

Exergoeconomic and exergoenvironmental analyses of a combined cycle power plant with chemical looping technology

Fontina Petrakopoulou*, Alicia Boyano, Marlene Cabrera, George Tsatsaronis

Technische Universität Berlin, Institute for Energy Engineering, Marchstr. 18, D-10587 Berlin, Germany

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ABSTRACT

CO₂ capture and storage from energy conversion systems is one option for reducing power plant CO₂ emissions to the atmosphere and for limiting the impact of fossil-fuel use on climate change. Among existing technologies, chemical looping combustion (CLC), an oxy-fuel approach, appears to be one of the most promising techniques, providing straightforward CO₂ capture with low energy requirements.

This paper provides an evaluation of CLC technology from an economic and environmental perspective by comparing it with a reference plant, a combined cycle power plant that includes no CO₂ capture. Two exergy-based methods, the exergoeconomic and the exergoenvironmental analyses, are used to determine the economic and environmental impacts, respectively. The applied methods facilitate the iterative optimization of energy conversion systems and lead towards the improvement of the effectiveness of the overall plant while decreasing the cost and the environmental impact of the generated product. For the plant with CLC, a high increase in the cost of electricity is observed, while at the same time the environmental impact decreases.

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1. Introduction

Due to concerns about rising concentrations of greenhouse gases in the atmosphere, CO₂ capture from power plants and its permanent storage in suitable geological formations (CCS) has become an important mitigation option for climate change (Herzog, 2001). There are currently a number of proposed methods for capturing the CO₂ produced in power plants. However, the majority of these techniques are energy intensive, resulting in a significant decrease in the overall efficiency of a system and a substantial increase in the monetary cost associated with the generated products.

Considering these factors, we investigate the economic and ecological aspects of an oxy-fuel power plant operating with chemical looping combustion (CLC). Previous studies (Richter and Knoche, 1983; Hossain and de Lasa, 2008) show that CLC has the potential to become a relatively efficient and low cost technology. The process was first introduced by Lewis and Gilliland in 1954, in 1968

it was proposed by Knoche and Richter as an option for decreasing irreversibilities in combustion processes (1968), but later it was identified as having important advantages due to its nitrogen-free CO₂ production. This allows CO₂ separation with minimal thermodynamic losses and minimal contribution to NO_x emissions (Hossain and de Lasa, 2008; Brandvoll and Bolland, 2004).

In the CLC unit, the combustion products are kept separate from the air through the use of a metal oxide oxygen carrier (OC), and of two separate reactors for the oxygen separation and the fuel combustion. The OC is circulated between the two reactors, reacting with part of the air's oxygen in the air reactor and transferring it to the reaction chamber (fuel reactor). Complete combustion of the fuel in the fuel reactor produces CO₂, and water vapor, thus the CO₂ formed can be readily recovered by condensing the water vapor. This method eliminates the need for an additional, energy intensive CO₂ separation technique.

The goal of this paper is to highlight differences between two theoretical energy conversion systems, a plant with CLC and a reference plant (a three-pressure level combined cycle plant with one reheat stage) that includes no CO₂ capture, using exergoeconomic and exergoenvironmental analyses.

The exergoeconomic analysis (Tsatsaronis and Winhold, 1985; Bejan et al., 1996; Tsatsaronis, 1999; Tsatsaronis and Czesla, 2004) combines an exergetic analysis with an economic analysis to provide crucial information that is not obtainable through conventional thermodynamic analysis and simple economic evaluations. It is conducted at the component level, and specific costs associated

Abbreviations: CLC, chemical looping combustion; AR, air reactor; FR, fuel reactor; OC, oxygen carrier; HRSG, heat-recovery steam generator; ST, steam turbine; LP, low pressure; TRR, total revenue requirement; PEC, purchase equipment cost; FCI, fixed capital investment; COE, cost of electricity; CEPCI, chemical engineering plant cost index; GT, gas turbine; MEA, monoethanolamine.

* Corresponding author. Tel.: +49 30 314 22851; fax: +49 30 314 21683.

E-mail address: f.petrakopoulou@iet.tu-berlin.de (F. Petrakopoulou).

URL: <http://www.energietechnik.tu-berlin.de/> (F. Petrakopoulou).

Nomenclature

b	environmental impact per unit of exergy (mPts/GJ)
\dot{B}	environmental impact rate associated with exergy (mPts/h)
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 (%)
f_b	exergoenvironmental factor (%)
r	relative cost difference (%)
r_b	relative environmental impact difference (%)
T	temperature (°C)
y	exergy destruction ratio (%)
\dot{Y}	component-related environmental impact (mPts/h)
\dot{Z}	cost rate associated with capital investment (€/h)

Subscripts

D	exergy destruction
F	fuel (exergy)
k	component
L	loss
P	product (exergy)
i	exergy streams

Greek symbols

ε	exergetic efficiency (%)
λ	excess air fraction

with all exergy streams in the system are calculated. The results of an exergoeconomic analysis provide useful information for improving the cost effectiveness of the components and the overall system, by pinpointing the required changes in structure and parameter values (Bejan et al., 1996; Tsatsaronis, 1999; Tsatsaronis and Czesla, 2002, 2004).

