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# Co-gasification of coal–petcoke and biomass in the Puertollano IGCC power plant



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## ABSTRACT

Integrated Gasification Combined Cycle plants (IGCC) are efficient power generation systems with low pollutants emissions. Moreover, the entrained flow gasifier of IGCC plants allows the combined use of other lower cost fuels (biomass and waste) together with coal. Co-firing with biomass is beneficial for the reduction of CO<sub>2</sub> emissions of fossil source.

In this paper the results of co-gasification tests with two types of biomass deriving from agricultural residues, namely 2% and 4% by weight of olive husk and grape seed meal, in the 335 MWe ISO IGCC power plant of ELCOGAS in Puertollano (Spain) are reported. No significant change in the composition of both the raw syngas and the clean syngas was observed.

Furthermore, a process simulation model of the IGCC plant of Puertollano was developed and validated with available industrial data. The model was used to assess the technical and economic feasibility of the process co-fired with higher biomass contents up to 60% by weight. The results indicate that a 54% decrease of fossil CO<sub>2</sub> emissions implies an energy penalty (a loss of net power) of about 20% while does not cause significant change of the net efficiency of the plant. The mitigation cost (the additional cost of electricity per avoided ton of CO<sub>2</sub>) is significantly dependent on the price of the biomass cost compared to the price of the fossil fuel.

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**Keywords:** Co-gasification; IGCC; Biomass; Carbon mitigation; Process simulation; Mitigation cost

## 1. Introduction

The conventional use of biomass in combustion plants has become an important component of the energy production and has received increasing attention for the global climate concerns due to the greenhouse gas emissions caused by the use of fossil fuels. Further incentive for the gasification of biomass comes from the much higher power generation efficiency that can be obtained in Integrated Gasification Combined Cycle (IGCC) plants with respect to power plants based on direct combustion and steam cycles.

As a result of the heightened concern over the global warming issue, there is an increased interest in renewable energy projects including those exploiting biomass. A number of

biomass gasification plant projects are under development in the USA and Europe. Most of these plants are based on the use of fluidized bed gasification technologies (Li et al., 2004; Corella et al., 2007). In addition, historically, there has been a large number of small fixed or moving bed gasifiers used throughout the world on a variety of biomass feedstocks particularly wood chips. However, fixed bed gasifiers have several limitations and drawbacks and because of these, gasifiers are not considered as prime candidates for biomass to power conversion in the OECD countries where more stringent emissions regulations would require very considerable additional cleanup equipment, labor and operating costs. Finally, no matter is the gasifier technology, these new projects have not paid sufficient attention to the flow properties (Tomasetta et al., 2013) and to

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### Nomenclature

3M	ternary mixture
CO <sub>2</sub> emi	fossil CO <sub>2</sub> emitted per unit energy, tCO <sub>2</sub> /MWh
COE	cost of electricity, €/MW <sub>net</sub>
GSM	grape seed meal
IGCC	Integrated Gasification Combined Cycle
LHV	low heating value, MJ/kg
HHV	high heating value, MJ/kg
MC	mitigation cost, €/tCO <sub>2</sub>
OH	olive husk

### Subscripts

base	base case with design fuel
bio	case for ternary mixture with biomass
gross	gross power
net	net power

the handling operations of biomass feedstock (Miccio et al., 2011) despite intermittent flow or blockage of these granular solids could often result in one of the major bottleneck of the whole process (Miccio et al., 2013; Barletta and Poletto, 2013).

In addition to this, there are also several biomass gasification projects in Europe in which the product gas from atmospheric fluidized bed gasification is co-fired with fossil fuels in existing boilers (Dai et al., 2008; Emami Tabatabaie et al., 2012). This approach has the advantage that it avoids having to build a completely dedicated biomass power plant. Biomass can be also added in small amounts to the feed in coal gasifiers without compromising the operations and quality of the slag, even though some ashes are high in alkalis such as Na and K at some percentage. On the other hand, the co-firing option could increase the cohesive properties of the feedstock powders and negatively affect their handling and feeding to the gasifier (Barletta et al., 2013).

One of the most significant cases of co-gasification is the IGCC plant in Buggenum (The Netherlands). As a consequence of the signature of the Kyoto protocol, this plant had to reduce its CO<sub>2</sub> emissions derived from coal. Therefore, in order to reach this target, it was decided to co-gasify biomass together with coal. The biomass feed has been mostly wood dust and dried sewage sludge. The wood was milled to ~1 mm at a central location and sieved out wood dust added from the wood board industry then transported by ship to Buggenum, where it is stored in silos and injected into the outlet of the coal mills (coal was milled to 90% <100 µm). It is a closed system apart from the single transfer from the ship. Sewage sludge, when dried to a similar level, can be fed below the classifier of the coal mills. Buggenum gasifier can handle up to 30% by weight of biomass (Damen et al., 2011). Also in recent years ELCOGAS has been running a project aimed at evaluating the effects of adding small percentages of biomass (up to 10 wt.%) on the performance of their IGCC plant in Puertollano (Spain), usually fed by mixtures of coal and petcoke (Casero and García-Peña, 2007; Fermoso et al., 2009).

Gasification of biomass and co-gasification of fossil fuels and biomass in IGCC plants has also been studied on modeling ground by means of thermodynamic analysis, detailed process simulations and economic analysis. The performance of an entrained bed gasifier as a function of the biomass content in the feed with coal and petcoke and of the oxygen and steam to fuel ratios was extensively analyzed by means of an

equilibrium reactor model (Pérez-Fortes et al., 2011). Srinivas et al. (2012) addressed the optimization of the operating conditions of a gasifier fed by rice husk in order to maximize the overall IGCC plant efficiency. In some studies a direct comparison in technical and economic terms between a fossil fuel fired IGCC and a biomass fired IGCC (also named BGCC or BIGCC) was addressed by means of process simulation tools including the optimization of the process heat integration (Klimantos et al., 2013; Tsakomakas et al., 2012). Tsakomakas et al. (2012) showed a higher profitability of a plant fed by wood chips in comparison to a plant fed by lignite when the CO<sub>2</sub> emissions price is larger than 27 €/t. Klimantos et al. (2013) instead assessed the possibility to operate in parallel a conventional coal fired IGCC and a BGCC and to integrate the gas turbines and the steam turbines streams of the two processes. According to the analysis by Klimantos et al. (2013) the coal fired IGCC plant and the combined BGCC and IGCC plant require cost of electricity comparable with that of a BGCC for CO<sub>2</sub> allowance larger than 55 €/t and 75 €/t, respectively.

The combination of biomass co-firing with pre-combustion carbon capture technologies was studied in order to obtain zero or negative carbon emissions for co-fired IGCC and BGCC, respectively (Rhodes and Keith, 2005; Cormos, 2009, 2012; Padurean et al., 2012; Pérez-Fortes et al., 2011). In some of these studies also the option of co-producing pure hydrogen in addition to electricity was analyzed in terms of electrical efficiency, hydrogen yield and CO<sub>2</sub> emissions (Cormos, 2009, 2012) and also in terms of gross profit calculated by accounting only the price of the fuels of electricity and of hydrogen (Pérez-Fortes et al., 2011).

The present work aims at studying the effect of co-gasification with biomass on the process performance of the IGCC plant of Puertollano on both experimental and modeling grounds. In particular, the paper reports in the first part the experimental results of co-gasification tests in the industrial plant with low biomass content and in the second part a techno-economic analysis performed by means of process simulation techniques also for biomass content up to 60% by weight.

## 2. Experimental

### 2.1. Plant description

ELCOGAS S.A. was founded on April the 8th 1992 to construct and exploit the Puertollano 335 MWe<sub>ISO</sub> IGCC demonstration power plant. The founding members were European electrical companies along with the main plant suppliers (ELCOGAS, 2000).

