Co-Gasification of Coal and Biomass in an IGCC Power Plant: Gasifier Modeling

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Abstract

Co-gasification of coal and biomass in an existing coal-fired IGCC power plant is proposed as an efficient, flexible and environmentally friendly way to increase the biomass contribution to electricity generation. A model of an entrained flow gasifier is described and validated with nearly 3,000 actual steady-state operational data points (4,800 hours). The model is then used to study co-gasification of coal, petroleum coke and up to 10 percent of several types of biomass. As a result, the influence of fuel variations on gasifier performance and modifications in operation that should be made in co-gasification are obtained. A conclusion of our study is that co-gasification is possible provided that operation is properly adapted. A validated model can be very useful for predicting operating points for new fuel mixtures.

Keywords: Co-gasification, model, gasifier, IGCC, biomass

1. Introduction

The use of biomass to produce electricity has several advantages. First, biomass is a renewable energy with near zero net CO_2 emissions. Second, it is a local resource that reduces energetic dependence and creates jobs. Last but not least, biomass power plants can be easily integrated in the electric grid because their load does not depend on weather (unlike wind farms or solar stations).

However, since biomass is a dispersed resource, and transport can increase its cost, large biomass power plants cannot be built. Although the most usual option is to use small biomass power plants (less than 25 MWe) based on combustion and steam cycles, interest in gasification is growing because it provides higher efficiency and better environmental results (Overend, 2000; VTT, 2002). Biomass gasifiers are usually combined with internal combustion engines, either in CHP or for producing electricity only. Another interesting option that can achieve higher efficiency is the use of biomass-fired IGCC (Integrated Gasification Combined Cycle) power plants like the Värnamo, ARBRE, and Thermie Energy Farm demonstration projects (Overend, 2000; Stahl and Neergard, 1998; Pitcher, 2000; De Lange and Barbucci, 1998). The first plant provided 6 MWe and 9 MWth to a district heating system, while the others generated 8 and 14 MWe, respectively. They all use air blown gasifiers.

However, the use of small biomass-fired plants has several disadvantages (high specific cost, low efficiency, and shut-down risk when there is a biomass shortage), which can be avoided by mixing biomass and coal in the same power station, either by building new bi-fuel plants or (what is quicker and cheaper) by adapting an existing coal plant. One way to do this is by burning coal and biomass (Hein and Bemtgen,1998). Another option consists of gasifying biomass and burning the gaseous fuel in a coal boiler (VTT, 2002). Here, the co-gasification of coal and biomass in an existing IGCC power plant is proposed. Due to the advantages of coal IGCC power plants, this option provides high efficiency and low environmental impact.

Co-gasification of coal and biomass in fixed-bed and fluidised bed gasifiers has been studied at a bench scale level by several authors. Some of them have detected synergistic effects (Sjöström et al., 1999) while others disagree with them (Collot et al., 1999). On a pilot scale, Rheinbraum AG and the British Coal Corporation have successfully studied the co-gasification of sewage sludge and coal in a high temperature Winkler gasifier and in a fluidised

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bed gasifier (Minchener, 1999). British Coal has studied the techno-economic feasibility of the co-gasification of coal with straw or sewage sludge in an oxygen-blown integrated gasification combined cycle (IGCC) and in an air-blown gasification combined cycle (ABGC) (Minchener, 1999). The University of Essen has investigated the use of coal/biomass combinations for IGCC applications, concluding that up to 10 percent biomass in an oxygen-blown entrained flow gasifier is technically feasible although net electrical efficiencies would be slightly lower due to the energy needed for biomass pre-treatment (Minchener, 1999). The co-gasification of coal and biomass in the Buggenum IGCC power plant has also been proposed in a study that consists of two parts: a preliminary desk study (Ree, 1997) and an exploratory experimental work (Korbee et al., 1998).

