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Integrated Gasifier Combined Cycle Plant with Integrated CO_2 – H_2S Removal: Performance Analysis, Life Cycle Assessment and Exergetic Life Cycle Assessment

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Abstract

The aim of this study is to assess the possibility of reducing CO_2 emissions from an Integrated Gasifier Combined Cycle plant, accounting for the entire life cycle.

With respect to the basic scheme of a conventional Integrated Gasification Combined Cycle - which already includes H_2S removal - a shift reaction section followed by CO_2 chemical absorption have been added, before the syngas combustion, in order to obtain a clean and H_2-rich synthesis gas.

The proposed plant reaches 38.8 % efficiency vs. 46.4 % efficiency of the corresponding IGCC without CO_2 removal. The main reason for the efficiency reduction is the complexity due to the addition of shift and CO_2 removal sections and, hence, the related losses in the complex heat recovery process.

From LCA results, a substantial decrease in the carbon dioxide total production, due to the applied removal technology, is highlighted. In fact, the CO_2 equivalent production in the life cycle, per energy unit, is one order of magnitude lower with respect to the parent IGCC.

Moreover, the major contribution to CO_2 production comes from operation/maintenance phase; likely, in the case of Exergetic Life Cycle Assessment: the major exergy destruction - in term of MJ of destroyed exergy per functional unit - comes from operation/maintenance, which is two orders of magnitude greater than the contributions of the other phases.

The whole cycle has been simulated by means of Aspen Plus 10.1-0.

Key words: IGCC (integrated gasification combined cycle), CO2 removal, LCA (life cycle assessment), ELCA (exergetic life cycle assessment), ammine, GHGs (greenhouse gases)

1. Introduction

The production of clean coal syngas and its use in a combined cycle has been simulated. The goal is to obtain a syngas containing a large amount of hydrogen (loosing as little as possible of the original energy content).

The basic scheme is an Integrated Gasifier Combined Cycle (IGCC), reflecting the present technology of coal gasification systems, including particulate and H_2S removal systems, coupled to a gas turbine combined cycle. With respect to this scheme two basic units have been added: a shift reaction section and a CO_2 chemical absorption process, both working on the syngas stream before its combustion. For this reason, this cycle scheme will be referred to in the following as IGCC with $DeCO_2$. A schematic overall flow diagram is showed in *Figure 1*

In the gasifier, the coal is converted with the gasifying mediums oxygen and steam to a CO and H_2 rich gas. This raw syngas contains several pollutants, especially carbonyl-sulphide (COS), hydrogen-sulphide (H_2S), carbon monoxide (CO) and carbon-dioxide (CO_2) which need to be removed.

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First the COS, if present, is converted into H_2S and carbon dioxide by means of COS-hydrolysis water gas reaction. In this way the COS sulphur content is converted in H_2S which is more easily removable.

Then H_2S removal takes place, followed by a catalytic shift reaction. Here, CO reacts with water to produce CO_2 and H_2. This stage is required to obtain a hydrogen rich syngas from the carbon-monoxide-rich raw syngas; moreover, the obtained stream has a relevant CO_2content, which can be conveniently removed through a subsequent chemical absorption process. The clean hydrogen rich syngas leaves this section in pressurised conditions and is finally delivered to the combined cycle.

The choice of two separated systems for CO_2 and H_2S removal was imposed by the necessity of obtaining two separate streams, almost pure in CO_2 and H_2S respectively. In fact, while CO_2 is sent to a disposal treatment, H_2S is typically further processed to obtain valuable by products by the Claus process.

For the removal of CO_2 and H_2S, chemical absorption with aqueous amines solutions has been selected; this is a well assessed process for H_2S removal, and also the most developed CO_2 removal technique.

All the processes above mentioned have been simulated by means of Aspen Plus 10.1-0 (Aspen Plus ^{TM} , 1996).

2. The Gasification Section

In the simulation, an entrained flow gasifier has been modelled at a constant 20 bar pressure, operating on a 31 kg/s coal mass flow.

As raw material, a reference coal (Manfrida et al., 1990) whose ultimate analysis is reported in TABLE I (ash and moisture free basis), has been considered.

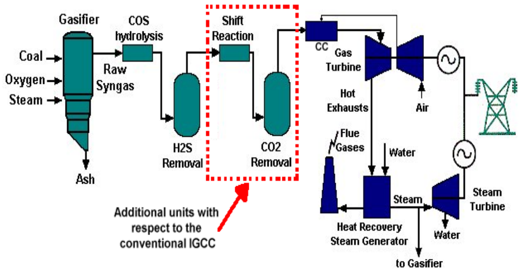


Figure 1. Schematic of IGCC + $DeCO_2$

The ash content is unknown for the used coal. Therefore, it has been supposed to be 10%, according to the ash contents of similar coals (Smith et al., 1994). The chlorine content of this coal is also unknown and it has been set to zero, since it is low for the major part of coals (Swaine, 1990). The moisture content has been set to 10%.

TABLE I. REFERENCE COAL COMPOSITION

Component	Weight %
Carbon	77.2
Oxygen	11.2
Nitrogen	1.4
Hydrogen	5.9
Sulphur	4.3

The gasification media are pure oxygen at ambient temperature of 25 °C and steam at 250 °C. The mass ratio between injected water and coal is set at 0.074 (Kloster, 1999). The gasifier reaction temperature is 1400 °C, so that the ash is forming a molten slag (Merrick, 1984). An oxygen/coal mass ratio of 0.85 (i.e. Equivalence Ratio ER = 0.32) has been assumed (Kloster, 1999). The exiting syngas is first cooled down to 900 °C - below the ash melting point - through a mixing cooler, where it is mixed with an appropriate flow rate of the same gas (about 42 kg/s), taken downstream the cyclone and repressurised (*Figure 2*).

