

## Modelling system efficiencies and costs of two biomass-fuelled SOFC systems

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### Abstract

Increasing demand for power and the depletion of fossil fuels are providing opportunities for the development of fuel cells as power generating systems. This paper investigates the integration of a solid oxide fuel cell (SOFC) with biomass gasification for the production of power and heat (combined heat and power (CHP) system). A steady-state model was developed in the gPROMS modelling tool to investigate the integrated system.

The system was modelled for two different options, a cold process involving gas cleaning at a reduced temperature and a hot process involving gas cleaning at a high temperature.

For each option, the model was used to determine the system efficiency and prospective costs. The electrical efficiency and overall system efficiency for the hot process were found to be 23 and 60% and for the cold process the efficiencies were 21 and 34%, respectively. Superior heat management in the gas cleaning stage of the hot process results in its higher system efficiency. The capital cost for the hot process appears higher than that for the cold process. This differential capital cost may be justified by the income earned from selling the extra heat produced in the hot process. Conversely, the cold process produces a gas stream with lower levels of impurities than the hot process.

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**Keywords:** Fuel cells; SOFC; Biomass fuel; System efficiency

### 1. Introduction

The pressure to reduce the impact on the environment of conventional energy conversion systems used for the generation of power and heat is increasing. The major areas of concern are greenhouse gas emissions and air pollution, in conjunction with the possible depletion of fossil fuels. However, energy demand is projected to increase and small-distributed power systems are attracting increased interest.

These issues have led to the exploration of alternative energy conversion systems. Fuel cell technology is one such alternative, with the solid oxide fuel cell (SOFC) offering bright prospects. Advantages of fuel cells include low to zero emissions during operation, flexibility of operation and ease of integration with other systems. However, fuel cells are not yet used commercially. Another area of growth in

alternative energy systems is the increased use of biomass, grown in a sustainable manner, as a replacement for fossil fuels [11].

This paper explores the possibility of combining two technologies, SOFC and biomass gasification, for the generation of power and heat. A steady state model for a 200 kW<sub>e</sub> SOFC combined heat and power (CHP) system is developed using the gPROMS modelling tool [6]. Two different systems are modelled; the first involves a hot gas cleanup process while the other uses a cold gas cleanup process. The models determine system efficiency and preliminary costs for the two systems.

The hot gas cleanup process involves treatment of the producer gas at a high temperature with little or no cooling. It offers good heat management since the gasifier and the fuel cell operate at similar temperatures. However, equipment for gas treatment at such temperatures is not yet reliable and is expensive. Cold gas cleanup involves significant cooling of the producer gas before and during the gas treatment. Equipment used for cold gas cleanup is well developed and readily available, though much of

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the heat energy in the exhaust cannot be used under these circumstances.

## 2. Fuel cells

Fuel cells directly convert the chemical energy of a fuel into electrical energy at a constant temperature [3,9], via electrochemical reactions. They consist of two electrodes (the anode and the cathode) separated by an electrolyte. The electrodes are electronic conductors, with electrons flowing from the anode to the cathode. The electrolyte is an ionic conductor.

The SOFC uses oxide ion-conducting ceramic as the electrolyte, which becomes a conductor of oxygen ions at temperatures of about 800 °C [9]. It is simpler in concept than some other fuel cell types since there are no problems associated with electrolyte management. On the other hand, the high operating temperature of a SOFC makes it difficult to find materials for its construction with the necessary physical and thermal properties.

## 3. Biomass gasification

### 3.1. Introduction

The gasification of biomass results in the production of a gaseous fuel with a calorific value of about 4–20 MJ N<sup>-1</sup> m<sup>-3</sup> [1]. Biomass undergoes thermal decomposition at high temperatures (600–1000 °C) to form gaseous components like carbon monoxide and hydrogen as well as impurities such as particulates and tars. The biomass is generally treated prior to its introduction into the gasifier [5]. This involves drying and proper sizing of the feedstock before feeding to the gasifier. The characteristic of the gas produced varies with the type of fuel material and gasifier type and conditions. The product gas from the gasification process can be used (after cleaning and conditioning) in turbines, boilers and engines for the generation of heat and electricity and for the production of synthesis gases for the manufacture of fuels and chemicals, or hydrogen for fuel cells.

### 3.2. Gas cleanup

The raw product gas from a biomass gasifier contains impurities that may interfere with downstream utilisation technologies such as fuel cells [14]. The end-use application of the gas determines the extent to which it is cleaned, and the design of the cleanup system influences the level and type of impurities going into downstream power generation equipment. Some applications such as kilns require little or no cleanup, while others may need extensive gas cleanup to meet stringent fuel quality requirements. Generally, the types of contaminants and their amounts depend on the gasification process, type of gasifier used and the type of biomass feedstock.

### 3.2.1. Types of impurities

The major contaminants produced during gasification are particulates, alkali compounds, tars, nitrogen-containing compounds and sulphur [14].

Nitrogen-containing components and sulphur contaminants are generally a minimal problem in biomass gasification. Most biomass feedstocks contain low percentages of sulphur, while the production of NO<sub>x</sub>, the major nitrogen contaminant, is limited by the nature of the reactive environment.