The Puertollano IGCC power plant consists of three completely integrated main units: gasification unit, air separation unit (ASU) and combined cycle (CC). A simplified flow diagram of the Puertollano IGCC is shown in Fig. 1. The design fuel is a mixture 50% by weight of autochthonous coal (with approximately 45% of ash) and 50% by weight of petcoke (with a 6% of sulphur). The technology is based on a coal gasification process, namely the conversion of coal into synthetic gas, which is then subjected to an exhaustive cleaning process, resulting in a combustible gas free of pollutants which can be burned with high efficiency in a combined-cycle electricity-generating unit.

The gasification unit uses the pressurized entrained flow process, called PRENFLO process (supplied by Krupp Koppers), for coal gasification. Syngas is produced by reaction of

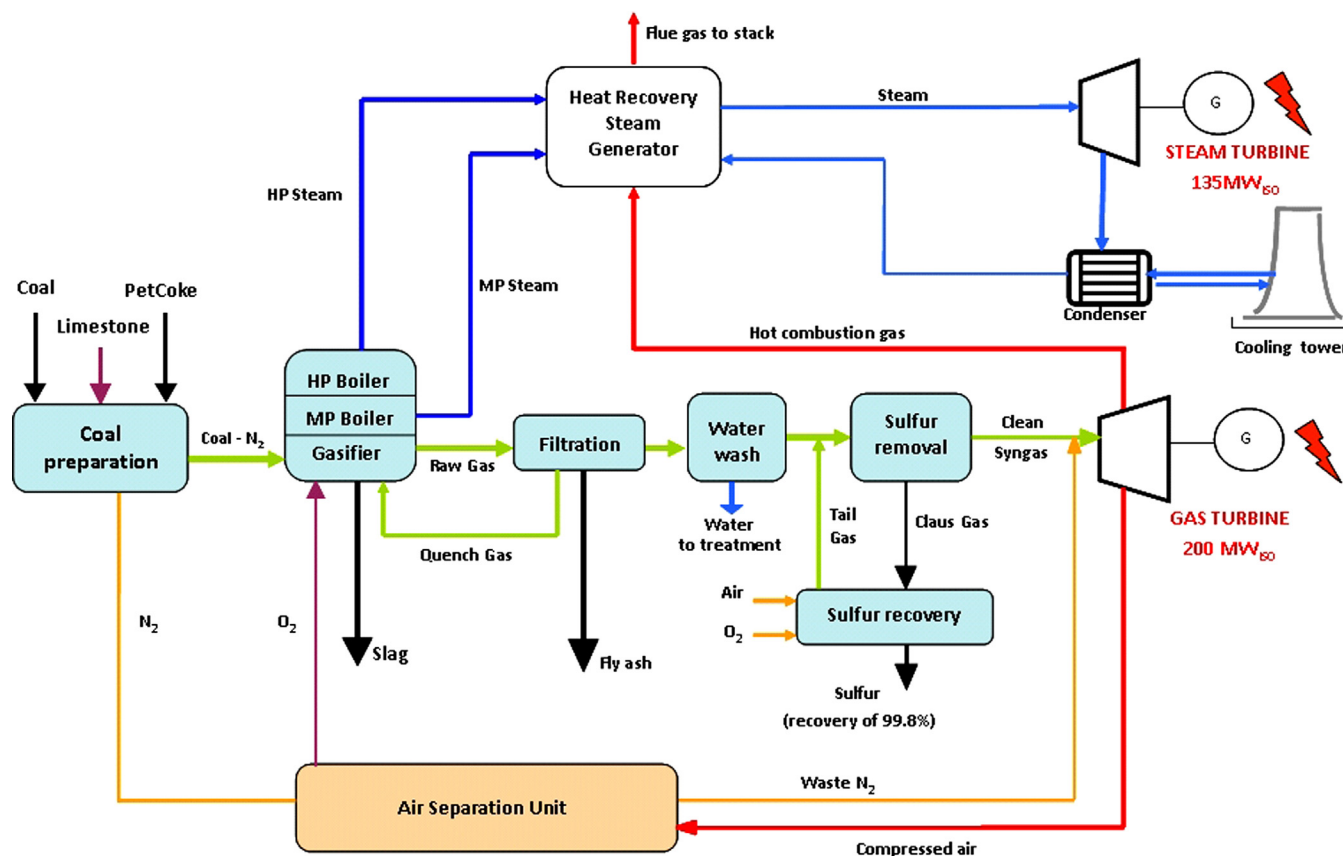


Fig. 1 – Puertollano IGCC simplified process block diagram.

a solid fuel with oxygen and steam at high temperatures up to 1600–1700 °C. The reducing raw gas produced in the gasifier is cooled down to approximately 800 °C using recycle gas free of solid particles (quench gas). Afterwards, the raw gas temperature is reduced down to 235 °C in two convective boilers (high and intermediate pressure boilers, HP and IP) with water tube syngas coolers. Subsequently, the syngas at the outlet of IP boiler is treated by removing the solid particles by candle filters, and by washing it in the Venturi scrubber for ammonia, HCl, HCN and trace component removal. Then, the sulphur compounds are eliminated (sulphur 99.9% removal) by means of a fixed bed of catalyst for carbonyl sulphide (COS) hydrolysis to H<sub>2</sub>S, and subsequently, removal of H<sub>2</sub>S is carried out using the MDEA (Methyl DiEtanol Amine) process. Then, the so-called clean gas passes firstly through a conditioning phase (saturation with water), before being sent to the gas turbine. The air separation unit, ASU, uses the air extracted from the gas turbine air compressor to generate, by means of a cryogenic process, both high purity oxygen (85%) for feeding the gasifier and nitrogen for pneumatic transportation of the fuel and as a safety system by rendering the installations inert. The combined cycle, CC, has a gas turbine, V94.3 model supplied by Siemens, which produces power and feeds the residual heat of the exhaust gases to a heat recovery steam generator. The latter equipment generates steam that is used to produce additional electric power in a conventional steam turbine with a condensation cycle. The gas turbine is able to operate with both syngas and natural gas allowing a great plant flexibility. The whole IGCC process has a gross efficiency of 47.2% and a net efficiency of 42.2% with respect to the low heating value of the fuel feed. Further details can be found in the ELCOGAS report (ELCOGAS, 2000).

## 2.2. Feedstock

The selection of suitable biomass was made at laboratory scale prior to performing the co-gasification tests in the IGCC plant. The following criteria to select biomass were followed: (a) the original particle size must be lower than 25 mm; (b) the biomass moisture content must be lower than 12%; (c) the biomass price must be lower than 100 €/t; and (d) the biomass must be available in large quantities. Accordingly, olive husk (locally named as orujillo), OH, and grape seed meal, GSM, were selected for the co-gasification tests. Table 1 reports the proximate analysis and the ultimate analysis of the design fuel (coal–petcoke 50% by weight each), of the OH and of the GSM upstream of the mills of the process preparation section. Ternary mixtures of coal, petcoke and biomass were prepared by adding different amounts of one biomass type (2% and 4% by weight) to the design fuel. Table 1 reports also the proximate analysis and the ultimate analysis of the four ternary mixtures prepared by the mills before feeding to the entrained flow gasifier.