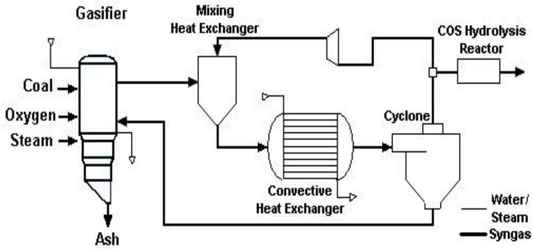


Figure 2. Gasification section scheme

An additional convective cooler lowers the temperature down to 140 °C, in order to safely supply the gas to the cyclone (Weber and Brocke, 1973), which removes all the ash together with 95% of the non-converted solid carbon. The removed stream is recirculated to the gasifier (*Figure 2*).

The raw syngas mass flow exiting the gasification section is about 57 kg/s and its composition – resulting from Aspen Plus simulation - is reported in TABLE II.

TABLE II. RAW SYNGAS FEATURES

Temperature [°C]	140	
Pressure [bar]	19.7	
Mass Flow [kg/s]	56.83	
Composition	Mass Frac.	Mole Frac.
H_2O	0.101	0.12
CO_2	0.148	0.072
N_2	0.006	0.005
COS	0.002	0.001
CO	0.699	0.534
H_2S	0.019	0.012
H_2	0.024	0.257

3. COS Hydrolysis Section

Organic sulphur compounds are generally much less chemically reactive than hydrogen sulphide and, therefore, they are very little removed in conventional H_2S removal processes. Hence, catalytic conversion is used to eliminate them. In particular, carbonyl sulphide is converted to hydrogen sulphide by hydrolysis in presence of a correct catalyst. Chromia-alumina and copper-chromia-alumina catalysts are used for selective hydrolysis of carbonyl sulphide in the presence of large amounts of carbon monoxide and are not affected by the presence of H_2S in the inlet gas. Chromia-alumina catalysts assure a complete conversion at temperatures ranging from about 300 to 450 °C, while copper-chromia-alumina catalysts have a much higher activity permitting operation at lower temperatures (Kohl and Riesenfeld, 1985).

The main raw syngas stream, after diverting the necessary fraction for the mixing cooler, is fed to the COS hydrolysis reactor.

According to Kloster (1999), in presence of the catalyst (copper-chromia-alumina), at ca. 140 °C, only the following reaction occurs, with a COS conversion of 99.9%:

$$COS + H_2O \rightarrow H_2S + CO_2$$

The operating conditions in the gasification section are reported in TABLE III. The composition of the syngas leaving the COS hydrolysis section is reported in TABLE IV.

TABLE III. GASIFICATION SECTION OPERATING CONDITIONS

Gasification temperature [°C]	1400
Gasification pressure [bar]	20
Coal flow rate [kg/s]	31
Oxygen/Coal mass ratio	0.85
Steam/Coal mass ratio	0.074
Gasification steam temperature [°C]	250
Mixing HE outlet temperature [°C]	900
Con. HE outlet temperature [°C]	140
Gasifier HP steam production [kg/s]	7
Con. HE HP steam production [kg/s]	51
Con. HE LP steam production [kg/s]	1.07

4. H_2S Removal Section

For combined cycle power generation applications, hydrogen sulphide has to be removed in order to comply with sulphur emission limits. In this case, a sulphur removal efficiency of 90% is acceptable (Swaine, 1990), in fact, the H_2S content in the final stack stream is of the order of part per million (PPM).

The removal of H_2S from the syngas stream takes place in an absorber column using a solution of aqueous amines as absorbing medium. In these solutions H_2S and CO_2 react in an acid base buffer mechanism with alkanolamines. Aqueous amines solutions have the capability to react with both H_2S and CO_2. Actually it is necessary to remove them separately, in order to have the possibility of distinct further treatment for the two pollutants. For this reason, the tertiary amine methyldiethanolamine (MDEA) has been chosen, which has capability of reacting selectively with H_2S, differently from primary and secondary amines monoethanolamine (MEA) and diethanolamine (DEA) (Merrick, 1984).

TABLE IV. SYNGAS LEAVING THE COS HYDROLYSIS SECTION

Temperature [°C]	140.8	
Pressure [bar]	19.1	
Mass Flow [kg/s]	56.83	
Composition	Mass Frac.	Mole Frac.
H_2O	0.1	0.119
CO_2	0.15	0.073
N_2	0.006	0.005
CO	0.699	0.534
H_2S	0.021	0.013
H_2	0.024	0.257

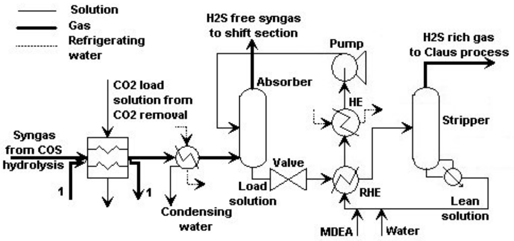


Figure 3. H_2S removal section scheme.

The H_2S load solution needs to be thermally regenerated in a second column, working with an externally heated reboiler, before being recirculated back to the absorber. The absorber is working necessarily at pressurised conditions (about 18 bar), in order to preserve the syngas pressure level obtained in the gasifier. Since the solubility of H_2S in the absorbing medium decreases with decreasing pressure (Austgen et al., 1989), then it is useful to depressurise the H_2S rich solution down to ambient pressure before entering the stripper. Before recirculation of the H_2S lean solution into the absorber, the stream passes through a regenerative heat exchanger, heating up the load solution entering the stripping column. The lean solution has to be further cooled down to 30 °C and then repressurised at the absorber inlet condition level.