Methodological development has evolved to include not only technical and economical data, but environmental factors as well. With this in mind, we also evaluate both considered plant configurations from an exergoenvironmental viewpoint. An exergoenvironmental analysis is based on an *exergetic analysis* and a *life cycle assessment*, following the principles of an exergoeconomic analysis. This analysis is also conducted at the component level of a system and identifies the environmental impact of each component included in the system, as well as options for reducing this impact (Meyer et al., 2009).

2. Methodology

2.1. Exergoeconomic analysis

A complete *exergoeconomic analysis* consists of (1) an *exergetic analysis*, (2) an *economic analysis*, and (3) an *exergy costing* that leads to the *exergoeconomic evaluation*. The exergetic analysis is suitable for identifying the sources of irreversibilities as exergy can be destroyed or lost, while energy is always conserved. The exergoeconomic analysis is conducted with a system of balance equations, stated at the component level, and a general equation for the overall system. Definitions of the exergy of the product and the exergy of the fuel, as well as important issues of the analyses are provided by Tsatsaronis and Winhold (1985), Bejan et al. (1996) and Lazzaretto and Tsatsaronis (2006).

The exergetic efficiencies of the k th component, ε_k , and of the overall system, ε_{tot} , are defined by the following equations

Table 1

Selected parameter and assumptions for the economic analysis.

Plant economic life (years)	20
Levelization period (years)	10
Average general inflation rate (%)	3
Average nominal escalation rate for natural gas (%)	4
Average real cost of money (%)	10
Date of commercial operation	2012
Average capacity factor (%)	85
Unit cost of natural gas (€/GJ-LHV)	7

(Tsatsaronis, 1999).

$$\varepsilon_k = \frac{\dot{E}_{P,k}}{\dot{E}_{F,k}} = 1 - \frac{\dot{E}_{D,k}}{\dot{E}_{F,k}} \quad (1)$$

$$\varepsilon_{tot} = \frac{\dot{E}_{P,tot}}{\dot{E}_{F,tot}} = 1 - \frac{\sum_{k=1}^{NC} \dot{E}_{D,k} + \dot{E}_{L,tot}}{\dot{E}_{F,tot}} \quad (2)$$

A useful variable calculated from the exergetic analysis is the exergy destruction ratio, $y_{D,k} = \dot{E}_{D,k}/\dot{E}_{F,tot}$ that shows which percentage of the total exergy of the fuel provided to the overall plant is destroyed in each of the components. In other words, it is a measure of the contribution of the exergy destruction within each component to the reduction of the overall exergetic efficiency.

After completion of the exergetic analysis, an economic analysis is conducted. Here we apply the *total revenue requirement* (TRR) method (Bejan et al., 1996). In this method, first the *fixed capital investment* (FCI) of the plant is estimated (Tsatsaronis and Winhold, 1984; EPRI report, 2000; Tsatsaronis and Czesla, 2002; Turton et al., 2002; Frammer, 2006). Costs are escalated to the reference year (2008) by means of the *chemical engineering plant cost index* (CEPCI) as published in the *Chemical Engineering Magazine* (CEM). Assumptions made concerning market conditions, plant operation and plant construction are summarized in Table 1. Using these assumptions, we calculate the levelized TRR with a levelization period of 10 years. Finally, the cost rate \dot{Z}_k associated with the k th component is used as input in the exergoeconomic analysis. The main part of the variable \dot{Z}_k stems from the investment costs. Thus, here this variable is referred to as the *investment-related cost*.

To perform the exergoeconomic evaluation, we use variables that help us rank the components, depending on the influence they have on the improvement of the overall plant. An important variable of the exergoeconomic evaluation is the relative cost difference (Tsatsaronis, 1999), which expresses the difference between the specific cost of the product, c_P , and the fuel, c_F :

$$r = \frac{c_{P,k} - c_{F,k}}{c_{F,k}} = \frac{1 - \varepsilon_k}{\varepsilon_k} + \frac{\dot{Z}_k}{\dot{C}_{D,k}}$$

The contribution of the cost rate \dot{Z}_k to the total sum of costs associated with capital and exergy destruction ($\dot{Z}_k + \dot{C}_D$) is expressed by the exergoeconomic factor $f_k = \dot{Z}_k/(\dot{Z}_k + \dot{C}_{D,k})$ (Tsatsaronis and Winhold, 1985). The relationship of the monetary impact of each component's exergy destruction and investment can then be examined.

Using the sum of the cost rates $\dot{Z}_k + \dot{C}_D$, the exergoeconomic variable f_k and the ratio r_k , more or less efficient design changes to improve the cost effectiveness of the plant, as a whole, are proposed. The objective is to reduce the cost associated with the product of the overall plant.