## 2.3. Co-gasification test procedures

Since biomass is not the usual fuel used in the plant and has different properties with respect to those of the design fuel, ELCOGAS had to define a new procedure to introduce the biomass into the existing feeding system. As a result, the required amount of biomass was mixed with limestone (used to reduce the slag melting point in the gasifier) and fed to the limestone tank. Limestone and biomass were then mixed together with the Puertollano coal and petcoke, dried and ground in the existing mills. Subsequently, the homogeneous

**Table 1 – Fuel mixtures composition and thermal properties.**

Fuels	Design fuel	Upstream of mills			Downstream of mills		
		OH	GSM	3 M 2%OH	3 M 4%OH	3 M 2%GSM	3 M 4%GSM
Proximate analysis							
H <sub>2</sub> O (%)	9.40	9.97	13.84	0.76	1.03	0.73	0.89
Volatiles <sup>a</sup> (%)	17.57	67.22	66.50	19.41	19.91	18.64	19.12
Ashes <sup>a</sup> (%)	22.83	16.01	3.21	22.42	21.16	20.56	20.45
C <sub>fixed</sub> <sup>a</sup> (%)	59.60	16.76	30.30	58.16	58.92	60.79	60.43
Ultimate analysis							
C <sup>a</sup> (%)	65.35	45.78	52.20	65.04	65.35	66.90	67.36
H <sup>a</sup> (%)	3.09	5.19	5.69	2.38	3.52	3.24	3.68
N <sup>a</sup> (%)	1.50	1.13	1.78	1.40	1.36	1.46	1.36
S <sup>a</sup> (%)	3.54	0.16	0.12	3.38	3.30	3.55	3.39
Cl <sup>-a</sup> (mg/kg)	200	5798.6	178	314.9	370.7	282	233
LHV <sup>a</sup> (MJ/kg)	24.89	17.38	19.26	25.62	25.59	25.97	25.89
<sup>a</sup> Dry base.							

<sup>a</sup> Dry base.

ternary mixture, biomass/coal/petcoke, was fed to the gasifier. Previous studies demonstrated that the addition of biomass to the coal petcoke mixture might cause an increase of the cohesive properties of the resulting powder (Barletta et al., 2013a). Nevertheless, no significant change in the powder handling and feeding was observed during the industrial co-gasification tests.

## 2.4. Results

The total operating hours of the co-gasification tests were 506 h using 627.44 ton of olive husk and 605.10 ton of GSM. The operating hours and the biomass consumption for each test are reported in Table 2. The tests aimed at studying the biomass influence both on the IGCC power plant performance and on the CO<sub>2</sub> capture process of the pilot plant.

With concern to the influence of the biomass fed in the IGCC power plant, Table 3 reports the raw gas volumetric composition on dry basis obtained during the co-gasification tests. The concentration values for the test with 2% of OH are missing due to a temporary breakdown of the measurement system. The concentration values of main compounds (CO, H<sub>2</sub>, N<sub>2</sub>, CO<sub>2</sub>) are similar for all the tests and close to the average values obtained when using the common fuel. The concentration values of minor compounds such as H<sub>2</sub>S, COS and CH<sub>4</sub> are of the same order of magnitude of the values obtained when using the common fuel. As a result, it can be concluded that the addition of small amounts of biomass does not significantly affect the raw gas composition.

Table 4 reports the volumetric composition on dry basis of the clean gas obtained during the co-gasification tests. The concentration values of the main compounds are generally similar to those obtained with the common fuel. A closer inspection of the table reveals that CO<sub>2</sub> slightly decreases while CO slightly increases with increasing biomass content. Minor compounds such as H<sub>2</sub>S and CH<sub>4</sub>, keep concentration values of the same order of magnitude of those obtained with the common fuel. Differently, the COS concentration is lower than that obtained with the common fuel. This result can be explained considering that the COS concentration value for the common fuel is higher because it results from an average over much longer periods including cyclic variations of the performance of the COS hydrolysis catalyst and of the MDEA absorber. Despite these differences, on the whole the effect of low biomass content in the feed on the clean syngas composition is very limited.

During the entire period of the co-gasification tests, the IGCC power plant produced a gross power of 134,578.96 MWh whose 56,268.50 MWh were provided by the steam turbine, and 78,310.46 MWh were provided by the gas turbine. The total net electrical power produced was 117,387.45 MWh. The data corresponding to each of the tests are reported in detail in Table 5.

Finally, during the co-gasification tests the effect of biomass addition on the CO<sub>2</sub> capture pilot process was also studied. In particular, the CO<sub>2</sub> capture and H<sub>2</sub> production pilot plant of 14 MWth installed in the Puertollano IGCC power plant was operated for 249 h, being fed with synthesis gas with sulphur compounds, i.e. raw gas (Table 3). During these tests, no significant difference was observed in comparison with tests with common fuel. More details on these results are reported in a recently published paper (Casero et al., 2013, 2014).

## 3. Process simulation

### 3.1. Process simulation model

The IGCC plant of ELCOGAS in Puertollano was modeled by the research group at the University of Salerno using the steady-state process simulator Aspen Plus version 7.2. The process flowsheet of the base case IGCC plant includes a feed preparation section, an air separation unit (ASU), an entrained flow gasifier, a sequence of syngas cleaning section and a combined cycle power generation. The different sections of the IGCC plant were modeled as hierarchy blocks (sub-flowsheets) of the simulation flowsheet reported in Fig. 2. The modeling approach and main assumptions for each section are reported in the following sections.

#### 3.1.1. Gasifier

The Prenflo gasification reactor was simulated using a modular-sequential approach by a sequence of a yield reactor model (Ryield), a stoichiometric reactor (RStoic) and an equilibrium reactor (RGibbs) (Bhattacharyya et al., 2011). In the yield reactor the non-conventional components of the feed (coal, petcoke, biomass) are decomposed into basic conventional species according to atomic balances (C, O<sub>2</sub>, H<sub>2</sub>, S, N<sub>2</sub>, Cl<sub>2</sub>, H<sub>2</sub>O, ash). The stoichiometric reactor simulates the oxidation of carbon to CO<sub>2</sub> and sulphur conversion into H<sub>2</sub>S assuming conversion degree values of 98.8% and 90%, respectively, on the base of experimental values of the industrial gasifier of the Puertollano plant (ELCOGAS, 2000). The stream



**Table 2 – Main operating data of the co-gasification tests.**

Co-gasification test	Date	Duration (h)	Biomass consumption (t)
3 M 2%OH	November–December 2011	106	218.12
3 M 4%OH	October 2012	153.5	409.32
3 M 2%GSM	October 2012	127	179.36
3 M 4%GSM	November–December 2012	119.5	425.74
Total		506	1232.54

**Table 3 – Raw syngas composition (on dry basis) during the co-gasification tests.**

Parameters	3 M 2%OH	3 M 4%OH	3 M 2%GSM	3 M 4%GSM	Design fuel average 1999–2011
CO <sub>2</sub> (vol %)	N.A.	2.05	2.32	1.64	2.83 ( $\sigma = 0.62$ )
CO (vol %)	N.A.	60.75	60.28	60.62	60.28 ( $\sigma = 0.90$ )
H <sub>2</sub> (vol %)	N.A.	21.99	21.68	21.99	21.88 ( $\sigma = 0.76$ )
N <sub>2</sub> (vol %)	N.A.	13.56	14.17	13.84	13.11 ( $\sigma = 0.73$ )
Ar (vol %)	N.A.	0.97	0.97	0.97	0.97 ( $\sigma = 0.09$ )
CH <sub>4</sub> (ppmv)	N.A.	57	51.67	76	83.32 ( $\sigma = 22.55$ )
H <sub>2</sub> S (ppmv)	N.A.	4831	3628	7141	7329 ( $\sigma = 2200$ )
COS (ppmv)	N.A.	1908	2070	1617	1632 ( $\sigma = 1039$ )

N.A.: not analyzed due to problems in the sampler.