H_2S removal section working conditions are summarised in TABLE V.

Since the removal efficiency increases with the MDEA concentration in the solution, a 50% mass fraction MDEA/water solution has been chosen. This is the highest concentration allowable for this amine (Kohl and Riesenfeld, 1985) , limited by corrosion problems.

The total mass flow of the solution used in the absorber is set to 30 kg/s (0.97 kg solution/kg coal), thus achieving an H_2S removal efficiency of 95%.

TABLE V. H_2S REMOVAL SECTION WORKING CONDITIONS

Absorber working pressure [bar]	18.9
Absorber inlet syngas temperature [°C]	40
Absorber outlet syngas temperature [°C]	36.4
Absorber inlet solution temperature [°C]	30.5
Absorber outlet solution temp. [°C]	47.7
Stripper working pressure [bar]	1
Stripper inlet solution temperature [°C]	79.4
Stripper outlet solution temperature [°C]	99.5
Stripper outlet gas temperature [°C]	81
MDEA make up [kg/s]	0.001
Water make up [kg/s]	0.512

4.1 Heat recovery in the H_2S removal section

The raw syngas coming from COS hydrolysis section is first cooled from 140 to 40 °C, in order to condense water before entering the H_2S absorption column (*Figure 3*).

This cooling is regenerative: the released heat is recovered heating the CO_2 load solution in the CO_2 removal section.

The raw syngas reaches 87.8 °C and is further cooled down to 40°C - with external cooling water -, condensing thus 6.625 kg/s of water.

The stripping column reboiler heat duty is supplied in the steam cycle section (about 3294 kW – exactly 3269 kW per kg of H_2S - to rise the solution's temperature in the column bottom stage up to 99 °C). In fact the steam extracted for the shift section – at the required pressure – has a temperature higher than required; hence, it is possible to recover the needed reboiler duty by cooling down this steam.

The gas stream exiting from the stripping column is substantially H_2S and can be fed to a Claus process section.

The gaseous stream exiting from the absorption column - syngas free from H_2S (TABLE VI) - is directed to the shift reaction section.

5. Claus Process Section

The Claus process section contains two reactors. The third part of the H_2S rich gas enters the first reactor, where it is burnt to sulphur dioxide at 1100 °C (Kloster, 1999):

$$2\ H_2S + 3\ O_2 \leftrightarrow 2\ SO_2 + 2\ H_2O$$

The gas exiting from this reactor is mixed with the rest of the H_2S rich gas in the second reactor to produce elemental sulphur over a bauxite or alumina catalyst at about 200 °C (Merrick, 1984):

$$2\ H_2S + SO_2 \leftrightarrow 3\ S + 2\ H_2O$$

TABLE VI. SYNGAS LEAVING THE H_2S ABSORBER

Temperature [°C]	36.4	
Pressure [bar]	18.9	
Mass Flow [kg/s]	48.836	
Composition	Mass Frac.	Mole Frac.
H_2O	0.003	0.003
CO_2	0.156	0.076
N_2	0.007	0.005
CO	0.805	0.615
H_2S	0.001	0.001
H_2	0.028	0.3

On the whole, the heat released in the first reactor is greater than the heat required in the second one. This means that the process is self-sufficient from an energetic point of view (Lombardi, 2000).

With this configuration it is possible to convert 96% of the H_2S of the entering stream into elemental sulphur (Kloster, 1999), (Merrick, 1984).

6. Shift Reaction Section

The syngas stream, which leaves the H_2S removal absorbing column, is fed to a catalytic shift reaction section (*Figure 4*), where CO and water are converted into CO_2 and H_2:

$$CO + H_2O \rightarrow CO_2 + H_2$$

To enhance the conversion of CO, this reaction is realised in two reactors at about 450 and 250 °C respectively (Chiesa and Consonni, 1998).

Before entering the first reactor, the syngas is heated up to 230 °C - in a regenerative heat exchanger - by the syngas leaving the second shift reactor.

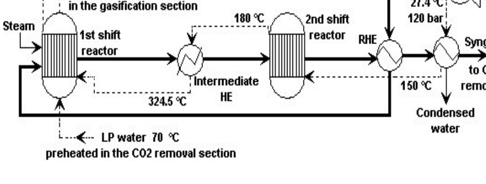


Figure 4. Shift reaction section scheme.

In the first reactor, steam at 230 °C and 18.6 bar is supplied for the reaction to take place in the ratio of 1.5 moles of water per mole of CO (37.91 kg/s) (Chiesa and Consonni, 1998). This steam comes from an extraction from the steam turbine.

The first reactor outlet stream (450 °C) is cooled down to 230 °C, in an intermediate shift heat exchanger, and fed to the second reactor.

Thanks to the catalysts, at the exit of the reactors, the CO conversion is 90 %; the CO_2 mass fraction passes from 0.156 to 0.729 and to 0.868 after the water condensation. The composition of the syngas leaving the shift section is reported in TABLE VII.

TABLE VII. SYNGAS LEAVING SHIFT SECTION

Temperature [°C]	50	
Pressure [bar]	18.3	
Mass Flow [kg/s]	65.439	
Composition	Mass Frac.	Mole Frac.
H2O	0.006	0.006
CO2	0.868	0.378
N2	0.005	0.004
CO	0.06	0.041
H2S	0.001	0
H2	0.06	0.57

6.1 Heat recovery in the shift reaction section

The shift reaction section is crucial for increasing the CO_2 concentration (thereby decreasing CO_2 total emissions after removal). The resulting syngas is very rich in H_2.

