combustion capture has the same configuration as the reference plant. The changes needed to incorporate post-combustion capture are: (1) the addition of the absorption unit at the outlet of the exhaust gases, (2) steam extraction from the *steam turbine* (ST) of the plant to produce the required thermal energy for complete regeneration of the chemical solvent, and (3) power generation in the ST used to drive the CO_2 compressors. The last two points result in a significant decrease in the power output and, consequently, in the efficiency of the overall system. No solvent losses are taken into account in the simulation. Therefore, the lean sorbent CO_2 loading (mol CO_2/mol MEA) is set to zero, resulting in a relatively high solvent regeneration requirement. Computational calculations are based on Rubin and Rao (2002). Assuming a 10% increase in the capital cost of the plant working with chemical absorption, with respect to the reference plant, the COE of the MEA

Table 3. Results of the exergetic and exergoeconomic analyses at the component-level for the reference case.

Component, k	$\dot{E}_{F,k}$ [MW]	$\dot{E}_{P,k}$ [MW]	$\dot{E}_{D,k}$ [MW]	ε_k [%]	y_k [%]	$c_{F,k}$ [€/GJ]	$c_{P,k}$ [€/GJ]	$\dot{C}_{D,k}$ [€/h]	$\dot{Z}_k$ [€/h]	f_k [%]	r_k [%]
Compressor	242.68	231.30	11.38	95.3	1.56	16.67	19.05	682.8	1297.0	65.5	14.3
CC	729.62	508.76	220.87	69.7	30.23	9.15	13.63	7276.3	926.5	11.3	48.9
GT	551.15	530.67	20.47	96.3	2.80	15.30	16.67	1127.9	1482.3	56.8	8.9
Reheater	26.47	23.89	2.58	90.3	0.35	15.30	19.13	141.9	105.4	42.6	25.0
HPSH	35.07	31.72	3.35	90.5	0.46	15.30	19.16	184.5	149.5	44.8	25.2
HPEVAP	43.64	39.91	3.73	91.5	0.51	15.30	18.83	205.3	183.6	47.2	23.1
HPECON	28.92	24.91	4.00	86.2	0.55	15.30	20.16	220.5	88.6	28.7	31.8
IPSH	0.18	0.12	0.06	69.0	0.01	15.30	34.61	3.1	3.8	55.2	126.1
IPEVAP	6.10	5.67	0.43	92.9	0.06	15.30	20.32	23.8	65.0	73.2	32.8
IPECON	1.06	0.87	0.19	82.5	0.03	15.30	22.06	10.2	5.2	33.5	44.2
LPSH	1.43	1.04	0.38	73.3	0.05	15.30	28.97	21.0	18.3	46.6	89.3
LPEVAP	19.03	15.48	3.55	81.4	0.49	15.30	23.93	195.4	172.8	46.9	56.4
LPECON	11.49	7.71	3.78	67.1	0.52	15.30	30.48	208.5	92.7	30.8	99.2
HPST	31.29	29.18	2.11	93.2	0.29	20.10	23.77	152.9	165.6	52.0	18.3
IPST	37.39	35.21	2.18	94.2	0.30	20.03	24.19	157.4	299.7	65.6	20.7
LPST	70.99	61.35	9.64	86.4	1.32	21.15	29.01	734.3	696.3	48.7	37.2
Condensate Pump	0.04	0.04	0.01	78.8	0.00	19.64	80.52	0.7	6.7	91.0	310.0
HP Pump	1.12	0.96	0.17	85.3	0.02	19.64	35.63	11.7	38.2	76.6	81.4
IP Pump	0.03	0.02	0.01	65.3	0.00	19.64	140.35	0.7	7.3	91.0	614.6
LP Pump	0.00	0.00	0.00	67.2	0.00	19.64	384.64	0.1	2.4	97.3	1858.4
De-aerator	0.56	0.53	0.03	95.4	0.00	24.79	40.05	2.3	26.1	92.0	61.6
Mixer 1	1.81	1.63	0.18	90.1	0.02	20.03	22.49	12.9	0.0	−	12.2
Mixer 2	0.63	0.58	0.04	92.9	0.01	20.10	24.35	3.2	0.0	−	21.2
Mixer 3	0.18	0.18	0.00	99.9	0.00	15.30	15.32	0.0	0.0	−	0.1
Condenser	12.43	−	7.53	−	1.70	21.15	−	946.4	85.7	8.3	−
Total (E_L=14 MW)	**730.58**	**411.40**	**305.15**	**56.3**	**41.77**	**9.15**	**20.53**	**10053.1**	**6459.9**	**39.1**	**124.4**

Table 4. Results of the exergetic and exergoeconomic analyses at the component-level for the plant with CLC.