**Table 4 – Clean syngas composition (on dry basis) and LHV during the co-gasification tests.**

Parameters	3 M 2%OH	3 M 4%OH	3 M 2%GSM	3 M 4%GSM	Design fuel average 1999–2011
CO <sub>2</sub> (vol %)	2.12	1.59	2.01	1.43	2.18 ( $\sigma = 0.46$ )
CO (vol %)	61.12	61.91	60.64	60.96	61.00 ( $\sigma = 0.80$ )
H <sub>2</sub> (vol %)	23.47	22.72	22.15	23.06	22.02 ( $\sigma = 0.45$ )
N <sub>2</sub> (vol %)	12.33	12.94	14.19	13.7	13.72 ( $\sigma = 0.65$ )
Ar (vol %)	0.95	0.84	1	0.85	1.01 ( $\sigma = 0.04$ )
CH <sub>4</sub> (ppmv)	79.1	109.5	84.63	89.3	99.82 ( $\sigma = 9.34$ )
H <sub>2</sub> S (ppmv)	0	<0.1 <sup>a</sup>	<0.1 <sup>a</sup>	<0.1 <sup>a</sup>	5.00 ( $\sigma = 8.49$ ) <sup>b</sup>
COS (ppmv)	34.1	5.45	2.85	3.7	50.07 ( $\sigma = 39.36$ ) <sup>b</sup>
LHV (MJ/kg)	10.27	10.24	9.92	10.21	9.926 ( $\sigma = 177$ )

<sup>a</sup> Values are lower than the quantification limit.

<sup>b</sup> Values include operation periods with final lifetime of the COS catalyst and/or the sorbent.

**Table 5 – Gross power and net electrical power during the co-gasification tests.**

Test	Gross power (MWh)			Net electrical power (MWh)
	Gas turbine	Steam turbine	Total	Total
3 M 2%OH	18,247.2	12,694.7	30,941.9	26,625.28
3 M 4%OH	21,772.37	16,148.26	37,920.63	34,341.33
3 M 2%GSM	16,065.54	12,353.23	28,418.77	24,391.52
3 M 4%GSM	22,225.35	15,072.31	37,297.66	32,029.32
Total	78,310.46	56,268.50	134,578.96	117,387.45

leaving this reactor is fed to an equilibrium reactor together with other streams namely a 85% pure oxygen stream and a 99.9% pure nitrogen stream coming from the ASU, the medium pressure steam coming from the heat recovery section and the limestone feed (95% CaCO<sub>3</sub>, 5% ash). The oxygen flow rate was chosen on the basis of the equivalence ratio, ER (the ratio between the oxygen mass flowrate and the stoichiometric oxygen mass flowrate for the fuel mixture), while the steam flow rate was chosen on the basis of the steam ratio, SR (the mass ratio between steam and carbon in the fuel mixture) reported for real cases. The equilibrium reactor RGibbs evaluates at constant temperature and pressure the final raw syngas composition (or partial pressure distribution) by minimizing the Gibbs free energy of the reacting system subject to constraints deriving from the atomic mass balances. Further details on the optimization method can be found in thermodynamics

**Table 6 – Reactions accounted in the gasifier model.**

Reaction	Model assumption	Stoichiometry
RG1	Equilibrium	C + H <sub>2</sub> O ↔ CO + H <sub>2</sub>
RG2	Equilibrium	C + 0.5O <sub>2</sub> ↔ CO
RG3	Equilibrium	H <sub>2</sub> + S ↔ H <sub>2</sub> S
RG4	Temperature approach	CO + H <sub>2</sub> O ↔ CO <sub>2</sub> + H <sub>2</sub>
RG5	Equilibrium	CH <sub>4</sub> + H <sub>2</sub> O ↔ CO + 3H <sub>2</sub>
RG6	Temperature approach	N <sub>2</sub> + 3H <sub>2</sub> ↔ 2NH <sub>3</sub>
RG7	Equilibrium	COS + H <sub>2</sub> O ↔ CO <sub>2</sub> + H <sub>2</sub> S
RG8	Equilibrium	H <sub>2</sub> + Cl <sub>2</sub> ↔ 2HCl
RG9	Equilibrium	CaCO <sub>3</sub> ↔ CaO + CO <sub>2</sub>
RG10	Equilibrium	CO + NH <sub>3</sub> ↔ HCN + H <sub>2</sub> O

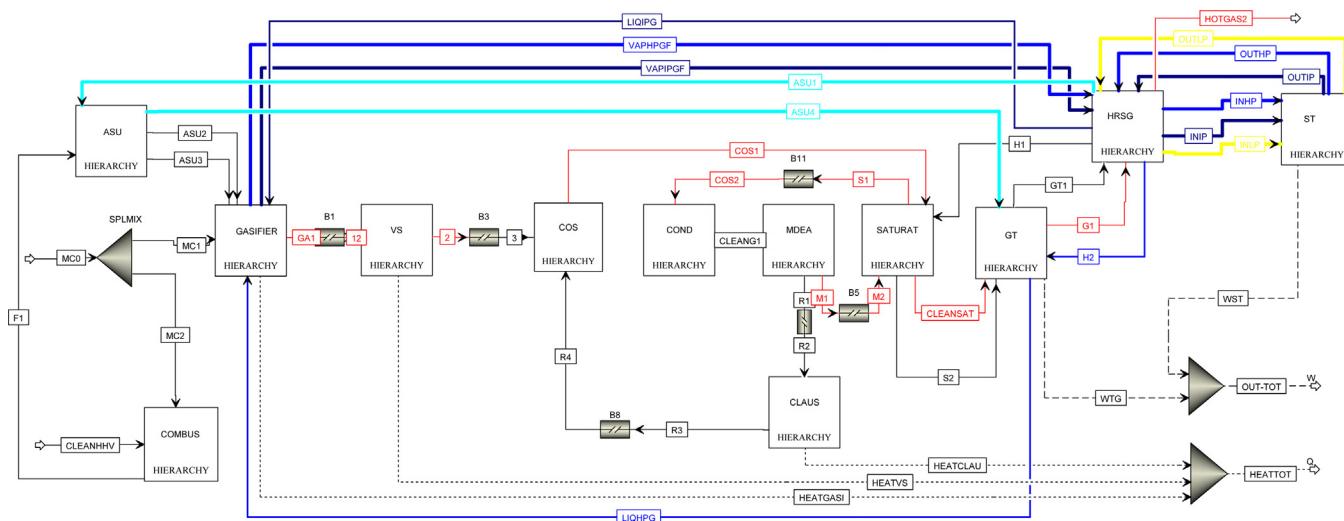


Fig. 2 – Simulation flow sheet of the IGCC power plant of ELCOGAS.

textbooks (i.e. Smith et al., 2001). The reactions considered are listed in Table 6. Reactions R4 and R6 were not considered at equilibrium, while restricted equilibrium was assumed and as a result two “temperature approach to equilibrium”  $\Delta T$  were introduced as model parameters (Bhattacharyya et al., 2011). In particular, this assumption is based on a restricted chemical equilibrium in which the chemical equilibrium constant is evaluated by the classical exponential equation with the standard Gibbs energy change at  $T + \Delta T$  in spite of  $T$ , where  $T$  is the actual gasifier temperature and  $\Delta T$  is the temperature approach. As a result, the molar fraction of the species involved in reaction R4 and R6 are also subject to the constraint of the relevant equilibrium constant evaluated at  $T + \Delta T$ . Therefore, the Gibbs energy objective function was minimized accounting for the constraints deriving from the atomic mass balances and the constraints of the restricted chemical equilibrium constant equations for the reactions R4 and R6. Optimal  $\Delta T$  values for these reactions were found by searching the best fitting between the simulation results and the experimental data of the raw syngas composition for different coal-petcoke mixtures available in ELCOGAS report (ELCOGAS, 2000). The thermodynamic model used in this section was the Peng & Robinson equation of state. Heat recovery from the gasifier was accounted for and coupled with the Heat Recovery Steam Generator of the Combined Cycle, as explained in further detail in Section 3.1.3. The separation of the slag (including the bottom ashes, the unconverted calcium and sulphur) occurring by means of the withdrawal from the gasifier bottom and the raw syngas filtration to remove the fly ashes are simulated together by means of a single ideal separator.