Of course, the better syngas quality is paid in terms of heating value (from 11535 to 7808 kJ/kg). This heat is recovered producing steam.

A high pressure (HP) water stream (17 kg/s) is preheated from 27.4 to 150 °C in the heat exchanger that condenses water from the syngas after the two shift reactors. Then, it is circulated in the second shift reactor where it is heated up to 180 °C and then fed to the intermediate shift heat exchanger, where it is brought to the saturation temperature (324.5 °C) and vaporised. The steam flow is then circulated through the first shift reactors where it is superheated to 440 °C. The gasification section - as previously seen – provides further superheating up to 560 °C in the convective heat exchanger upstream the cyclone.

A low pressure (LP) water stream (about 1 kg/s) - preheated form 25.7 to 70 °C by the stream of almost pure CO_2 exiting the stripper in the CO_2 removal section - is brought to saturation temperature (188 °C) and vaporized in the first shift reactor. Then it is superheated up to 255 °C in the convective heat exchanger upstream the cyclone.

7. CO_2 Removal Section: $DeCO_2$ Unit

The CO_2 removal - 92.3% - takes place in an absorbing column by chemical absorption with an amines aqueous solution. A blend of 50 % water, 25 % MDEA and 25 % DEA (mass basis) has been chosen, which was found to be a composition effectively reducing the energy demand for its regeneration (Corti and Manfrida, 1998). In order to reach a 92.3 % carbon dioxide removal, a solution mass flow of 17 kg solution per kg of coal (526 kg/s) has been selected. Such a high degree of removal is required to obtain a global removal efficiency of 85% referred to the CO_2 content in the exhausts of a standard IGCC, i.e. without shift and CO_2 removal sections.

The CO_2 load solution is first depressurised down to atmospheric pressure and then routed to a stripper column for thermal regeneration. The stripper column is assisted by a reboiler, which keeps the bottom stage temperature at 99 °C; the heat is supplied by steam extraction from the LP steam turbine. The lean solution, leaving the stripper bottom stage, goes through a regenerative heat exchanger, to heat the load solution before entering the stripper; it is then further cooled down to 30 °C; finally, it is repressurised and sent back to the absorber, after the eventual water and amines make up (*Figure 5*).

The gaseous stream leaving the stripper top is basically CO_2 and water; cooling to 30 °C allows to condense water and non-regenerated amines. The resulting gaseous stream contains 98 % in mass carbon dioxide; after compression to 80 bar in an intercooled staged compressor and liquefaction, it can be transported for final disposal or reuse.

TABLE VIII resumes the clean syngas and the separated CO_2 stream compositions.

7.1 Heat recovery in the CO_2 removal section

The CO_2 load solution temperature needs to be upgraded before entering the stripper, in order to minimise the heat duty to be supplied in the reboiler. To do this, first it goes through a regenerative heat exchanger - with the lean solution on the opposite side - where it reaches 83 °C, then through a multiple heat exchanger (MHE; see *Figure 5* for section layout and temperature details).

Inside the MHE, the syngas stream coming from the shift section, together with the H_2S absorber inlet stream (stream 2 in *Figure 5*), releases heat increasing the CO_2 load solution temperature. A temperature level of 97.5 °C is thus reached before entering the stripping column, and the reboiler heat duty results about 8 MW, which are supplied by steam extraction in the LP steam turbine.

The syngas is further cooled down to 50 °C, condensing 21.307 kg/s of water, before entering the absorber.

The gaseous stream leaving the stripper top end, at 97.5 °C, is first used to preheat 1.071 kg/s of LP water from 25.7 up to 70 °C (the same LP water stream was considered before, when dealing with the shift section) and then its temperature is further reduced - by means of external cooling water - down to 30 °C, condensing 65.7 kg/s of water and a small amount of amines previously evaporated in the stripper.

TABLE VIII. CLEAN SYNGAS AND CO_2 RICH STREAM CONDITIONS

	Syngas		**CO_2 stream**
Temperature [°C]	73.3		30
Pressure [bar]	18.3		1
Mass Flow [kg/s]	13.134		53.563
Composition	Mass Frac.	Mole Frac.	Mass Frac.
H2O	0.052	0.017	0.018
CO2	0.332	0.044	0.979
N2	0.026	0.005	0
CO	0.291	0.061	0.002
H2S	25 ppm	4 ppm	0.001
H2	0.298	0.872	0

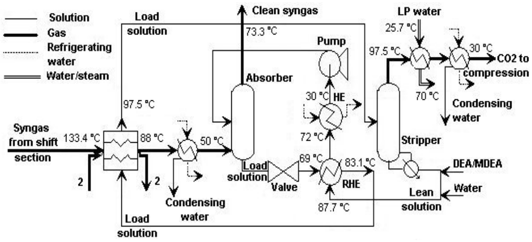


Figure 5. Schematic of the CO2 removal section

8. Power Generation Section

The Carbon Dioxide content of the gas stream leaving the CO_2 absorption column is decreased of 92.3 %; the syngas composition at this stage is reported in TABLE IX. Actually, this syngas is very rich in H_2 and has a quite good LHV of about 38700 kJ/kg, due mainly to absence of the dilution effect of CO_2.