2.2. Exergoenvironmental analysis

The concepts of exergy and environmental impact are combined in the exergoenvironmental analysis. The method used to determine the environmental impact is *life cycle assessment* (LCA)

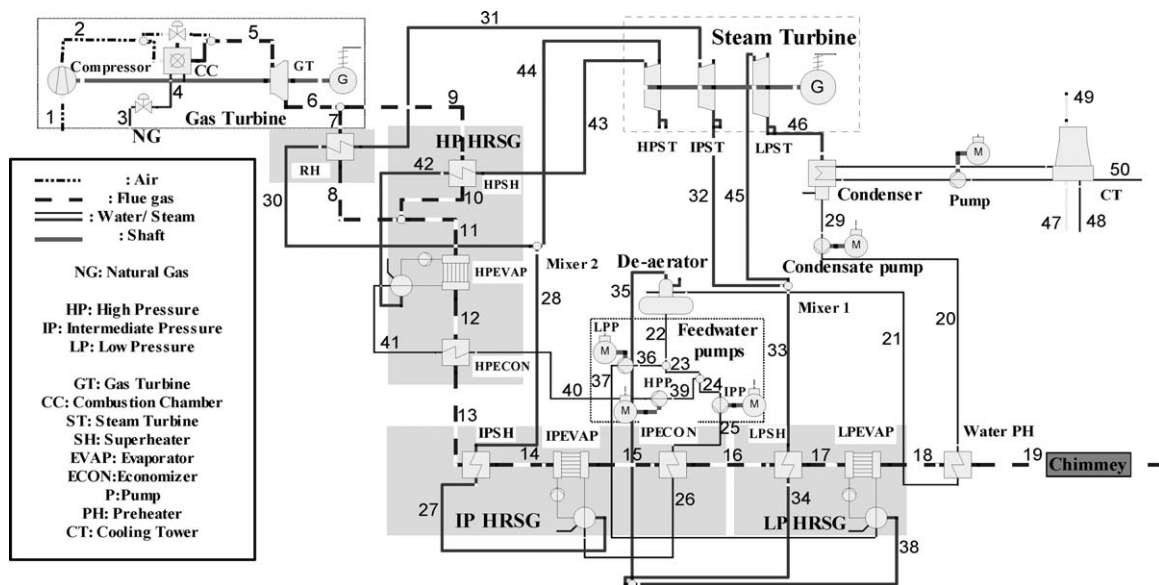


Fig. 1. Diagram of the reference plant.

(Meyer et al., 2009). Similar to the exergoeconomic analysis, the exergoenvironmental analysis is realized in three steps: (1) exergetic analysis, (2) LCA, and (3) exergoenvironmental costing: the assignment of environmental impact to the exergy streams of the system, which allows the calculation of the exergoenvironmental variables and the realization of the exergoenvironmental evaluation.

LCA is a technique used for assessing the environmental impact associated with a product over its lifetime. The quantification of the environmental impact caused by depletion and emissions of a natural resource can be carried out using different methodologies. In this work we used the impact analysis method Eco-indicator 99. The Eco-indicator 99 defines three categories of damage: human health, the ecosystem and the depletion of resources. LCA is carried out following the guidelines of international standard approaches (ISO 14004). In the LCA, the component variables, the consumption and the release of materials are identified and quantified (Fiaschi and Lombardi, 2002). In order to identify the raw material inlet

flows, the overall life cycle of each component of the plant has been considered and, hence, the phases of construction, operation and dismantling should be taken into account. However, according to previous studies (Tsatsaronis and Morosuk, 2008), the operation phase is included in the performance of the overall system and the dismantling phase can be regarded as a percentage of the construction phase. Accepting these assumptions, the construction phase is the main phase considered and examined for the purpose of this paper. For the quantification of the raw materials needed during the construction phase, it is necessary to approximate the size of the plant and collect information about the main materials, their production processes and weights, as well as the scrap output of all equipment assembled for the plant. To shift from the manufactured materials to the raw substances and emissions inventory, the commercially available software Package *SimaPro 7.1* is used. The environmental impact of the methane, as used in these plants, is assumed to be 180 mPts/kg (*SimaPro 7.1 manual, 2000*).

Table 2
Operating parameters^a.

<p>Ambient air 15 °C, 1.013 bar, 60% relative humidity Composition (mol%): N₂ (77.3), O₂ (20.73), CO₂ (0.03), H₂O (1.01), Ar (0.93)</p> <p>Fuel 14 kg/s, 15 °C, 50 bar, LHV = 50,015 kJ/kg Natural gas composition (mol%): CH₄ (100.0)</p> <p>Gas turbine system & CO₂/H₂O gas turbine Compressor: polytropic efficiency 94.0%, mechanical efficiency: 99%, pressure ratio: 16.8 Air turbine: polytropic efficiency: 91%, mechanical efficiency: 99%, cooling air: 11% of incoming air CO₂/H₂O turbine: polytropic efficiency: 91% Generators: electrical efficiency: 98.5%</p> <p>Steam cycle HRSG: 1 reheat stage, 3-pressure-levels: HP (124 bar), IP (22 bar), LP (4.1 bar) HRSG pressure drop: hot side: 30 mbar, cold side: 10% SHs, ECONs (HP, IP, LP): ΔT_{\min}: 20 °C EVAPs (HP, IP, LP): approach temperature: 6 °C, pinch point: 10 °C Live steam temperature: (ref. plant) 560 °C, (plant with CLC) 497 °C Steam turbine polytropic efficiency: HP (90%), IP (92%), LP (87%) Condenser operating pressure: 0.05 bar Pumps: efficiency: 62–86% (incl. motors and mechanical efficiency: 98%) Cooling water temperature: 21 °C</p>	<p>CLC unit (reactors) Adiabatic reactors, oxygen carrier: NiO/Ni (no losses) Inlet pressure: 17 bar Reactors pressure drop: 3% Fuel conversion: 98%</p> <p>CO₂ compression unit (4 intercooled stages) Compressors polytropic efficiency (4 stages): 80, 79, 78, and 77% CO₂ end pressure: 103 bar Cooling water: inlet/outlet temperature: 21 °C/31 °C CO₂ condenser exit temperature: 30 °C Coolers exit temperature: 40 °C Coolers pressure drop (4 stages): 0.15, 1.5, and 2.4 bar</p> <p>Overall plants Plant exergetic efficiency (ref. plant – without CO₂ capture): 56.3% Plant exergetic efficiency (plant with CLC): 51.3% Plant exergetic efficiency (plant with MEA): 45.8%</p>
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^a If not otherwise stated, the common components of the plants operate under the same conditions.