Component, k	$\dot{E}_{F,k}$ [MW]	$\dot{E}_{P,k}$ [MW]	$\dot{E}_{D,k}$ [MW]	ε_k [%]	y_k [%]	$c_{F,k}$ [€/GJ]	$c_{P,k}$ [€/GJ]	$\dot{C}_{D,k}$ [€/h]	$\dot{Z}_k$ [€/h]	f_k [%]	r_k [%]
Compressor	281.78	268.57	13.21	95.3	1.81	18.32	20.16	871.4	903.7	50.9	10.0
Reactors	694.73	500.67	194.06	72.1	26.56	9.06	15.25	6,332.4	4,823.1	43.2	68.3
GT	540.99	521.34	19.65	96.4	2.69	17.13	18.32	1,211.7	1,032.8	46.0	7.0
CO_2/H_2O Expander	54.16	51.35	2.81	94.8	0.38	14.61	16.51	147.6	202.2	57.8	12.9
NG PH	6.32	1.12	5.20	17.7	0.71	14.61	84.49	273.8	7.4	2.6	478.1
Reheater	16.60	14.36	2.24	86.5	0.31	17.13	22.20	138.2	52.3	27.4	29.6
HPSH	22.43	20.43	1.99	91.1	0.27	17.13	21.06	123.0	102.5	45.5	23.0
HPEVAP	35.74	32.97	2.77	92.2	0.38	17.13	20.52	170.8	142.6	45.5	19.8
HPECON	24.25	20.58	3.67	84.9	0.50	17.13	22.54	226.1	57.5	20.3	31.6
IPSH	0.63	0.49	0.14	78.1	0.02	17.13	28.74	8.5	7.6	47.2	67.8
IPEVAP	11.97	11.00	0.97	91.9	0.13	17.13	21.77	60.0	92.5	60.7	27.1
IPECON	2.03	1.70	0.33	83.6	0.05	17.13	23.65	20.5	8.7	29.8	38.1
LPSH	1.12	0.81	0.32	71.9	0.04	17.13	31.82	19.4	13.1	40.3	85.8
LPEVAP	17.03	13.99	3.03	82.2	0.42	17.13	25.51	187.1	137.9	42.4	48.9
LPECON	11.28	7.52	3.76	66.6	0.51	17.13	32.90	232.0	74.0	24.2	92.1
SH II	0.48	0.42	0.05	88.6	0.01	14.61	20.04	2.9	3.7	56.2	37.1
EVAP II	2.03	1.93	0.10	95.2	0.01	14.61	17.36	5.1	10.8	67.7	18.8
ECON II	1.47	1.20	0.27	81.8	0.04	14.61	20.42	14.1	2.6	15.4	39.7
HPST	24.14	22.41	1.73	92.8	0.24	22.01	25.59	137.0	95.9	41.2	16.2
IPST	22.98	21.57	1.41	93.9	0.19	22.27	26.08	112.8	138.4	55.1	17.1
LPST	48.80	42.17	6.63	86.4	0.91	23.55	31.02	561.8	360.8	39.1	31.8
ST for CO_2 supply	20.73	15.66	5.06	75.6	0.69	22.29	34.96	406.3	146.0	26.4	56.9
Condensate Pump	0.04	0.03	0.01	78.6	0.00	20.37	72.99	0.7	5.4	89.0	258.3
HP Pump	0.98	0.84	0.14	85.3	0.02	20.37	34.75	10.6	28.0	72.6	70.6
IP Pump	0.05	0.04	0.02	70.1	0.00	20.37	95.73	1.1	8.2	87.9	369.9
LP Pump	0.00	0.00	0.00	66.5	0.00	20.37	348.84	0.1	1.9	96.8	1612.4
CO_2 Compressor 1	3.85	3.24	0.61	84.1	0.08	34.96	182.09	76.9	294.1	79.3	420.8
CO_2 Compressor 2	3.96	3.32	0.64	83.8	0.09	34.96	73.28	81.0	302.4	78.9	109.6
CO_2 Compressor 3	3.91	3.27	0.64	83.5	0.09	34.96	75.71	81.0	298.4	78.7	116.6
CO_2 Compressor 4	3.93	3.26	0.67	83.0	0.09	34.96	76.53	84.0	299.8	78.1	118.9
De-aerator	0.44	0.42	0.02	95.6	0.00	26.45	41.59	1.8	20.7	91.8	57.3
Mixer 1	0.85	0.79	0.06	92.5	0.01	22.27	24.80	5.1	0.0	0.0	11.4
Mixer 2	0.08	0.08	0.00	99.4	0.00	22.01	22.22	0.0	0.0	0.0	0.9
Mixer 3	1.99	1.97	0.02	99.1	0.00	17.13	17.37	1.1	0.0	0.0	1.4
Mixer 4	0.77	0.73	0.05	93.9	0.01	22.10	24.10	3.7	0.0	0.0	9.1
Mixer 5	0.12	0.12	0.00	100.0	0.00	23.55	23.55	0.0	0.0	0.0	0.0
Flue gas condenser	22.36	-	18.65	-	2.55	14.70	-	1,183.0	70.6	5.6	-
Cooler 1	0.77	-	0.62	-	0.08	33.04	-	91.8	8.7	8.7	-
Cooler 2	0.91	-	0.74	-	0.10	37.19	-	122.0	8.7	6.6	-
Cooler 3	0.85	-	0.70	-	0.10	40.80	-	124.4	7.9	5.9	-
Cooler 4	0.86	-	0.71	-	0.10	43.94	-	136.7	10.2	6.9	-
Condenser	11.52	-	6.99	-	0.96	23.25	-	964.5	66.2	6.4	-
Total (E_L=43 MW)	**730.73**	**374.82**	**312.89**	**51.29**	**42.82**	**9.15**	**25.46**	**10,308.1**	**10,422.8**	**50.3**	**178.2**