The Elcogas Power Station in Puertollano, Spain is a 330 MWe IGCC plant that consumes a mixture of local coal and petroleum coke (50/50 percent weight ratio). Since a continuous stable operation has been achieved, the use of new fuels is being considered. The work shown in this paper is part of a techno-economic feasibility study of the co-gasification of coal, coke, and biomass. Since a resource evaluation and a pre-treatment study showed that between 5 and 10% of the fuel could be replaced by biomass, the aim of this work is to verify that operation with the new mixtures is possible without big operational modifications. To do so, a validated model of the gasifier is used.

2. The Model of the Gasifier

2.1 Description of the gasifier

The Puertollano IGCC power plant furnishes a PRENFLO (PRessurised ENtrained FLOw) gasifier, which is fed with fuel (a mixture of high-ash local coal, high-sulphur petroleum coke, and a small amount of limestone to favour ash fluidisation), oxygen (85% purity), steam and nitrogen. These flows react very quickly at high temperatures generating a combustible gas (mainly composed of CO and H_2) that leaves the reaction chamber through its upper part. A flow of slag is removed from the bottom. Gas leaving the reaction chamber is quenched with a cold gas stream in order to stop gas phase reactions and enter the evaporators with adequate operating conditions. The gas is then cooled in an HRSG (heat recovery steam generator) to arrive at a temperature at which it can be cleaned. The gas used for quenching is taken from the cold gas flow that leaves the HRSG (*Figure 1*). Since the amount of fuel is fixed by the synthesis gas that the turbine demands and the nitrogen flow depends on the fuel (pneumatic conveying),

operators can control gasification reactions by adjusting the flows of oxygen and steam (actually oxygen/fuel and steam/fuel ratios). The quench gas flow is also modified, but this parameter is used to control temperature distributions and fouling in the HRSG and is not considered here.

Several models have been proposed to simulate the reaction chamber of an entrained flow gasifier. Van der Burgt (1998) uses a simple model based on constant fuel conversion ratio (that avoids the simulation of the gasification process) and gas phase equilibrium. Wen (1979) proposes a model based on the division of the gasification process into three stages (volatilisation and volatiles combustion, char combustion, and char gasification) and by using an unreacted-core-shrinking model in the simulation of gas-particle interactions.

Since the gasifier has been built, a CFD model that provides the value of properties throughout the gasifier is not needed. However, dependence of the fuel conversion ratio on operating conditions should be considered.
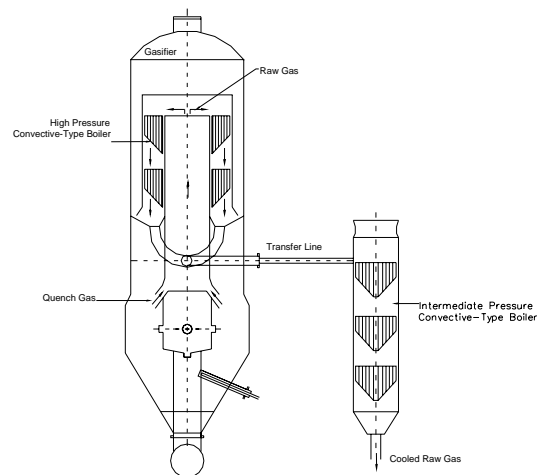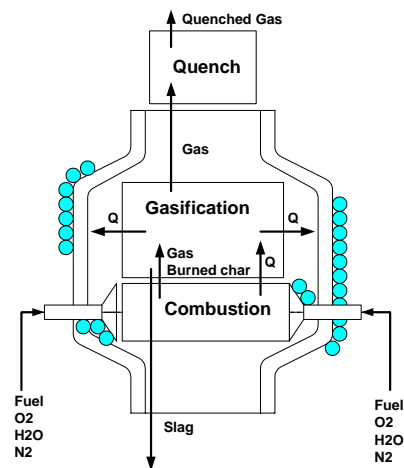

Figure 1. Gasifier and HRSG. Source: Elcogas

Figure 2. Gasifier schematic.