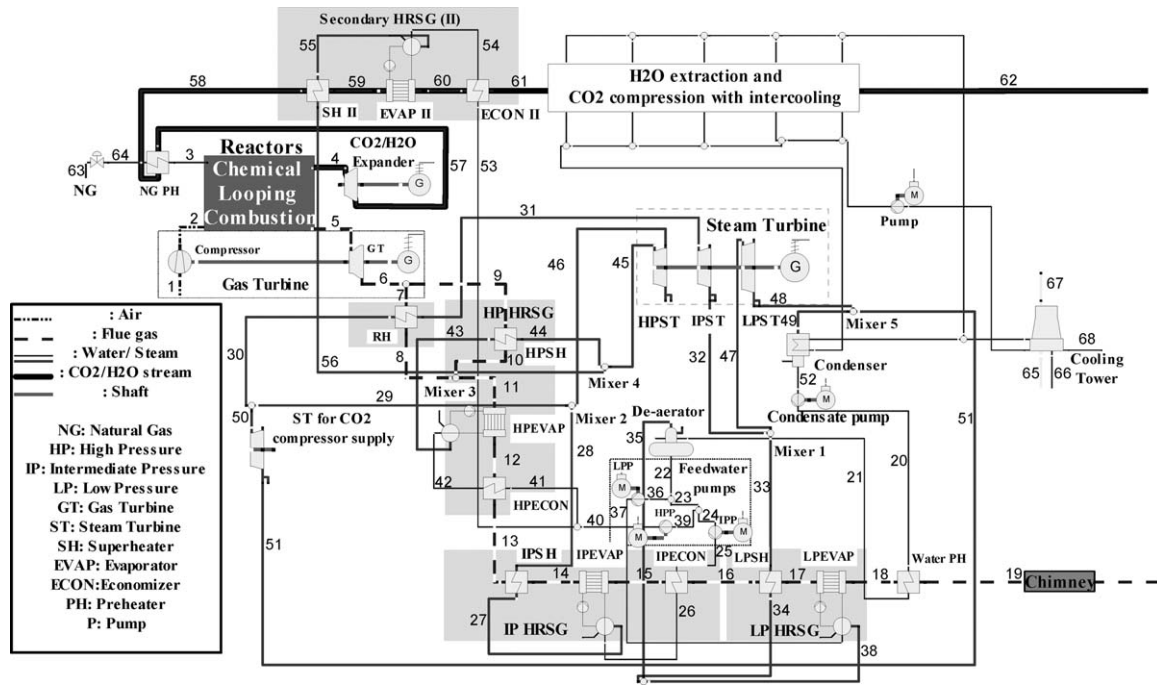


Fig. 2. Diagram of the plant with CLC.

The exergoenvironmental analysis does not only identify the components with the highest environmental impact, but also reveals the possibilities and trends for improvement, in order to decrease the environmental impact of the overall plant. These trends can be identified through the relative environmental impact difference, $r_{b,k}$, and the exergoenvironmental factor, $f_{b,k}$.

The environmental impact difference of component k , $r_{b,k} = (b_{F,k} - b_{P,k})/b_{F,k}$, depends on the environmental impact of its exergy destruction, $\hat{B}_{D,k}$, and the component-related environmental impact, \hat{Y}_k , and is an indicator of the reduction potential of the environmental impact associated with the component.

At the component level, the contribution of the component-related environmental impact, \hat{Y}_k , to the total environmental impact, $\hat{Y}_k + \hat{B}_{D,k}$, is expressed by the exergoenvironmental factor, $f_{b,k} = \hat{Y}_k / (\hat{Y}_k + \hat{B}_{D,k})$.

When the value of $f_{b,k}$ is high, the component-related environmental impact, \hat{Y}_k , is dominant, whereas when the value of $f_{b,k}$ is low, exergy destruction is the dominant source of the environmental impact. Thus, the higher the exergoenvironmental factor, the higher the influence of the component-related environmental impact to the overall performance of the plant from the environmental perspective.

After definition and calculation of the above variables, each component's total environmental impact is evaluated and design changes are formulated, in order to decrease the environmental impact of the components and the overall process. With a similar objective to the exergoeconomic analysis, the exergoenvironmental analysis indicates options for reducing the environmental impact associated with the product of the overall plant.

3. Plant configuration description

3.1. The reference plant

The reference plant (Fig. 1) is a plant used as the basis for the simulation and evaluation of the oxy-fuel plant with CLC. It is a combined cycle with a three-pressure level *heat-recovery steam generator* (HRSG) and one reheat stage that includes no CO₂ capture. The gas turbine system of the plant incorporates a cooling

mechanism that uses 11% of the incoming air to ensure longer safe operation of the turbine. Computational assumptions made for the simulation of the plant are provided in Table 2. The plant has only one product – electricity – and works with natural gas (assumed to be pure methane). The CO₂ emission rate of the plant is 339 g/kWh.