The syngas exiting the CO_2 removal section - about 13 kg/s at 18.3 bar, 73.3 °C - is routed to the combustion chamber of the gas turbine. The compressor delivers 479 kg/s of air at 18.3 bar. The turbine inlet temperature is limited at 1277 °C. The products of combustion expand through the gas turbine to 1.1 bar, 582.6 °C with a 492 kg/s mass flow (see TABLE IX).

TABLE IX. POWER CYCLE WORKING CONDITIONS

Compressor power [kW]	230738
Compressor inlet flow rate [kg/s]	479
Compression ratio	18
Gas turbine inlet temperature [°C]	1277
Gas turbine outlet temperature [°C]	582.6
Gas turbine outlet pressure [bar]	1.1
Gas turbine outlet flow rate [kg/s]	492
Gas turbine power [kW]	430397
Net GT power [kW]	199659
Net ST power [kW]	132557
Cycle power [kW]	332216
Net cycle power [kW]	287760
Efficiency [%]	38.80
Specific CO_2 production [kg/MWh]	129.75
Specific Power [kJ/kg]	600

The gas turbine exhaust stream is passed through a heat recovery steam generator (HRSG), where about 69 kg/s HP (120 bar, 560 °C) and 4 kg/s LP (12 bar, 225 °C) superheated steam are produced.

Inside the HRSG, the gas temperature changes from 582.6 to 150.6 °C.

The HP steam produced in the HRSG is added to the 7 kg/s HP steam produced in the gasification reactor and to the 51 kg/s HP steam produced in the convective heat exchanger upstream of the cyclone; all HP streams are at 120 bar and 560 °C. Hence, about 127 kg/s of steam can be fed to the HP steam turbine.

The steam exiting the HP steam turbine (12 bar) is added to the 4 kg/s LP steam from the HRSG and to the 1 kg/s LP steam from the cyclone upstream convective heat exchanger, and then sent to the LP steam turbine.

Three steam extractions take place from the HP and LP steam turbines:

- 2.5 kg/s (20 bar level) for the gasification reactors process steam;
- 38 kg/s (18.6 bar level) for the shift reaction process;
- 3.795 kg/s (2 bar level) satisfy the CO_2 load solution heat duty.

The second extraction (18.6 bar) takes place at 300 °C, while it is needed at 230°C for the shift reaction. Hence, the excess heat is recovered for the regeneration of the H_2S load solution.

9. Overall Efficiency Calculation

The cycle power production is 332216 kW; the coal input energy is 741677 kW; this leads to a First Law efficiency level of 44.79 %.

From the above calculated net power, the consumption of the air separation unit (ASU) plus the power for O_2 compression from ASU outlet pressure - assumed to be 5 bar - to gasification reactor pressure - 20 bar - and the power for compression of carbon dioxide up to 80 bar still have to be subtracted. TABLE X reports these additional power requirements. The resulting net power is then 287760 kW and the efficiency 38.80% (TABLE IX).

TABLE X. ADDITIONAL POWER LOADS

Device	Specific consumption
ASU consumption	826 kJ/kgO_2
O_2 compression	154 kJ/kgO_2
CO_2 compression	355 kJ/kgCO_2

How much does it cost to remove CO_2?

Once obtained the above result, it is worth to compare it with the efficiency of a conventional IGCC plant, without any reduction of the CO_2 emissions.

To do this, a simulation based on the same power plant without CO_2 separation has been carried out: the syngas burnt in the GT combustion chamber comes directly from the H_2S absorption column. Some differences in the heat recovery section have also been accounted.

10. Power Generation Section for a Standard IGCC

The gaseous stream leaving the H_2S absorption column has the composition reported in TABLE VI. This syngas is much poorer in H_2 and, on the other hand, has a quite high content of CO. Its LHV is about 11535 kJ/kg: almost half than the previous one, but its mass flow is more than double (48 kg/s); the inlet energy to the GT combustor is thus higher: 563361 vs. 509121 kW.

The main gas turbine operating conditions are basically unchanged: the compressor units deliver 516 kg/s of air at 18.9 bar. The 565 kg/s combustion gas expands through the gas turbine and exits at 1.1 bar, 588 °C.

Due to the higher energy input, the air needed to reach the same turbine inlet temperature level is higher, hence also the GT exhaust heat content is larger.

In the HRSG, 76 kg/s HP (120 bar, 560 °C) and 4 kg/s LP (12 bar, 225 °C) superheated steam are produced. The HRSG working conditions have been left unchanged with respect to the plant with CO_2 removal.

The same amount of HP steam - 7.114 kg/s - is produced in the gasifier, while in the convective heat exchanger upstream of the cyclone 34.134 kg/s of HP steam and 2.07 kg/s of LP steam are produced.

Compared with that of the plant with CO_2 removal, only 2.5 kg/s at 20 bar process steam are extracted from the steam turbine for operation of the gasifier.

11. Overall Efficiency Calculation for the Standard IGCC

The power output of the standard IGCC is 370262kW; from this the ASU and O_2 compression power consumption must be subtracted; these are the same as for the IGCC with CO_2 removal. The net power production is 344439 kW that, referred to the same coal entering energy - 741677 kW - gives a First Law efficiency of 46.44 %.

TABLE XI summarises the working conditions for the ICGG without CO_2 removal.