### 3.1.2. Cleaning section

The syngas cleaning section consists in a sequence of a Venturi scrubber, VS, a carbonyl sulphide, COS, hydrolysis reactor, a sour water steam stripper, SWS, a methyl diethanol amine, MDEA, absorber, a sulphur recovery Claus plant.

**3.1.2.1. Venturi scrubber.** The absorption of hydrogen sulfide, ammonia pollutants and carbon dioxide from the syngas in a water–NaOH solution in the Venturi scrubber was modeled by a rigorous multistage vapor–liquid model addressing mass transfer phenomena as well. Electrolytic reactions in the liquid phase (reported in Table 7) were accounted for and

Table 7 – Reactions accounted in the Venturi scrubber model.

Reaction	Stoichiometry
RS1	$\text{HCN} + \text{H}_2\text{O} \leftrightarrow \text{CN}^- + \text{H}_3\text{O}^+$
RS2	$\text{H}_2\text{O} + \text{HSO}_4^- \leftrightarrow \text{H}_3\text{O}^+ + \text{SO}_4^{--}$
RS3	$\text{H}_2\text{O} + \text{H}_2\text{SO}_4 \leftrightarrow \text{H}_3\text{O}^+ + \text{HSO}_4^-$
RS4	$\text{NH}_3 + \text{HCO}_3^- \leftrightarrow \text{H}_2\text{O} + \text{NH}_2\text{COO}^-$
RS5	$\text{H}_2\text{O} + \text{NH}_3 \leftrightarrow \text{OH}^- + \text{NH}_4^+$
RS6	$\text{H}_2\text{O} + \text{HCO}_3^- \leftrightarrow \text{CO}_3^{--} + \text{H}_3\text{O}^+$
RS7	$2\text{H}_2\text{O} + \text{CO}_2 \leftrightarrow \text{HCO}_3^- + \text{H}_3\text{O}^+$
RS8	$\text{H}_2\text{O} + \text{HS}^- \leftrightarrow \text{H}_3\text{O}^+ + \text{S}^{--}$
RS9	$\text{H}_2\text{O} + \text{H}_2\text{S} \leftrightarrow \text{H}_3\text{O}^+ + \text{HS}^-$
RS10	$2\text{H}_2\text{O} \leftrightarrow \text{OH}^- + \text{H}_3\text{O}^+$

consequently the thermodynamic model used in this section was the electrolyte NRTL. The model parameter was the ratio between the water–NaOH solution flow rate and the raw syngas flow rate ( $\text{H}_2\text{O}$ –NaOH/RSG). The polluted water regeneration and recycle was also addressed by modeling the acid and basic water treatment columns.

**3.1.2.2. COS hydrolysis reactor.** The COS hydrolysis reactor converting COS into  $\text{H}_2\text{S}$  was modeled as a stoichiometric reactor with an assigned conversion degree of 0.9. The thermodynamic model used in this section was the Peng & Robinson EOS.

**3.1.2.3. MDEA absorber.** The MDEA absorption column to remove  $\text{H}_2\text{S}$  and  $\text{CO}_2$  was modeled with a rigorous multistage vapor–liquid equilibrium model including the electrolytic reactions in the liquid (reported in Table 8). Solvent recovery was modeled by a stripper column with the same modeling approach. The thermodynamic model used in this section was the electrolyte NRTL. The model parameter was the flow rate

Table 8 – Reactions accounted in the MDEA absorber model.

Reaction	Stoichiometry
RA1	$\text{H}_2\text{O} + \text{MDEA}^+ \leftrightarrow \text{MDEA} + \text{H}_3\text{O}^+$
RA2	$\text{H}_2\text{O} + \text{HCO}_3^- \leftrightarrow \text{CO}_3^{--} + \text{H}_3\text{O}^+$
RA3	$2\text{H}_2\text{O} + \text{CO}_2 \leftrightarrow \text{HCO}_3^- + \text{H}_3\text{O}^+$
RA4	$\text{H}_2\text{O} + \text{HS}^- \leftrightarrow \text{H}_3\text{O}^+ + \text{S}^{--}$
RA5	$\text{H}_2\text{O} + \text{H}_2\text{S} \leftrightarrow \text{H}_3\text{O}^+ + \text{HS}^-$



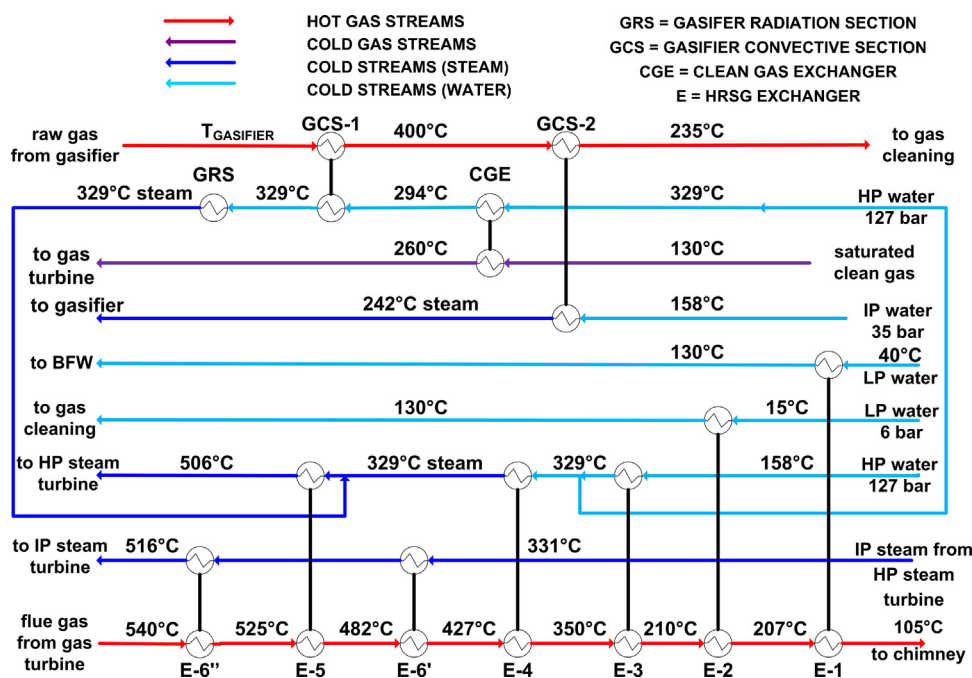


Fig. 4 – Heat exchanger network scheme.

the outlet temperature of the hot flue gas leaving the reboiler E4 could be 350 °C in order to ensure a  $\Delta T$  non-lower than 20 °C with the vaporizing HP water at 329 °C. The flow rate of the high-pressure water stream affects the amount of steam that expands in the steam turbines and, thus, the generated electric power.

The HEN is finally completed by a heat exchanger coupling a split of the high pressure boiling water stream (HP water from E-3) with the clean gas which needs to be heated up to 260 °C before entering the gas turbine (heat exchanger CGE in Fig. 4).

Heat integration was also carried out to exploit the hot air leaving the compressor train to generate additional low pressure steam. Finally, the hot syngas leaving the COS reactor was thermally coupled with cold syngas leaving the MEA absorption column.

### 3.2. Comparison between process simulation results and industrial data

The process simulation model was previously validated with the industrial data published by ELCOGAS (ELCOGAS, 2000) for coal–petcoke mixtures. In particular, main model parameters (namely, the gasifier “temperature approach to equilibrium”  $\Delta T$ s, the  $H_2O$ – $NaOH$ /RSG of the Venturi scrubber and the makeup stream flow rate of amines of the MDEA column) were found by searching the best fitting between the simulation results and the experimental data of raw syngas and clean syngas compositions for four different coal–petcoke mixtures (39%/61%, 45%/55%, 54%/46%, 58%/42% by weight). The model was then validated by comparing simulation results and experimental data for the design case with the coal–petcoke mixture (50% by weight each) in terms of raw syngas and clean syngas compositions and of electrical power (Sofia et al., 2013).