3.2. The plant with chemical looping combustion (CLC)

The plant with CLC (Fig. 2) is a combined cycle including a primary HRSG with three-pressure levels and a secondary single-pressure-level HRSG to remove heat from the combustion products. It performs with approximately 100% CO₂ capture and assumptions considered for the simulation of the plant, are given in Table 2. Due to material limitations, the temperature of the gases exiting the CLC unit is not higher than 1200 °C. No blade cooling is considered in the CLC system.

3.2.1. The CLC unit

The CLC unit (Fig. 3) consists of two interconnected, fluidized bed reactors: an air and a fuel reactor that take the place of the com-

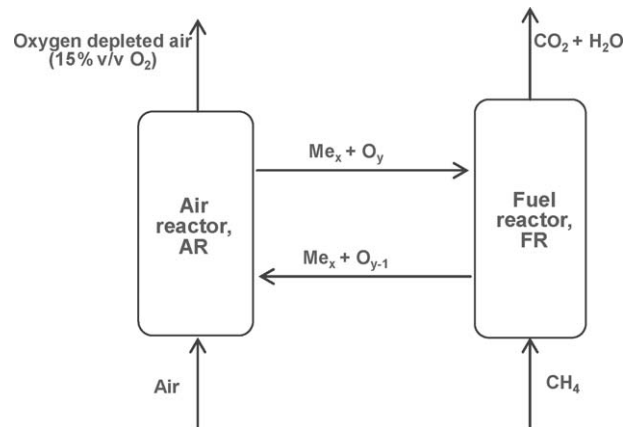


Fig. 3. Configuration of chemical looping combustion.

bustion chamber in a conventional gas turbine system. The reactors are designed as fluidized beds with the advantages of good contact between gas and solid materials and temperature homogeneity. Various arrangements of these reactors have been proposed in recent years (Brandvoll and Bolland, 2004; Naqvi et al., 2005; Abad et al., 2006; Naqvi and Bolland, 2007). In order to keep the combustion products separate from the air, a solid metal oxide is used as the oxygen carrier (OC). This OC is continuously circulated between the reactors, transferring part of the air's oxygen from the air to the fuel reactor.

In more detail, the atmospheric air is introduced into the air (or oxidizing) reactor, AR, where the metal oxide is oxidized. The metal oxide then exits the AR and is fed to the fuel reactor (FR), where the transported oxygen reacts with the fuel to produce CO₂ and H₂O. Simultaneously, the metal oxide is reduced back to its initial state and led back to the AR, continuing its loop between the two reactors. No direct contact between the air and the fuel takes place, and the carbon dioxide produced can be readily recovered after water condensation without costly energy requirements (Mattison and Lyngfelt, 2001).

The net reaction in the CLC unit and the heat generation are equal to that of the fuel combustion. The oxidation is an exothermic reaction, whereas the reduction can be either endothermic or slightly exothermic, depending on the fuel and metal oxide combination. Many different metals have been investigated for oxygen carriers, mainly based on nickel, iron or manganese, as presented by Lyngfelt et al. (2001) in a detailed review of the status of the development of oxygen carriers. According to literature (Bolh ar-Nordenkamp et al., 2008; Johansson et al., 2006; Kolbitsch et al., 2008; Lyngfelt and Thunman, 2005), nickel seems to be the most suitable carrier when CH₄ is used, due to its higher reactivity and greater durability; consequently, Ni is assumed in this work. Due to limitations of the simulation software, no losses of the oxygen carrier were assumed. Therefore, the carrier is considered completely oxidized when leaving the AR and completely reduced when leaving the FR.

In this study, CLC is simulated as a black box and replaces the combustion chamber of a conventional gas turbine system (Naqvi et al., 2005). The two streams entering the CLC unit (Fig. 2) are the compressed air (stream 2) and the preheated methane (stream 3) and the two streams exiting the unit are the combustion products (stream 4), consisting of CO₂ and water vapor, and the oxygen depleted air (stream 5) with 15% (v/v) O₂. The latter is expanded in the expander of the main gas turbine system and it is then led to the primary HRSG of the plant. A 98% conversion of methane in the FR was assumed, along with an air ratio of 2.9 with respect to the stoichiometric ratio and temperatures of 1200 and 930 °C in the AR and FR, respectively (Naqvi and Bolland, 2007; Lewis and Gilliland, 1954; Abad et al., 2007).

To calculate the costs of the reactors included in the CLC unit, we use the estimates reported in Petrakopoulou et al. (2009). Data from the publications of Klara (2007), Wolf et al. (2005) and Lyngfelt and Thunman (2005) were used to approximate the necessary input parameters. The cost of the metal oxide was not taken into account. The total FCI of the CLC unit was then found to be €128 million, with an adjustment to the year 2008 according to the CEPSI index.