2.2 Description of the model

The model used to simulate the reaction chamber has been developed by Martínez (Usón, 2002). In order to introduce the dependence of coal conversion with gasification conditions, the gasification process suffered by a fuel particle has to be simulated. As indicated above, this process is divided into three stages: i) volatilisation and volatiles combustion, ii) combustion, and iii) gasification. In each stage, products are calculated by mass balances, and the kinetics is studied. These kinetics studies are used in the first two stages to calculate their duration and in the last stage to calculate its degree of development from the time available (calculated by subtracting total residence time minus the time taken by the first two stages).

Volatilisation is the decomposition of coal into volatiles and a carbon residue called char. This process can be represented by the following reaction:

$$CH_{hf}O_{of}N_{nf}S_{sf}\left(H_2O\right)_w Z \rightarrow$$
$$CH_hO_oN_nS_sZ + V + w \cdot H_2O \quad (1)$$
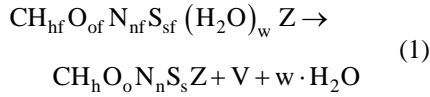
where $CH_{hf}O_{of}N_{nf}S_{sf}(H_2O)_wZ$ is the molecular formula of the fuel, $CH_hO_oN_nS_sZ$ is the molecular formula of the char, and V are the volatiles. To simulate this stage, Loison and Chauvin (1964) provide correlations for volatiles composition and Badzioch and Hawksley (1970) provide expressions for the real amount of volatiles (which differ from the proximate analysis) and for the kinetics. As volatiles are released, they are burned.
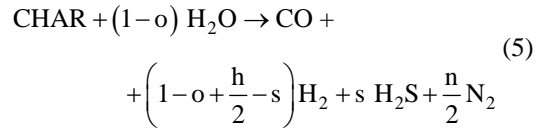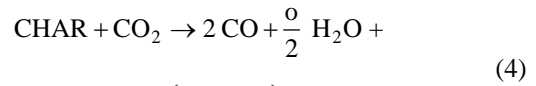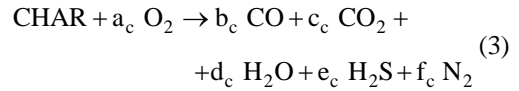
In the combustion and gasification stages, char particles react with the gas. To represent this interaction, the unreacted-core-shrinking model mentioned above is used (Wen, 1968). This model assumes that chemical reactions take place on a spherical surface that separates the core that has not reacted with the ash cover where non-mineral matter has already been consumed. As combustion or gasification advances, this surface is displaced towards the centre. In this model, mass convection around the particle, diffusion through the ash layer, and reaction kinetics on the core surface are considered. Thus,

$$\frac{1}{k_t} = \frac{1}{h_{gas}} + \frac{r_p - r_c}{y}\frac{1}{k_{ash}} + \frac{1}{y^2 k_q} \quad (2)$$

where k_t is the specific total reaction rate, h_{gas} the external convection coefficient, k_{ash} the specific diffusion rate of the gas through the ash layer, r_p the particle radius, r_c the core radius, y the ratio between them, and k_q the chemical rate constant. Since there are several gas-char reactions, there is one value of k_t for each one. These values and the partial pressures of the gaseous reactants are used to obtain the fraction of the char involved in each reaction and to relate the char consumed in each stage to the duration of this stage.

During the combustion stage, the particle reacts with O_2 (combustion), CO_2 (Boudouard reaction), and H_2O (steam gasification), which have flowed through the ash layer from the gas to the core. Thus,

$$CHAR + a_c\ O_2 \rightarrow b_c\ CO + c_c\ CO_2 +$$
$$+d_c\ H_2O + e_c\ H_2S + f_c\ N_2 \quad (3)$$

$$CHAR + CO_2 \rightarrow 2\ CO + \frac{o}{2}\ H_2O +$$
$$+\left(\frac{h}{2} - s - o\right)H_2 + s\ H_2S + \frac{n}{2}\ N_2 \quad (4)$$

$$CHAR + \left(1-o\right)H_2O \rightarrow CO +$$
$$+\left(1 - o + \frac{h}{2} - s\right)H_2 + s\ H_2S + \frac{n}{2}N_2 \quad (5)$$
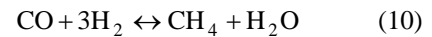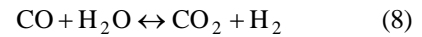
The products of these three reactions flow from the core surface to the gas where combustible species are burned.