TABLE XI. POWER CYCLE WORKING CONDITIONS WITHOUT CO_2 REMOVAL

Compressor power [kW]	252811
Compressor inlet flow rate [kg/s]	516
Compression ratio	18
Gas turbine inlet temperature [°C]	1275
Gas turbine outlet temperature [°C]	588.4
Gas turbine outlet pressure [bar]	1.1
Gas turbine outlet flow rate [kg/s]	564.8
Gas turbine power [kW]	468957
Net GT power [kW]	216146
Net ST power [kW]	154115
Cycle power [kW]	370262
Net cycle power [kW]	344439
Efficiency [%]	46.44
Specific CO_2 production [kg/MWh]	725
Specific Power [kJ/kg]	667

12. Life Cycle Assessment

The described power plants have been studied by means of Life Cycle Assessment.

This is a relatively uncommon application of LCA, usually carried out for material products easily identifiable in well defined objects, while the final product here considered is the power output. Here too, however, the different processes to obtain the identified product and all their requirements, in terms of materials and energy for the entire life cycle, are compared to assess their environmental impacts and, in particular, their contribution to the Greenhouse Effect, since the aim of the studied cycle is primarily the reduction of carbon dioxide emission in power generation.

The whole life cycle of the plant has been considered and, hence, the phases of construction, operation and dismantling have been included within the boundary of the study. This means that the inlet and outlet streams to and from the selected limit consist of the raw material flows and waste/emission flows respectively.

In order to identify the raw materials inlet flows, it is first necessary to perform an approximate sizing of the plant and to collect information about the weights, main materials, production processes and scrap outputs of all the relevant pieces of equipment needed to assembly the plant.

From the roughly-calculated amount of the employed main materials, it is possible to go back to the raw materials and to their manufacturing processes, i.e. energy consumptions, calculating the use of raw substances use and emissions, and hence the environmental impacts starting from the mining of the resources.

The same procedure has been used also for the operating phase, during which the inlet material flows - mainly fuel - have been traced back to use of natural resources and emissions, while the waste output from the plants - mainly stack gases - have been included on the basis of the model itself.

Concerning dismantling of the plant, the energy - and hence the related emissions - to disassemble, dispose and/or recycle materials has been calculated, according to the assumption of a waste scenario.

In order to pass from the manufactured materials to the raw substances and emissions inventory, SimaPro 4.0 (SimaPro, 1997) has been used. This is basically a data base able to reconstruct the "history" of several processes and materials and to aggregate the elemental pollutants inventory in order to obtain values for the selected environmental effect indicators.

Since the sizes of the compared plants, IGCC with $DeCO_2$ and standard IGCC, are different, as functional unit for the comparison the unit power output of 1 MJ has been chosen. Hence, the total Greenhouse Effect score in the whole life time is divided by the overall power production in the same period, obtaining the score per functional unit.

12.1 Plant construction

The construction phase consists in the assembly of the plant. As it is quite difficult to find detailed data about the weights, materials and scrap outputs, it was necessary to up-scale or down-scale the weights of known models.

For this plant, materials and weights of the equipment in TABLE XII have been considered. Globally, the total amount of the main materials used in the construction are summarised in TABLE XIII.

Heat exchangers have been roughly sized with the help of CHED (Intercept Software, 1987), while the data about weights and materials refers to Alfa Laval database. Information on materials and weights of a ASU refer to a Kobe Steel contact.

TABLE XII. EQUIPMENT CONSIDERED IN THE CONSTRUCTION OF THE IGCC WITH $DECO_2$.

Air Separation Unit
Gasifier Unit
Cyclone
Chemical reactors (COS-Hydrolysis, Claus process, shift reaction)
Heat exchangers
Gas turbine, compressor, combustion chamber, electric generator
Steam turbine, electric generator, steam condenser
HRSG
Absorber columns (H_2S & CO_2)
Stripper columns (H_2S & CO_2)
Cooling tower
Activated carbon filters
Compressor for CO_2

The gasifier unit weight is obtained from (Smith and Nieuwlaar, 1994); the cyclone has been sized with the help of Aspen (Aspen Plus ^{TM}, 1996); all the chemical reactors have been considered as absorption columns because of the lack of more accurate data; gas and steam turbine units and HRSG have been scaled from the data retrieved from previous works (Tober, 1997) (Mariabelli, 1999).

TABLE XIII. AMOUNT OF MATERIALS FOR THE CONSTRUCTION OF THE IGCC WITH $DECO_2$.

Material	**IGCC with $DeCO_2$**
Steel [kg]	9.18E+06
Steel Construction [kg]	2.84E+06
Cast Iron [kg]	1.96E+05
Copper [kg]	1.71E+05
Plastic [kg]	1.73E+05
Rubber [kg]	2.13E+03

Concrete [kg]	5.69E+07
Asphalt [kg]	6.00E+06

Transportation of the installed equipment has been considered, assuming an average distance of 1000 km. Moreover, on-site energy consumption has been considered, proportionally to the total weight, with reference to (Emmerson et al., 1995); 85% of the energy is produced by a dedicated diesel engine and the remaining is taken from the grid.

Absorption and stripping columns have been sized with the help of Sulzer indications. Activated carbon filters - needed for the regeneration of exhausted amines - (Alessandri, 1998) and cooling tower data come from Italian manufacturer contacts. The Carbon Dioxide compressor is scaled from a GE - NP model. Other items considered for the construction are piping, civil buildings - in particular the concrete and the steel to reinforce it - and the copper for the cabling. These data are obtained from a report scaling the weights on the basis of the total power of the plant (Smith and Nieuwlaar, 1994). Data about PVC for cabling and insulation of piping were retrieved from Cornelissen (1997).

12.2 Plant operation

A life time of fifteen years has been assumed.

During operation the consumption of coal,, activated carbon and amines (TABLE XIV) and maintenance for the major devices in the plant have been considered, (in particular, for gas turbine, heat exchangers, and HRSG).