4. Results and discussion

4.1. Exergetic and exergoeconomic evaluation

The results of the analyses for selected components of the reference plant and the plant with CLC are presented in Tables 3 and 4, respectively. The main exergy destruction in both plants occurs within the gas turbine (GT) system, essentially due to the combustion chambers. Nevertheless, the use of CLC decreases the exergy destruction ratio of the reactors ($y_{D,Reactors}$), by 12%, when com-

Table 3 Results of the exergetic, exergoeconomic and exergoenvironmental analyses for selected components of the reference plant.

Component, k	Reference plant without CO ₂ capture																		
	$\dot{E}_{F,k}$ (MW)	$\dot{E}_{P,k}$ (MW)	$\dot{E}_{D,k}$ (MW)	ϵ_k (%)	$y_{D,k}$ (%)	$c_{F,k}$ (€/GJ)	$c_{P,k}$ (€/GJ)	$\dot{C}_{D,k}$ (€/h)	\dot{Z}_k (€/h)	$\dot{C}_{D,k} + \dot{Z}_k$ (€/h)	f_k (%)	r_k (%)	$b_{F,k}$ (mPts/GJ)	$b_{P,k}$ (mPts/GJ)	$\dot{B}_{D,k}$ (mPts/h)	\dot{Y}_k (mPts/h)	$\dot{B}_{D,k} + \dot{Y}_k$ (mPts/h)	$f_{b,k}$ (%)	$r_{b,k}$ (%)
Compressor	242.68	231.30	11.38	95.3	1.56	16.7	19.0	683	1,297	1,980	65.5	14	5579	5853	228,496	236	228,731	0.10	5
CC	729.62	508.76	220.87	69.7	30.23	9.2	13.6	7,276	926	8,203	11.3	49	3599	5162	2,861,563	381	2,861,944	0.01	43
GT	551.15	530.67	20.47	96.3	2.80	15.3	16.7	1,128	1,482	2,610	56.8	9	5371	5579	395,831	1,126	396,957	0.28	4
Reheater	26.47	23.89	2.58	90.3	0.35	15.3	19.1	142	105	247	42.6	25	5371	6247	49,795	835	50,630	1.65	16
HPSH	35.07	31.72	3.35	90.5	0.46	15.3	19.2	184	149	334	44.8	25	5371	6230	64,738	1,237	65,975	1.88	16
HPEVAP	43.64	39.91	3.73	91.5	0.51	15.3	18.8	205	184	389	47.2	23	5371	6123	72,063	139	72,202	0.19	14
HPECON	28.92	24.91	4.00	86.2	0.55	15.3	20.2	220	89	309	28.7	32	5371	6664	77,384	96	77,481	0.12	24
IPSH	0.18	0.12	0.06	69.0	0.01	15.3	34.6	3	4	7	55.2	126	5371	9022	1,071	18	1,088	1.65	68
IPEVAP	6.10	5.67	0.43	92.9	0.06	15.3	20.3	24	65	89	73.2	33	5371	5986	8,351	4	8,392	0.49	11
IPECON	1.06	0.87	0.19	82.5	0.03	15.3	22.1	10	5	15	33.5	44	5371	7081	3,596	4	3,600	0.10	32
LP SH	1.43	1.04	0.38	73.3	0.05	15.3	29.0	21	18	39	46.6	89	5371	8334	7,366	110	7,477	1.48	55
LPEVAP	19.03	15.48	3.55	81.4	0.49	15.3	23.9	195	173	368	46.9	56	5371	7216	68,581	128	68,709	0.19	34
LPECON	11.49	7.71	3.78	67.1	0.52	15.3	30.5	209	93	301	30.8	99	5371	9322	73,179	100	73,279	0.14	74
HPST	31.29	29.18	2.11	93.2	0.29	20.1	23.8	153	166	318	52.0	18	6473	7137	49,230	276	49,507	0.56	10
IPST	37.39	35.21	2.18	94.2	0.30	20.0	24.2	157	300	457	65.6	21	6427	6993	50,483	317	50,800	0.62	9
LPST	70.99	61.35	9.64	86.4	1.32	21.2	29.0	734	696	1,431	48.7	37	6671	8142	231,559	493	232,052	0.21	22
Total ($E_L = 14.0$ MW)	730.58	411.40	305.15	56.3	41.77	9.2	20.5	10,053	6,460	16,513	39.1	124	3599	6398	39,536,056	49,422	39,585,477	0.12	78

Table 4
Results of the exergetic, exergoeconomic and exergoenvironmental analyses for selected components of the plant with CLC.