The end of oxygen determines the end of the combustion stage and the beginning of gasification. In this third stage, the particle also reacts with CO_2 and H_2O. Since there is no O_2, there is no combustion but instead a H_2 reaction (when there was O_2, H_2 reacted with it in the gas phase and did not flow to the core surface) given by

$$CHAR + \left(2 + o + s - \frac{h}{2}\right)H_2 \rightarrow$$
$$CH_4 + o\ H_2O + s\ H_2S + \frac{n}{2}\ N_2 \quad (7)$$
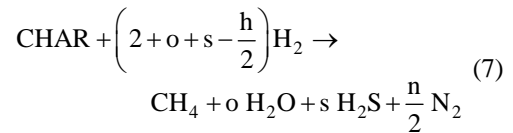
At the same time, in the gas phase, the composition of the gas leaving the reaction chamber is determined by the shift reaction, COS formation, and CH_4 formation equilibrium reactions, i.e.

$$CO + H_2O \leftrightarrow CO_2 + H_2 \quad (8)$$

$$H_2S + CO_2 \leftrightarrow COS + H_2O \quad (9)$$

$$CO + 3H_2 \leftrightarrow CH_4 + H_2O \quad (10)$$

Finally, the reaction chamber is divided into two isothermal zones: combustion and gasification. Volatilisation and volatiles combustion and char combustion stages take place in the first one and gasification in the second. In each zone, temperature is calculated by an energy balance. In these balances, heat

transfer between two zones and heat transfer from the gasification zone to the reactor walls are also considered.

The model presented above is suitable for simulating the reaction chamber and provides the composition of the gas that leaves this chamber. However, the quench (equilibrium reaction freezing due to mixing with a cold gas stream) is not instantaneous and gas phase equilibriums are slightly displaced before cooling locks them down. Since the final gas composition is more interesting than the composition of gas leaving the reaction chamber, the model has been completed with an additional stage. This stage simulates the evolution of shift and COS formation reactions up to new equilibrium points determined by adjusted equivalent lock temperatures. The model has been implemented using the EES (Engineering Equation Solver) software.

2.3 Model tuning

To tune and validate the model, information provided by the TDG system is used (García-Peña et al., 2000, 2001). This name is an acronym for Thermoeconomic Diagnosis, which is the most important feature of the system. The diagnosis consists of comparing two situations, identifying the causes that determine what particular situation is more efficient or has more electrical production than another, and quantifying how much each one of these causes is responsible for the deviation in efficiency or electrical production.

TDG connects with the plant information system, detects steady-state operating periods and uses mass and energy balances and data reconciliation to calculate the thermodynamic state of the plant. The information from 2874 real operating periods (which means 4812 hours) filtered and processed by this system is used to tune the model: coal conversion values are used to adjust the particle residence time and gas composition is used to adjust equivalent lock temperatures.

To validate the model, the relative average experimental discrepancy (taking into account the history) is calculated for the gasification temperature, the main gas composition, and the CGE (Cold Gas Efficiency, or a quotient between chemical energy of the gas and chemical energy of the fuel), i.e.

$$\text{Discrepancy}(\%) = \frac{1}{p} \sum_{i=1}^{p} \left| \frac{x_{model}^i - x_{plant}^i}{x_{plant}^i} \right| 100 \quad (11)$$

Results are shown in TABLE I. As can be seen, errors in the CGE and the concentrations of most abundant species are very small. Relative

errors of the other gas components are higher because the concentrations of these species are low. In *Figures 3, 4* and *5*, values provided by the model for the CGE and the main gas component compositions are compared to the plant data provided by TDG.