The elemental resources and emissions inventory have been calculated by means of SimaPro.

During operation carbon dioxide due to combustion is discharged into the atmosphere and its total production in the life time is calculated on the base of 8760 working hours per year. The total amount of carbon dioxide emitted from combustion in the life time, according to the Aspen Plus simulation, results 4.91E+09 kg.

TABLE XIV. MAIN CONSUMPTION DURING THE IGCC WITH DECO$_2$ OPERATION.

Coal [ton]	1.47 E+07
Activated carbon [ton]	8909
Amines (DEA+MDEA) [ton]	15457

12.3 Plant dismantling

Dismantling includes disassembly of the plant, transportation on an average distance of 1000 km and recycling or disposal of materials. A disposal scenario for dismantled materials has been assumed:

- steel is recycled for 80 % and the rest is land filled
- cast iron is recycled for 90 % and the rest is land filled
- copper is 100 % recycled
- aluminium is 100% recycled
- concrete and asphalt are crushed and reused as a low quality landfill
- plastic is incinerated
- rubber is land filled
- steel in reinforced concrete is land filled.

On-site energy is also needed for dismantling, proportionally to the total weight of the installed equipment; the same reference data as of above (Emmerson et al., 1995) were used.

Due to the intensive recycle of some materials, the Greenhouse Effect indicator value can result in negative figure. In fact, the amount of the recycled materials is accounted as avoided new materials production and, hence, avoided (i.e. negative) emissions of pollutants. On the other hand, the dismantling/recycling processes are energy consuming and contribute with a positive amount of emissions. The overall result of the dismantling phase depends on the balance between these positive and avoided emissions.

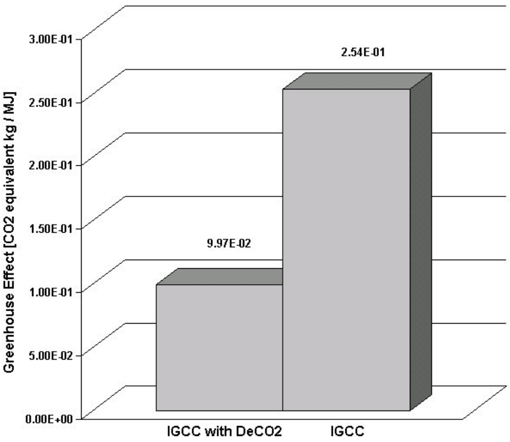
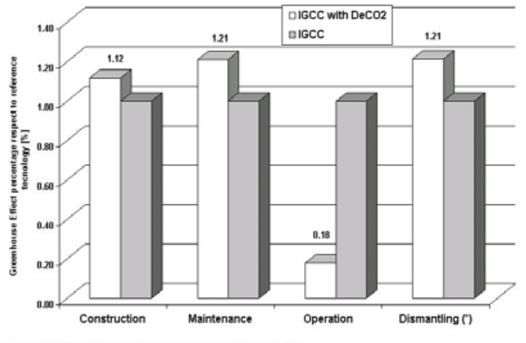


Figure 6. Comparison of Greenhouse Effect per functional unit for theIGCC with DeCO$_2$ vs. conventional IGCC.

13. LCA Results

The comparison between the IGCC with DeCO$_2$ and a conventional IGCC is shown in *Figure 6*, in term of Greenhouse Effect per functional unit. The amount of equivalent CO$_2$ emitted per MJ of power output is two and a half times higher in the conventional technology. In percentage, the emission per MJ in the IGCC with DeCO$_2$ is only 39% of that in to the IGCC.

Different effects are added together to obtain this result, as shown in *Figure 7*. In fact the IGCC with DeCO2 has a higher emission for construction and maintenance - respectively 12 and 21% more than the conventional plant. For the dismantling, the results should be interpreted in terms of avoided emissions; hence, the IGCC with DeCO2 has a higher value of avoided emissions (21%) due to the greater amount of recycled materials. The emissions in the operation phase are reduced to 18%. The same detailed comparison is shown in *Figure 8* in terms of Greenhouse Effect per functional unit.



(*) For dismantling the percentage is to be intended in term of avoided emissions

Figure 7. Greenhouse Effect, in the life time phases, for the IGCC with $DeCO_2$in percentage with respect to the reference conventional IGCC

In conclusion, it is true that the addition of all the sections necessary for CO_2 removal leads to an increase in the Greenhouse Effect production for construction, maintenance and dismantling; but it is also true that these phases are of very little importance when compared to operation, hence, the reduction in this phase is preponderant in the life cycle balance.

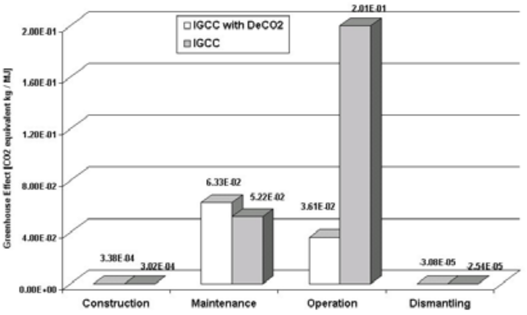


Figure 8. Comparison of Greenhouse Effect per functional unit, in the life time phases, between the IGCC with $DeCO_2$ and the conventional IGCC

14. ELCA Analysis

Exergetic Life Cycle Assessment (ELCA) is an analysis method based on a life cycle approach in combination with exergy analysis, developed by Cornelissen (1997). In this framework, exergy can be viewed as one of the possible impact indicators in a life cycle and, moreover, the most appropriate parameter for the depletion of natural resources, addressing the life cycle irreversibility.