Component, <i>k</i>	Plant with chemical looping combustion																		
	$\dot{E}_{F,k}$ (MW)	$\dot{E}_{P,k}$ (MW)	$\dot{E}_{D,k}$ (MW)	ε_k (%)	$y_{D,k}$ (%)	$c_{F,k}$ (€/GJ)	$c_{P,k}$ (€/GJ)	$\hat{C}_{D,k}$ (€/h)	\dot{Z}_k (€/h)	$\hat{C}_{D,k} + \dot{Z}_k$ (€/h)	f_k (%)	r_k (%)	$b_{F,k}$ (mPts/GJ)	$b_{P,k}$ (mPts/GJ)	$\dot{B}_{D,k}$ (mPts/h)	\dot{Y}_k (mPts/h)	$\dot{B}_{D,k} + \dot{Y}_k$ (mPts/h)	$f_{b,k}$ (%)	$r_{b,k}$ (%)
Compressor	281.78	268.57	13.21	95.3	1.81	18.3	20.2	871	904	1,775	50.9	10	5729	6,011	272,483	190	272,672	0.07	5
Reactors	694.73	500.67	194.06	72.1	26.56	9.1	15.3	6,332	4,823	11,155	43.2	68	3617	5,021	2,527,096	2,537	2,529,633	0.10	39
GT	540.99	521.34	19.65	96.4	2.69	17.1	18.3	1,212	1,033	2,245	46.0	7	5521	5,729	390,619	906	391,525	0.23	4
CO ₂ /H ₂ O expander	54.16	51.35	2.81	94.8	0.38	14.6	16.5	148	202	350	57.8	13	3878	4,091	39,175	192	39,367	0.49	5
NG PH	6.32	1.12	5.20	17.7	0.71	14.6	84.5	274	7	281	2.6	478	3878	21,936	72,645	3	72,648	0.00	466
Reheater	16.60	14.36	2.24	86.5	0.31	17.1	22.2	138	52	190	27.4	30	5521	6,794	44,564	477	45,040	1.06	23
HPSH	22.43	20.43	1.99	91.1	0.27	17.1	21.1	123	103	226	45.5	23	5521	6,323	39,647	934	40,581	2.30	14
HPEVAP	35.74	32.97	2.77	92.2	0.38	17.1	20.5	171	143	313	45.5	20	5522	6,203	55,056	123	55,180	0.22	12
HPECON	24.25	20.58	3.67	84.9	0.50	17.1	22.5	226	57	284	20.3	32	5522	6,965	72,893	69	72,962	0.09	26
IPEVAP	11.97	11.00	0.97	91.9	0.13	17.1	21.8	60	93	153	60.7	27	5522	6,240	19,341	76	19,417	0.39	13
LPEVAP	17.03	13.99	3.03	82.2	0.42	17.1	23.7	21	9	29	42.4	49	5522	7,107	6,612	8	6,620	0.13	29
LPECON	11.28	7.52	3.76	66.6	0.51	17.1	32.9	232	74	306	24.2	92	5522	9,576	74,803	93	74,897	0.12	73
EVAP II	2.03	1.93	0.10	95.2	0.01	14.6	17.4	5	11	16	67.7	19	3878	4,206	1,363	6	1,370	0.47	8
ECON II	1.47	1.20	0.27	81.8	0.04	14.6	20.4	14	3	17	15.4	40	3878	5,309	3,729	2	3,731	0.05	37
HPST	24.14	22.41	1.73	92.8	0.24	22.0	25.6	137	96	233	41.2	16	6568	7,276	40,876	229	41,105	0.56	11
IPST	22.98	21.57	1.41	93.9	0.19	22.3	26.1	113	138	251	55.1	17	6635	7,239	33,611	234	33,845	0.69	9
LPST	48.80	42.17	6.63	86.4	0.91	23.5	31.0	562	361	923	39.1	32	6912	8,404	164,905	375	165,280	0.23	22
ST for CO ₂ supply	20.73	15.66	5.06	75.6	0.69	22.3	35.0	406	146	552	26.4	57	6590	9,555	120,135	185	120,320	0.15	45
CO ₂ compressor 1	3.85	3.24	0.61	84.1	0.08	35.0	182.1	77	294	371	79.3	421	9555	39,502	21,023	119	21,142	0.56	313
CO ₂ compressor 2	3.96	3.32	0.64	83.8	0.09	35.0	73.3	81	302	383	78.9	110	9555	12,545	22,137	49	22,185	0.22	31
CO ₂ compressor 3	3.91	3.27	0.64	83.5	0.09	35.0	75.7	81	298	379	78.7	117	9555	12,972	22,128	24	22,152	0.11	36
CO ₂ compressor 4	3.93	3.26	0.67	83.0	0.09	35.0	76.5	84	300	384	78.1	119	9555	13,077	22,965	35	23,000	0.15	37
Condenser flue gas	22.36	–	18.65	–	2.55	14.7	–	1,183	71	1,254	6.4	–	–	–	313,918	89	314,006	0.03	–
Total ($E_L = 43.0$ MW)	730.73	374.82	312.89	51.3	42.82	9.2	26.4	10,715	11,790	22,504	50.3	178	3900	1,000	4,053,861	51,937	4,105,798	1.26	95

pared to that of the combustion chamber ($y_{D,CC}$) of the reference plant.

The two plants are provided with the same amount of fuel, resulting in the same exergy of fuel (E_F). Although the exergy destruction (E_D) in the plant with CLC is lower by about 8 MW than that of the reference plant, the exergy of the product (E_P) of the latter is higher. With the same E_F and lower E_D , the lower E_P for the plant with CLC is explained by its larger exergy loss (E_L). 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. Moreover, a comparison of the total y_D of the common components of the plants shows that the plant with CLC has a lower y_D when compared to the reference plant, mainly due to the more efficient combustion process. However, the components added for CO_2 capture, as well as the secondary HRSG increase the overall y_D of the plant by 14%.

From an economic perspective, 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. From the latter, almost 13% is due to the added equipment for CO_2 compression, i.e., the intercooled CO_2 compression unit and the steam turbine used to drive it. Additionally, 35% of the total FCI of the plant stems from the reactors of the CLC unit. Thus, it is clear that there is a large difference between the total FCI per produced kW in the reference plant (522€/kW) and in the plant with CLC (980€/kW).