This model can be integrated in the TDG system as an off-design simulator for the gasifier. The model reproduces a large amount of plant data; and, since it simulates all the gasification process, it can take into account the dependence of the fuel conversion ratio with operating conditions. Therefore, when the CGE is plotted versus oxygen and steam ratios, the graph shows a maximum (*Figure 6*).

TABLE I. RELATIVE AVERAGE ERRORS OF THE MODEL

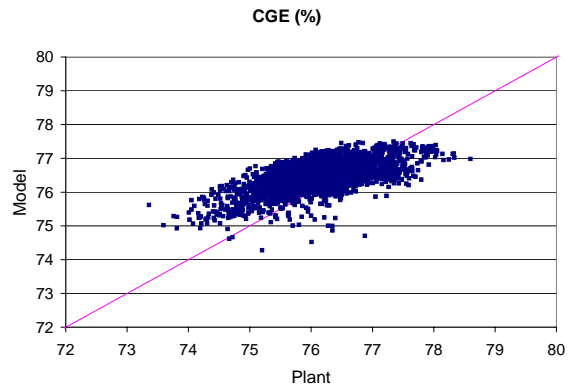| Variable	Experimental discrepancy
Temperature	2.40%
CGE	0.80%
CO	0.56%
H_2	0.43%
CO_2	4.90%
H_2O	5.30%
H_2S	7.90%
COS	8.50%

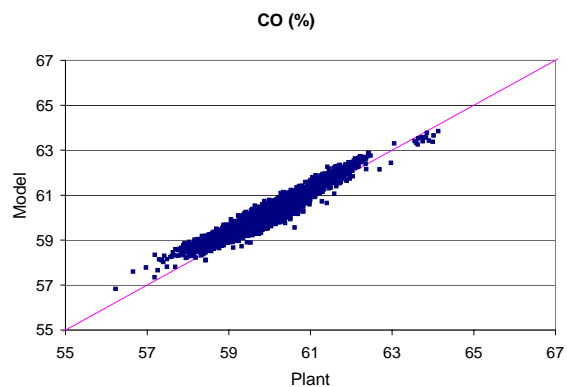*Figure 3. CGE experimental discrepancy.*

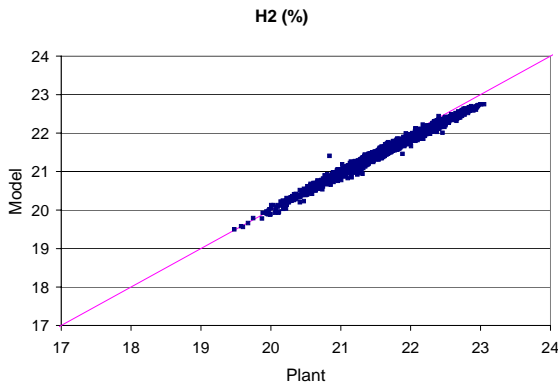*Figure 4. CO experimental discrepancy.*

H2 (%)

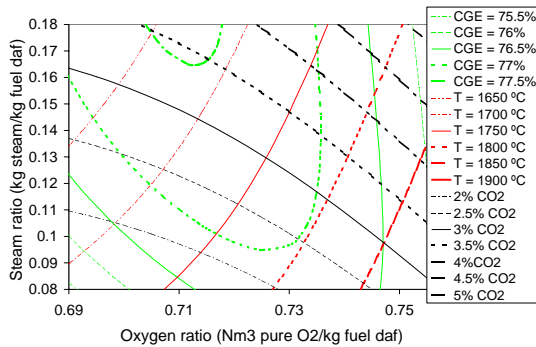*Figure 5. H₂ experimental discrepancy.*

Figure 6. CGE, temperature and CO₂ concentrations versus oxygen and steam ratios.