As for the conventional LCA analysis, also for ELCA the life cycle of the IGCC with $DeCO_2$ has been divided in the three phases of construction, operation/maintenance and dismantling. The basic assumptions concerning the system boundary, the life time duration, etc. are the same of the LCA case and hence the same inventory analysis has been used as basis for ELCA calculation. The functional unit is again the unit power output: the results are expressed as MJ of destroyed exergy per MJ of power output.

The exergy destruction calculation for the operation phase has been carried out by means of a plant exergy analysis, while for construction, maintenance and dismantling the exergy of the system input streams has been calculated. Obviously the major contribution comes from the fuel, not only as direct source of energy, but also for material production processes. The exergy input to the system related to the fuel has been calculated from the low heating value of the fuels themselves (Szargut et al., 1988). Actually in this way the chemical exergy of other raw materials used in the production processes is not accounted for, but this amount seems to be negligible if compared with the fuel contribution (Lombardi, 2000). Moreover, another reason for not considering the chemical exergy of raw materials is that they don't undergo a structural transformation, and hence change in exergy content, during their use (as it happens for fuels), so the chemical exergy entering into the system with these materials is almost the same exiting from the system when the plant is dismantled.

15. ELCA Results

The exergy destruction in the life time of the IGCC with $DeCO_2$ is shown in *Figure 9*. The contribution to the total exergy destruction due to the construction and dismantling phase is almost negligible when compared to operation, where the transformation from fuel chemical exergy into thermal exergy takes place, and maintenance, in which the production process of the fuel itself is accounted. This trend is confirmed by the ELCA of other power generation cycles (Lombardi, 2000), (Tober, 1997), Mariabelli (1999).

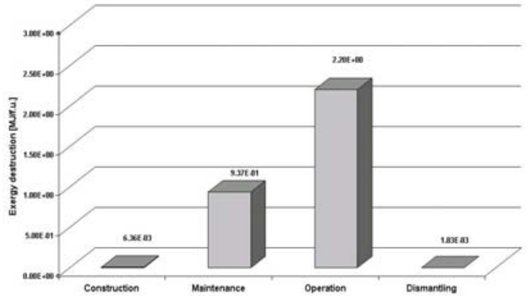


Figure 9. Exergy destruction per functional unit in the life cycle of the IGCC with $DeCO_2$

Figure 10 shows the different contributions to operation &maintenance exergy destruction. In this phase the exergy invested for the fuel production is the most important part, while the other contributions are almost negligible compared to this. In fact, in *Figure 10*, it is evident the negligible terms due to the devices operation & maintenance, and to the production/use of the activated carbons and amines.

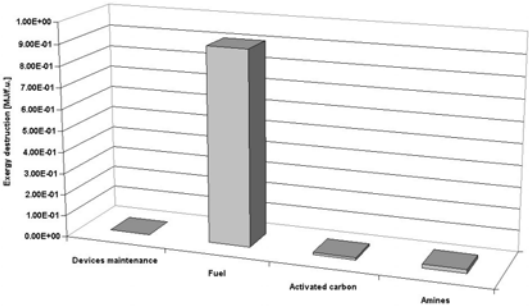


Figure 10. Different contributions to the exergy destruction of maintenance

16. Conclusions

The IGCC with CO_2 removal has a 38.8 % efficiency, which should be compared to a 46.4 % efficiency of a conventional IGCC without CO_2 removal: the complexity due to the addition of shift and CO_2 removal sections, and hence the related losses in the complex process of heat recovery - are the main responsibles for the efficiency reduction, while a minor role is played by the LP steam turbine extraction for the CO_2 load solution regeneration. Overall, a decrease of 7.6 efficiency points has been estimated.

No economic evaluations have been carried out in this study, but from this point of view it could be possible to estimate that the loss in term of cost efficiency would be lower with respect to energy efficiency. In fact, the commercial value of pure CO_2 and pure sulphur - obtained from the Claus process – should also be accounted for. These by-products can in fact improve the cost efficiency of the process partially compensating the decrease of the overall energy efficiency.

The LCA results show that the application of carbon dioxide absorption leads to a substantial decrease in the emissions.

Moreover, the main responsible phase for CO_2 production is operation, which is an order of magnitude greater than the other phases.

A similar conclusion can be drawn for the ELCA: the major exergy destruction is due to the operation & maintenance phase, which is two orders of magnitude greater than the other phases, in terms of MJ of destroyed exergy per functional unit.

These results confirm that, when studying the life cycle of this kind of products - i.e. electricity unit, power production cycles - the attention must be focused on the operation & maintenance phase, while the other phases, construction and dismantling, are almost negligible when compared to it.

Probably, it could be worth to consider again construction and dismantling phases when comparing systems with zero or very low CO_2 emissions in the operation. In fact, in this case, the additional construction and consumption of products for the removal could have an important weight in the total balance of Greenhouse Effect and exergy destruction.

List of Acronyms

IGCC	Integrated Gasification Combined Cycle
LCA	Life Cycle Assessment
$DeCO_2$	Decreasing CO_2 unit
ER	Equivalence Ratio
HE	Heat Exchanger
LP	Low Pressure
HP	High Pressure
DEA	DiEthanolAmine
MDEA	MethylDiEthanolAmine
MHE	Multiple Heat Exchanger
LHV	Low Heating Value
GT	Gas Turbine
ST	Steam Turbine
HRSG	Heat Recovery Steam Generator
ASU	Air Separation Unit
PVC	PoliVinilChloride
ELCA	Exergetic Life Cycle Assessment

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