The process simulation model was used to predict the IGCC performance when fed by the ternary mixtures formed by coal–petcoke and biomass used in the experimental industrial tests reported in Section 2. The total fuel feed rate and the feed

ratios for oxygen and steam used in the simulations were the same of the industrial tests.

The comparison between the model prediction results and the industrial data was performed in terms of raw and clean syngas composition on dry basis ( $N_2$ ,  $H_2$ ,  $CO$ ,  $CO_2$ ,  $Ar$ ). Fig. 5 reports the clean syngas composition for the 2% and 4% GSM mixtures and for the 2% and 4% olive husk mixtures. For both raw and syngas composition a good agreement between the model results and the experimental results was obtained. In fact, the root mean square error calculated as a function of raw and clean gas composition values is lower than 10%. With concern to single species errors, larger discrepancies were observed for low concentration substances like  $CO_2$ . Fig. 6 shows the comparison between the model results and the industrial data in terms of gross power and net power for

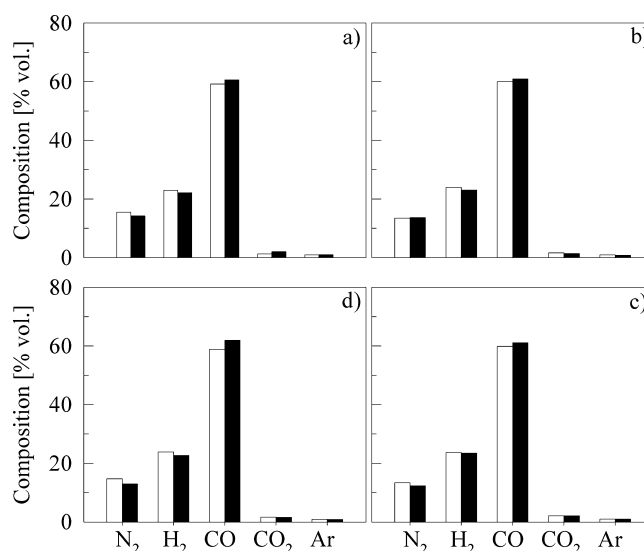
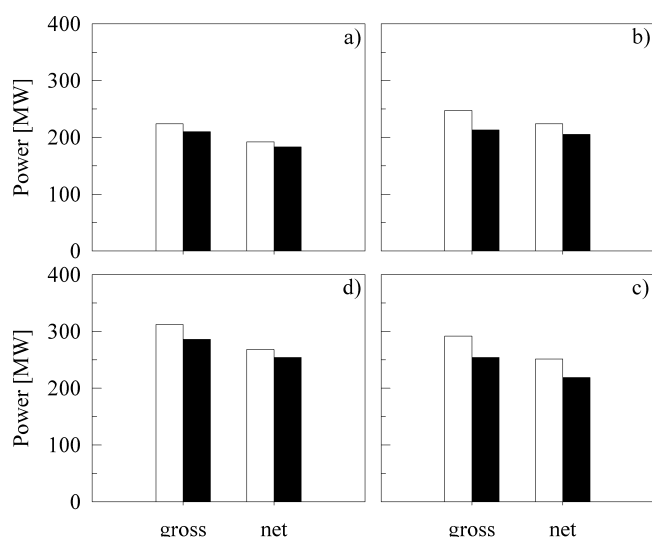


Fig. 5 – Clean syngas composition on dry basis for coal–petcoke–biomass mixtures: (a) 2% grape seed meal, (b) 4% grape seed meal, (c) 2% olive husk, and (d) 4% olive husk. White bars, experimental results; black bars, model results.





**Fig. 6 – Gross and net power for coal-petcoke-biomass mixtures: (a) 2% grape seed meal, (b) 4% grape seed meal, (c) 2% olive husk, and (d) 4% olive husk. White bars, experimental results; black bars, model results.**

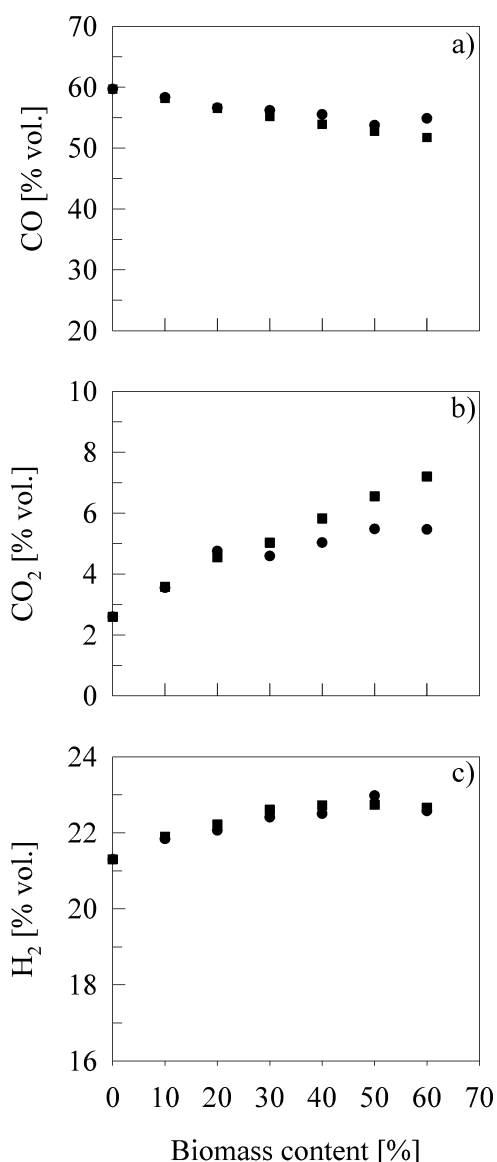
the 2% and 4% GSM mixtures and for the 2% and 4% olive husk mixtures. Also in this case the model results are in good agreement with the experimental results since the root mean square error does not exceed 11% for all ternary mixtures.

### 3.3. Process simulation results for co-gasification with higher biomass content

The process simulation model was also used to predict the performance of the IGCC plant fed by ternary mixtures with biomass content in the range 0–60% by weight. In fact, industrial experimental data for feeds with high percentage of biomass are not available. The latter upper limit was chosen as a function of a chosen minimum acceptable heating value of the ternary mixtures. In fact, a decrease is due to the lower heating value of both biomass samples. In particular, the total LHV value decrease up to 19.7 MJ/kg and 20.6 MJ/kg for the OH 60% by weight and for the GSM 60% by weight mixtures, respectively. The total fuel fed rate was kept constant and equal to that of the base case with the design fuel, while the ER and the SR of the gasifier were changed according to an empirical function of the fuel high heating value, HHV, derived from industrial data provided by ELCOGAS (2000). In particular, the ER decreases while the SR increases with increasing HHV.

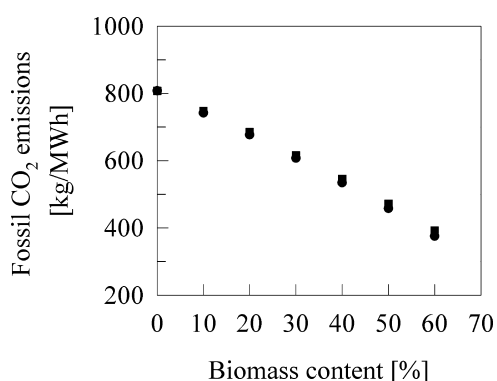
The effect of the biomass content on the composition of the clean syngas is shown in Fig. 7. Inspection of the plots reveals that increasing biomass content causes an increase of  $\text{CO}_2$  concentration at the expense of CO concentration. This result can be explained by considering the higher oxygen/fuel feed ratio to the gasifier for the mixtures with higher biomass content. In fact, as mentioned in Section 3.2, the gasifier ER increases with decreasing HHV of the fuel mixture and, thus, with increasing biomass content. Differently, an almost negligible effect of biomass content on hydrogen concentration in the clean syngas is observed.