3. Study of Co-Gasification of Biomass, Coal and Coke

In this section, the model previously presented and validated is used to study co-gasification. However, the model has been validated by the usual fuel (coal and coke at 50 percent by weight) and might not be suitable for mixtures including biomass. The main reason to think that the model is accurate enough to study co-gasification is that the model (and gasification itself) could be divided into two parts: first, most non-mineral matter of the fuel converts into gas (fuel conversion) and, second, gas phase equilibrium determines the distribution of the gasified matter into the different species. Since the gas phase equilibrium does not depend on the origin of the gas (due to high temperatures, tars and oils are not formed), the only difference may appear in the fuel conversion. The gasifier works with a quite high,constant fuel conversion ratio (around 98-99%), so the influence of fuel conversion is less important than differences in fuel composition and LHV and in operation. Besides, biomass has a high volatiles content (which is taken into account in the model) that can provide higher conversion ratios than coal and coke. Synergistic effects in co-gasification of

coal and biomass described in the literature (Sjöström et al. 1999; Jong et al., 1999) could also improve the fuel conversion ratio. In conclusion, the biomass with its high volatile content and a reduced biomass amount will ensure that fuel conversion remains almost constant, making the model useable.

In any case, the work presented here is a preliminary study to determine how the operation should be modified when introducing biomass into a gasifier where this fuel has not been used. Prior to continuous co-gasification in the plant, some tests should be carried out in order i) to determine the suitable particle size for biomass to ensure high conversion and ii) to adjust the tuning of the model if necessary.

3.1 Comparison of fuel characteristics

Prior to analysing the simulation results, it can be very useful to compare the characteristics of the possible fuels: coal, coke, and different types of biomass. This comparison can explain, at least qualitatively, results provided by the model. Information on coal and coke corresponds to the average of plant analysis in May 2001, while biomass data are provided from CIRCE (Cynara Cardunculus) and VTT (Kurkela et al., 2000). Maximum moisture content has been fixed at 15% because this is the upper limit for a right mill operation.

As can be seen in TABLE II, all biomass types are quite similar and have some differences with the other fuels. Biomass has more moisture and volatiles than coal or coke. Its LHV is roughly the same as that of the coal and half of that of the coke. The same can be said about carbon content. The difference between biomass and coal is that the first has high oxygen content and the second high ash content. Biomass has about twice the hydrogen than the other fuels. Last, biomass has low sulphur content, especially if compared with coke.
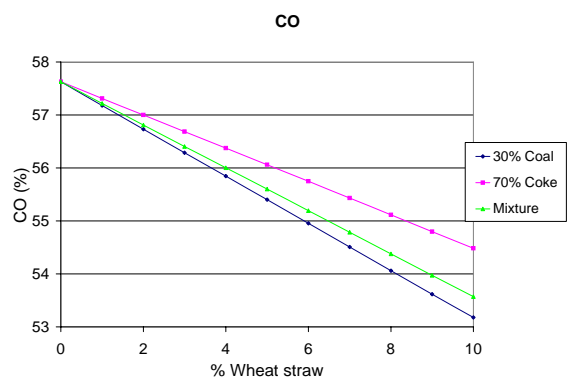
CO

Figure 7. CO versus wheat straw percentage.

TABLE II. FUEL ANALYSIS

	Coal	Coke	Wheat straw	Barley straw	Pine wood	Olive tree	Wine tree	Cynara
Moisture Wt %	2	2	12.1	13.8	8	15	15	8
Proximate analysis wt % d.b.								
Volatile matter	22.1	12.4	73.6	75	76.3	78.1	76.6	76.5
Fixed carbon	31.4	87.0	18.5	19.3	18.1	18.9	20.7	17.7
Ash	46.5	0.58	7.9	5.7	5.6	3.0	2.7	5.8
Ultimate analysis wt % d.b.								
C	40.60	87.70	45.60	45.60	47.20	49.80	49	46.8
H	2.80	3.80	5.70	5.60	5.70	6.00	5.70	5.80
O	8.40	0.19	40	42.50	39.20	40.40	41.80	40.70
N	0.82	1.50	0.70	0.50	2.20	0.70	0.70	0.70
S	0.88	6.20	0.09	0.09	0.09	0.06	0.05	0.13
Ash	46.50	0.58	7.90	5.70	5.60	3.00	2.70	5.90
LHV (kJ/kg, w.b.)								
LHV	15,109	33,228	14,472	14,403	16,365	15,784	15,189	16,041