plant increases by about 30%. The cost of the avoided CO_2 in the case with MEA would then be 78.3 €/t, a price 47% higher, than that of the plant with CLC. The difference in cost is caused by the high energy consumption during regeneration and the lower CO_2 capture percentage (85%) assumed for the plant with MEA.

5. Conclusions

In this paper, a three-pressure-level combined cycle plant using chemical looping combustion for approximately 100% CO_2 capture, has been compared to a simple combined cycle plant without CO_2 capture and briefly to a plant with CO_2 capture through chemical absorption using monoethanolamine.

An exergetic analysis showed lower irreversibilities for the chemical looping reactors in comparison to the conventional combustion chamber of the reference case. Moreover, the process resulted in a decrease of the exergetic efficiency by about 5 percentage points with respect to the case without CO_2 capture; yet it achieves an

exergetic efficiency of 51%, almost 6 percentage points higher than that of the conventional approach – chemical absorption with monoethanolamine. Nevertheless, the investment cost of such a technology is high and the operation of such a plant results in a significant increase in the cost of electricity.

Chemical looping combustion with interconnected beds is a promising technology with simultaneous inherent capture of the produced carbon dioxide. Issues that need further consideration are the scaling-up of pilot units and an examination of the impact of high temperatures on materials and engineering components that would result in more reliable and safer operation. Given that a price must be paid to implement CO_2 capture from fossil fuel power plants, CLC seems to be a relatively economical alternative, in comparison to other proposed approaches, such as chemical absorption.

Acknowledgements

Nomenclature

c	cost per unit of exergy (€/GJ)
$\dot{C}$	cost rate associated with an exergy stream (€/h)
$\dot{E}$	exergy rate (MW)
f	exergoeconomic factor (%)
$\dot{m}$	mass flow rate (kg/s)
p	pressure (bar)
r	relative cost difference (%)
T	temperature (°C)
y	exergy destruction ratio (%)
$\dot{Z}$	cost rate associated with capital investment (€/h)

Greek symbols

ε	exergetic efficiency (%)

Subscripts

D	destruction (exergy)
F	fuel (exergy)
i, j	entering and exiting exergy streams
k	component
L	loss (exergy)
P	product (exergy)

Abbreviations

AR	Air Reactor
CC	Combustion Chamber
CEPCI	Chemical Engineering Plant Cost Index
CLC	Chemical Looping Combustion
COE	Cost of electricity
FCI	Fixed Capital Investment
FR	Fuel Reactor
GT	Gas Turbine
HP,IP,LP	High-pressure, Intermediate-pressure, Low-pressure
HRSG	Heat Recovery Steam Generator
MEA	Monoethanolamine
OC	Oxygen Carrier
ST	Steam turbine
TRR	Total Revenue Requirement

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