The influence of biomass presence in the feed on the overall technical performance of the IGCC plant is shown in Figs. 8 and 9. In particular, Fig. 8 reports the  $\text{CO}_2$  emissions coming from fossil source as a function of the biomass content in the feed. A linear decrease of the  $\text{CO}_2$  emissions can

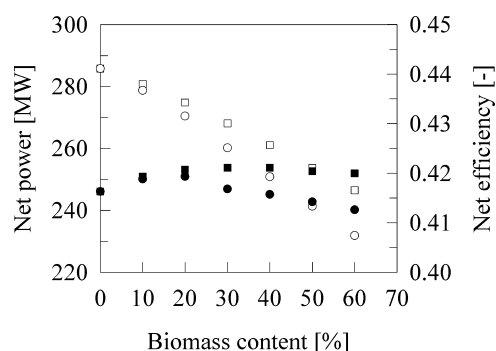


**Fig. 7 – Model results: concentration of main species in the clean syngas as a function of the biomass content in the fuel feed: (a) CO; (b)  $\text{CO}_2$ ; and (c)  $\text{H}_2$ ; ■, olive husk; ●, grape seed meal.**

be observed with increasing biomass percentage. In particular, more than 50% reduction of  $\text{CO}_2$  emissions was recorded for mixtures with 60% biomass content. This encouraging result is balanced by a loss of produced net power with increasing



**Fig. 8 – Model results: fossil  $\text{CO}_2$  emissions as a function of the biomass content in the fuel feed: ●, olive husk; ■, grape seed meal.**



**Fig. 9 – Model results: net power and net efficiency as a function of the biomass content in the fuel feed; ○, net power for olive husk; □ net power for grape seed meal; ●, net efficiency for the olive husk; ■, net efficiency for the grape seed meal.**

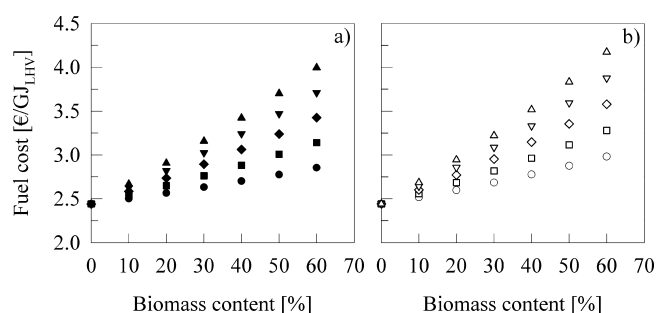
biomass content, as reported in Fig. 9. In fact, by inspecting the plots it can be noticed that the net power decreases with respect to the base case feed of about 14% for the 60% GSM mixture and of about 19% for the 60% OH mixture. This power loss is due to the lower LHV of the two biomass samples compared to the coal-petcoke mixtures. In fact, the power loss percentage values correspond to the decrease percentage values of the LHV of the ternary mixtures with respect to the base case coal-petcoke mixture. This finding is further confirmed by the plot of the net efficiency of the plant (calculated as the ratio between the net power and the product of the feed mass flow rate and its LHV), also reported in Fig. 9, showing a negligible effect of the biomass content in the feed.

### 3.4. Economic analysis

A simple economic analysis was carried out to assess the impact of the cost of the different fuels mixtures on the cost of energy. In particular, the fuel cost per unit energy, was calculated as follows:

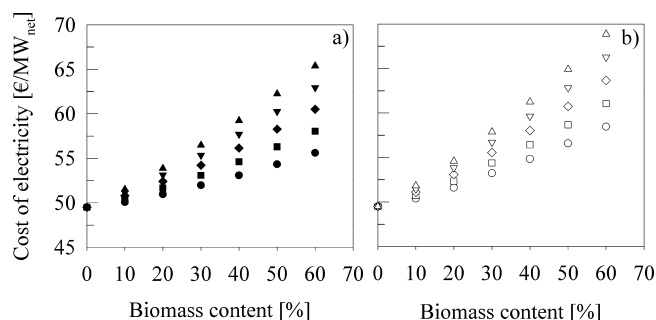
$$\text{Fuel cost} = \frac{\text{Cost}_{\text{fuel}}}{\text{LHV}_{\text{feed}}} = \left[ \frac{\text{€}}{\text{GJ}_{\text{LHV}}} \right] \quad (1)$$

where  $\text{Cost}_{\text{fuel}}$  is the cost per ton of feedstock and  $\text{LHV}_{\text{feed}}$  is the low heating value of the fuel mixture. A reference cost of 60 €/t was assumed for the coal-petcoke mixture. Since the cost of biomass is significantly affected by market fluctuations, the cost of OH and of GSM was varied in the range 60–100 €/t. Fig. 10 reports the fuel cost as a function of the biomass percentage in the feed for different biomass costs. The increase of the fuel cost of energy varies between 16% and 64% for the 60% GSM mixture and between 22% and 70% for the 60% OH mixture depending on the biomass cost. Moreover, the effect of the fuel cost on the final cost of electricity, COE, was assessed by a more rigorous profitability analysis based on the calculation of the discounted cash flows. In particular, a total investment cost of 460 M€ was assumed for the Puertollano IGCC plant at the time of the project (Campbell et al., 2000). This cost was not updated intentionally since the aim was to estimate the effect of the different fuel cost and plant performance on the COE over the entire life of a plant built years ago. Additional assumptions regarded approximate estimates of the operations and maintenance fixed costs (6.94 €/MWh<sub>gross</sub>), operations and maintenance variable costs (2.30 €/MWh<sub>gross</sub>). A construction period of 4 years and a linear depreciation over 22 years were assumed. Finally, an interest



**Fig. 10 – Fuel cost per unit low heating value as a function of the biomass content in the fuel feed; (a) grape seed meal for different biomass cost: ●, 60 €/ton; ■, 70 €/ton; ◆, 80 €/ton; ▼, 90 €/ton; ▲, 100 €/ton and (b) olive husk for different biomass cost: ○, 60 €/ton; □, 70 €/ton; ◇, 80 €/ton; ▽, 90 €/ton; △, 100 €/ton.**

rate of 3% was used to calculate the discounted cash flows. The cost of electricity was calculated as the break even electricity price (i.e. the price that makes the net present value, NPV, equal to zero after 25 years of plant life). Fig. 11 reports the COE as a function of the biomass percentage in the feed for different biomass costs in the range 60–100 €/t. The increase of the COE varies between 11% and 31% for the 60% GSM mixture and between 17% and 38% for the 60% OH mixture depending on the biomass cost. It is worth comparing these results with those of other techno-economic analyses on IGCC plants co-fired with coal and biomass. However, as reported in the survey of Section 1, published techno-economic studies concern coal fired or biomass fired IGCC, while data on co-fired plants are not available. Moreover, it is well known that the COE values are strongly dependent on the method of calculation and on the economic parameters used in the analysis. As a result, also a comparison with other analyses on IGCC plants using biomass is not straightforward. Nevertheless a comparison was attempted with the analysis performed by Craig and Mann (1996). This NREL report described the economic study on wood chips-based IGCC of smaller scale (50–150 MWe) with three types of gasifiers namely: (1) a high pressure air-blown single stage fluidized bed gasifier, (2) a low pressure indirectly heated gasifier, (3) a low pressure air-blown gasifier. Both an aeroderivative gas turbine and an utility gas turbine were examined for the combined cycle. The calculation of the COE used by Craig and Mann (1996) was derived as the break even price of electricity like in the present work,