3.2 Tuning of the gasification operating parameters

To simulate co-gasification, actual operating conditions are considered. The fuel mixture is then modified by changing the amounts of coal and coke and including up to 10 percent of biomass but keeping constant the total chemical energy of the fuel. Two experiments have been carried out. In the first, fuel composition is varied while keeping oxygen and steam flows constant. In the second, fuel composition is also modified, but oxygen and steam are varied to optimise the CGE.

When part of the fuel is replaced by biomass keeping oxygen and steam constant, fuel and gas mass flows increase because the biomass has a lower LHV and less ash than the coal/coke mix. Due to the biomass' high oxygen content, CO_2 and H_2O concentrations increase and CO concentration decreases which implies a reduction in the gas LHV. The gasification temperature slightly decreases, and CGE remains roughly constant. Finally, sulphur decreases; and ash mass flow usually decreases, although they can increase, depending on whether or not the biomass replaces coal, coke, or a mixture of the two.

In *Figure 7*, the evolution of CO concentration versus biomass is shown by using three lines. These lines are plotted by considering a basis coal/coke mixture (30/70 percent in energy, which means about 50/50 percent by weight) and substituting the coke (30 percent coal), coal (70 percent coke), or the mixture of both by wheat straw only. As can be seen, the effect of biomass is higher when coke is replaced.

Effects of fuel modification can be reduced if the operation is changed. One possibility is to adapt the oxygen and steam flows to maximise the CGE. If that is simulated, temperature, gas composition, CGE, and oxygen flow remain almost constant, reducing steam flow at the same time (*Figure 8*). This implies that more steam could be expanded in the turbine obtaining more electrical power (between 1 and 2 MWe in this case). It should be noted that maximisation of the CGE can only be done by using a non-linear model with a variable conversion ratio such as the one proposed here.

Although a 50/50 percent by weight of coal and coke is the usual fuel mixture, other combinations have been explored. In Figure 9, the amount of steam to maximise the CGE when the coal of several coal/coke mixtures is substituted by wheat straw is plotted. As can be seen, the lower the coke content is, the higher the saving in steam flow. Similar graphs have been built by substituting the coke or the mixture by biomass only.

Although the steam mass flow has been plotted, during operation, oxygen and steam ratios (obtained by dividing these mass flows into the

mass flow fuel without ashes and moisture) are used. Here, absolute mass flow has been used to separate the effects of steam mass flow and fuel mass flow variations. Since the biomass LHV on a dry and ash free basis is lower than that of the coal or coke, lines would have dropped more quickly if ratios had been used,. Results obtained by using all biomass types are quite similar and show that co-gasification is possible provided that the operation is modified. Since maximising the CGE is not the only operating strategy, how the temperature, CGE, and gas composition would vary if steam or oxygen were modified, keeping the fuel constant has also been studied.
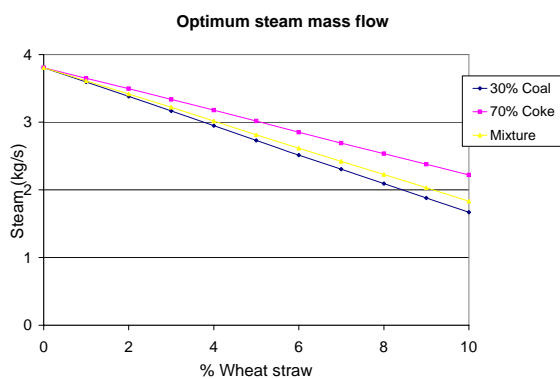