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; therefore, CO_2 capture causes an increase in the cost of electricity of about 24%.

To examine whether CLC is an economically viable solution for CO_2 capture, we calculated the cost of the CO_2 avoided, taking into account the equation provided in Rubin and Rao (2002). In this paper, this cost relates only to the capture of the CO_2 and it does not include transportation or storage costs. We compared this cost (53.1€/metric ton) to the cost associated with the conventional 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 minimum 10% increase in the capital cost for the plant working with chemical absorption, with respect to the reference plant, the COE of the MEA plant increased by about 30%. Its cost of avoided CO_2 was found to be 78.3€/metric ton, a price 47% higher than that of the plant with CLC. This increase in cost is caused mainly by the high energy supply to the regeneration process and by the lower CO_2 capture percentage (85%) in the plant working with chemical absorption.

4.2. Exergoenvironmental evaluation

The main results of the exergoenvironmental analysis at the component level for the reference plant and the plant with CLC are summarized in Tables 3 and 4, respectively. In the reference plant,

the highest total environmental impact ($\dot{B}_{D,k} + \dot{Y}_k$) is calculated for the GT system (essentially due to the combustion chamber) and for the high-pressure HRSG. On the other hand, in the plant with CLC, the GT system has the highest environmental impact, followed by the high-pressure HRSG and the equipment used for the compression and cooling of the separated CO_2 stream.

To compare the overall performance of the two configurations, we calculate the environmental impact per kWh produced in both plants. This impact is 32 mPts/kWh for the reference plant and 23 mPts/kWh for the plant with CLC. The difference of about 9 mPts/kWh between the two configurations is considered quite significant, taking into account the values of the estimated environmental impact of the low voltage electricity produced in Europe with an average of 26 mPts/kWh (SimaPro 7.1 manual, 2000). However, it should be mentioned that the environmental impact of both plants includes the production of electricity, as well as the environmental impact due to CO_2 and NO_x emissions, while the environmental impact of the plant with CLC does not include considerations for the transportation and the storage of the CO_2 . Using an average estimate provided by Khoo and Tan (2006) of 4.9 mPts/kg of CO_2 for transportation and storage, the environmental impact of the plant with CLC increases to 25 mPts/kWh. Overall, using CLC technology for CO_2 capture reduces the overall environmental impact of the electricity production by about 22%.

The exergoenvironmental variable, $f_{b,k}$, shows that the total environmental impact of the reference plant can be decreased mainly by increasing the exergetic efficiency of all components, as all of them show low $f_{b,k}$ values. The calculated exergoenvironmental factors do not differ significantly between the two plants. A reduction of the overall environmental impact could be achieved by increasing the exergetic efficiency of the majority of the components, but mainly that of the GT system. In this study, it happens that the results of the exergoenvironmental analysis do not differ much from those of the exergetic analysis. However, when newly introduced technologies (e.g. membranes) are examined, the exergoenvironmental factor can achieve higher values, provide interesting information, and lead to conclusions that diverge from those of the exergetic analysis.

5. Conclusions

In this paper, an oxy-fuel power plant with chemical looping combustion for approximately 100% CO_2 capture has been compared to a reference power plant of similar configuration without CO_2 capture. Oxy-fuel combustion can be considered as an effective means for producing cleaner electricity in the near future, mainly due to the fact that CO_2 is produced in a nitrogen-free environment. This advantage, as well as lower irreversibilities and lower NO_x emissions in the plant with CLC, make it an appealing approach for CO_2 capture from power plants.

As initially intended by the developers of this technology, the combustion process in the CLC unit (reactors) showed lower irreversibilities than that of the conventional combustion chamber of the reference case, due to the nitrogen-free combustion and the preheating of the fuel. Additionally, the overall process results in a relatively low decrease of about 5 percentage points in the exergetic efficiency, with respect to the reference plant.

As far as the economic analysis is considered, the fixed capital investment of the oxy-fuel process was estimated to be about 71% higher than that of the reference plant. This large difference in the costs results from the large size, and therefore, high cost of the reactors, as well as the equipment added for CO_2 capture. From this increase, almost 35% is directly linked to the reactors of the CLC unit and 13% to the newly added equipment used for CO_2 capture.

The exergoeconomic analysis shows an increase in the cost of electricity of about 24% when CO_2 capture is considered, where the

main source of additional cost is the combustion process in the GT system. The increase in the investment costs of the CLC reactors, with respect to the cost of the conventional CC in the reference plant, is not offset completely by the savings in the combustion process. Thus, the CLC reactor is the dominant cost source when compared to the rest of the components.

The environmental impact of the electricity produced by the plant with CLC is lower than that of the reference case, showing a benefit of CLC technology for producing environmentally friendlier electricity. The highest environmental impact of the electricity production was caused by the exergy destruction in the system components during the operation phase; therefore, in order to reduce the overall environmental impact of electricity, an increase in the exergetic efficiency of almost all important components is recommended.

The exergoeconomic and exergoenvironmental analyses can provide important information about how to improve both the structure and operating conditions of the plants, in order to improve the economic and environmental effectiveness. Further details, however, referring to the improvement potential and component interactions will be provided in upcoming publications, in which results of advanced exergy-based analyses of the same plants will be reported.

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