**Fig. 11 – Cost of electricity as a function of the biomass content in the fuel feed; (a) grape seed meal for different biomass cost: ●, 60 €/ton; ■, 70 €/ton; ◆, 80 €/ton; ▼, 90 €/ton; ▲, 100 €/ton and (b) olive husk for different biomass cost: ○, 60 €/ton; □, 70 €/ton; ◇, 80 €/ton; ▽, 90 €/ton; △, 100 €/ton.**

**Table 9 – Parameters and results of the economic analysis of the present work and of the work by Craig and Mann (1996).**

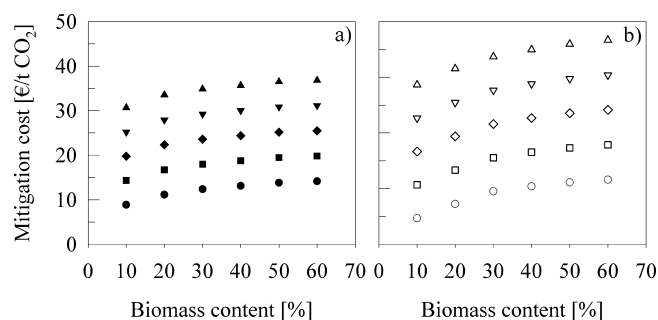
Parameter	246 MWe Puertollano IGCC 60% GSM (present work)		56 MWe NREL BIGCC 100% wood chips. HP gasifier, aero-derivative gas turbine (Craig and Mann, 1996)		132 MWe NREL BIGCC 100% wood chips. HP gasifier, advanced utility gas turbine (Craig and Mann, 1996)	
	This paper	This paper revised	NREL	This paper	NREL	This paper
<i>Economic assumptions</i>						
Book life (years)	25	25	30	25	30	25
Construction (years)	4	4	2	4	2	4
Contingency (%)	10	10	15	10	15	10
Discount rate (%)	3	3	6.9	3	6.9	3
Construction period (years)	4	4	2	4	2	4
Total plant cost (M€)	460.00	460.00	65.30	71.32	134.00	146.41
Cost of biomass (€/t)	60	31	31	31	31	31
Operating costs including fuel (M€/year)	61.86	49.26	9.95	10.87	21.26	23.22
Net power (MWh)	246.5	246.5	56	56	132	132
Specific investment (€/kW)	1866	1866	1209	1274	1016	1109
COE (€/MWh)	55.6	48.5	45.2	41.2	39.9	37.5
CO <sub>2</sub> neutral emission (%)	46	46	100	100	100	100

but different economic parameter values were assumed. A comparison between the economic parameters used in this present paper analysis and in the NREL analysis is reported in Table 9. Moreover, Table 9 reports also the results of the economic analyses estimated in different ways. In particular, the first column reports the results of the analysis on the Puertollano IGCC co-fired with 60% GSM and assuming the base case economic parameters of this study. The second column, instead, reports the results for the same case and with the same assumptions except for the biomass cost that was set to 31 €/t, the price assumed for wood chips by Craig and Mann (1996). The comparison between the results reported in the first and in the second column further confirmed that the cost of biomass had a significant effect on the final COE. The third column reports the results of the NREL analysis on a 56 MWe BIGCC with a high pressure air-blown gasifier fed by 100% wood chips and an aero-derivative gas turbine. The fifth column reports the results of the NREL analysis on a 132 MWe BIGCC with a high pressure air-blown gasifier fed by 100% wood chips and an advanced utility gas turbine. Economic parameters for columns three and five were those assumed by Craig and Mann (1996). For the sake of a more consistent comparison, the fourth and the sixth columns reports the results on the same 56 MWe BIGCC plant and 132 MWe BIGCC plant, respectively, recalculated in this work as follows: (a) the total capital investment and the operating costs reported by Craig and Mann (1996) for year 1993 were updated to 1999, the reference year for the costs of Puertollano IGCC, by using the Chemical Engineering Plant Cost Index (CEPCI) values of years 1993 and 1999 and (b) the economic parameters used for the base case analysis of the Puertollano IGCC were assumed. A comparison of the results indicates that the COE value for the Puertollano IGCC fed by 60% GSM is significantly higher than the COE values of both BIGCC considered in the NREL analysis. This result can be explained, despite its larger scale, considering the much higher specific investment

cost of the Puertollano IGCC plant which is characterized by a more complex gas cleaning section with respect to the NREL BIGCC. Coming back to the cases analyzed in this work, the difference between the cost of energy of a co-fired plant and that of a coal–petcoke fired plant was related to the avoided CO<sub>2</sub> emissions due to the partial substitution of fossil fuels with renewable fuels. This relationship was calculated in terms of the so-called mitigation cost, defined as follows:

$$MC = \frac{COE_{bio} - COE_{base}}{CO_2emi_{base} - CO_2emi_{bio}} = \left[ \frac{\text{€}}{tCO_2} \right] \quad (2)$$

where  $COE_{bio}$  and  $COE_{base}$  are the gross cost of energy for the ternary mixtures with biomass and for the base case with the coal–petcoke mixture, respectively,  $CO_2emi_{bio}$  and  $CO_2emi_{base}$  are the corresponding tons of fossil CO<sub>2</sub> emitted per MWh.  $CO_2emi_{bio}$  values were already reported in Fig. 8. The mitigation cost as a function of the biomass content in the mixture is



**Fig. 12 – Mitigation cost as a function of the biomass content in the fuel feed; (a) grape seed meal for different biomass cost: ●, 60 €/ton; ■, 70 €/ton; ◆, 80 €/ton; ▼, 90 €/ton; ▲, 100 €/ton and (b) olive husk for different biomass cost: ○, 60 €/ton; □, 70 €/ton; ◇, 80 €/ton; ▽, 90 €/ton; △, 100 €/ton.**

reported in Fig. 12. The results show the mitigation cost value increasing from 9 to 14 €/tCO<sub>2</sub> for increasing GSM content and from 15 to 22 €/tCO<sub>2</sub> for increasing OH content, assuming a biomass cost of 60 €/t. Instead, assuming a biomass price of 100 €/t. the MC values increase from 31 to 37 €/tCO<sub>2</sub> for increasing GSM content and from 39 to 47 €/tCO<sub>2</sub> for increasing OH content. Higher mitigation costs for the OH biomass are due to its lower LHV affecting the total net power and thus the consequent COE. Conclusions

Co-gasification tests performed in the Puertollano IGCC power plant with mixtures consisting of coal-petcoke and 2% or 4% by weight of two types of biomass (olive husk and grape seed meal) showed a negligible effect of low biomass content on the composition of both the raw syngas and the clean syngas. Consequently, the overall energy performance of the plant, in terms of power and of efficiency, is substantially unchanged.

Process simulations carried out for ternary mixtures with higher biomass contents (up to 60% by weight) showed a linear decrease of the CO<sub>2</sub> emissions of fossil source with increasing biomass percentage. A 50% reduction of CO<sub>2</sub> emissions was obtained for mixtures with 60% biomass. Co-gasification with biomass does not affect the net efficiency of the power plant. In fact, the decrease of the produced net power for increasing biomass content in the feed (up to about 14% for the 60% GSM mixture and of about 19% for the 60% OH mixture) is due by the lower LHV of the biomass with respect to that of the coal-petcoke mixture fuel.

The results of a simple economic analysis, based on the eventual additional cost of biomass with respect to the traditional fossil fuel, show mitigation cost values varying in a wide range: 9–39 €/tCO<sub>2</sub> for a 10% biomass mixture and 14–47 €/tCO<sub>2</sub> for a 60% biomass mixture in dependence of the assumed biomass cost varying in the range 60–100 €/t. This results in a critical role of the biomass price for making the mitigation cost sustainable and comparable with the values of the EU emissions allowance which dropped from 25 €/tCO<sub>2</sub> to 4 €/tCO<sub>2</sub> in the last five years (European Energy Exchange, 2013).

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