Optimum steam mass flow

Figure 8. Steam mass flow to maximise the CGE.
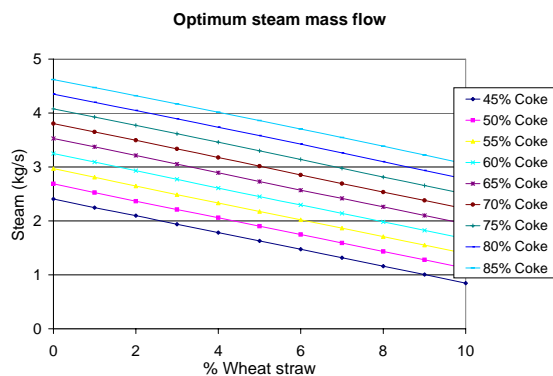

Optimum steam mass flow

Figure 9. Steam mass flow to maximise the CGE (replacing coal by biomass).

4. Conclusions

Due to high temperatures, which avoid tar and oil formation, and the efficient gas cleaning section, co-gasification at a large IGCC power plant is the most interesting way to use a wide range of fuels (not only biomass but also solid waste, used tires, etc.) with high efficiency and a negligible environmental impact.

A gasifier model has been described and validated with nearby 3,000 actual plant steady state operating periods. This adjusted model can be used to simulate gasifier operation with new

fuels, optimise operation, and understand how a gasifier works. The model has also been used to study co-gasification of coal, coke, and several types of biomass. Simulations show that operating strategies should be modified as the fuel mixture varies, so complete plant operatons maps for a wide range of coke/coal/biomass combinations have been built.

When a new fuel mixture is proposed, fuel, ash and sulphur mass flows should first of all be below limits determined by fuel feeding, ash removal and gas cleaning systems. If biomass is used, sulphur and ashes decrease, so that the coal/coke relation could vary over a wider range. Second, the model should be used to determine the correct operating set point (by building graphs for coke/coal/biomass mixtures). A small amount of biomass (up to 10 percent) implies that the steam amount can be decreased substantially. Additionally, although properties of most types of biomass are quite similar, moisture content variations can modify the optimal steam and oxygen ratios. Last but not least, there are other fuel composition aspects that do not affect the operating set point but should be taken into account. First, some types of biomass, such as straw, contain chlorine that could produce corrosion. Second, biomass contains elements like calcium or potassium that reduce the ashes' melting point, which implies that the limestone ratio should be reduced.

Nomenclature

a_c O_2 coefficient in the combustion equation
b_c CO coefficient in the combustion equation
c_c CO_2 coefficient in the combustion equation
CGE   Cold gas efficiency (%)
d.b.   Dry basis
d_c H_2O coefficient in the combustion equation
e_c H_2S coefficient in the combustion equation
f_c N_2 coefficient in the combustion equation
h   Hydrogen subscript in the char formula
hf   Hydrogen subscript in the fuel formula
h_{gas} External convection coefficient (kJ/(m²·s))
HRSG Heat recovery steam generator
k_t Specific total reaction rate (kJ/(m²·s))

k_{ash}	Specific diffusion rate of the gas through the ash (kJ/(m·s))
k_q	Chemical reaction rate constant (kJ/(m²·s))
LHV	Lower heating value (kJ/kg)
n	Nitrogen subscript in the char formula
nf	Nitrogen subscript in the fuel formula
o	Oxygen subscript in the char formula
of	Oxygen subscript in the fuel formula
p	Number of points
r_c	Core radius (m)
r_p	Particle radius (m)
s	Sulphur subscript in the char formula
sf	Sulphur subscript in the fuel formula
w	Moisture subscript in the fuel formula
w.b.	Wet basis
wt	Weight
x	Generic variable
y	Radius relation

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