### 3.2.2. Gas cleanup technologies

The choice and design of gas cleanup technology is affected in a number of ways by the nature of the contaminant. The primary types of systems for particulate removal include cyclone filters, barrier filters, electrostatic filters, and wet scrubbers [14]. Alkali vapours require more complex treatment and are removed by cooling the hot product gas below 600 °C to allow for condensation of the primary material into solid particulates. The solids are then removed using filtration systems similar to those for particulates. Alternatively, the use of alkali traps may be effective at removing alkali from the hot gas stream [14]. Methods to remove tars from producer gases fall into one of three categories [5]: physical removal, thermal conversion, and catalytic destruction.

### 3.3. Tolerance levels of impurities

The detailed fuel gas quality specifications for fuel cells are not well understood and documented. It is therefore difficult to specify tolerance limits for impurities. The temperature of operation has an impact on tolerance levels, with high temperature fuel cells like SOFC able to cope with higher levels of impurities than low temperature fuel cells. For example, tar decomposition may occur at the anode of the SOFC rendering it comparatively innocuous, but the effect of carbon deposition on catalyst surfaces might be a problem.

As an example, Table 1 lists the fuel quality requirements for internal combustion engines and gas turbines. For molten carbonate fuel cells (MCFC), Lobachyov and Richter [10] suggest a tolerance range of 1–10 ppm for alkali compounds and a limit of 100 ppm for particles greater than 3 μm diameter.

Table 1  
Gas quality for internal combustion (IC) engines and turbine generators [6]

	IC engine	Gas turbine
Particles (mg N <sup>-1</sup> m <sup>-3</sup> )	<50	<30
Particle size (μm)	<10	<5
Tar (mg N <sup>-1</sup> m <sup>-3</sup> )	<100	
Alkali metals (mg N <sup>-1</sup> m <sup>-3</sup> )		<0.25

## 4. System configuration and components

### 4.1. Introduction

A CHP system is used to generate both electricity and heat. The CHP system in this case is based around a fuel cell stack, and requires the integration of many components beyond the fuel cell stack itself.

For the two options (cold gas cleanup and hot gas cleanup) considered in this work, the system has been configured into three subsystems (Fig. 1).

- Gasification: biomass feed is gasified to a raw gaseous product.
- Gas cleanup: the raw producer gas from the gasifier is cleaned and conditioned.
- Fuel cell: cleaned fuel is reformed and oxidised, generating electricity and releasing heat.

The objective in building a model is to assess component and system characteristics based on different equipment options for each section of the system.

### 4.2. Equipment selection

#### 4.2.1. Gasification subsystem

The biomass feed is dried and gasified using air. It is assumed that the biomass is properly sized when supplied.

- System 1: cold gas cleanup
 

Most commercial gasifiers of the scale considered (200 kW<sub>e</sub>, 600 kW<sub>th</sub>) are fixed bed gasifiers. A co-current fixed bed gasifier suits the cold process better than a counter-current one, because it produces less tar (the most problematic of the impurities) and has a higher product gas temperature [8]. The higher product gas temperature increases the amount of heat recovered in the cooling process. Tar levels are then reduced to acceptable levels using a wet electrostatic precipitator.

- System 2: hot gas cleanup
 

The gasifier proposed for the hot gas process is a fluidised bed gasifier, with the catalyst embedded in the bed material. This eliminates the use of a second and expensive downstream reactor for the tar cracking process. Catalytic tar cracking takes place in the fluidised bed reactor and requires operating temperatures of about 900 °C [7].

#### 4.2.2. Cleanup subsystem

The impurities associated with the product gas are removed in the cleanup subsystem.

- System 1: cold gas cleanup

The best option for cold gas cleanup is to separate the particulates and alkali compounds first, before condensation and removal of tars. This is because the particulates and tars agglomerate, forming sticky particles that clog filters. Cyclones and filters are used for the particulates and alkali compounds and a wet electrostatic precipitator is used for tar removal. Cyclones help in removing large particulates (above 5 μm) and reduce the load burden on the filter. The gas stream is cooled to about 200 °C after the cyclone stage to allow for the condensation of alkali compounds. The tars remain in the vapour phase. A bag filter is used for the filtration of the condensed alkali compounds and particulates. A wet electrostatic precipitator is selected for tar removal because of its reliability over wet scrubbers for small-scale systems [14].

- System 2: hot gas cleanup

Since it is assumed that sufficient tar destruction occurs in the gasifier, the relevant impurities are particulates and alkali compounds. The gas stream leaving the fluidised bed gasifier passes through a cyclone and is cooled to allow for alkali condensation. Ceramic filters, capable of filtration for temperatures as high as 700 °C, are used for the removal of alkali compounds and remaining particulates.

#### 4.2.3. Fuel cell subsystem

The equipment in the fuel cell subsystem is almost the same for both processes. The only difference is that additional heat is needed for the cold gas process to raise the temperature of the gas stream entering the SOFC to 850 °C.

### 4.3. Process description

#### 4.3.1. System 1: cold gas process

The flowsheet for the cold gas process is shown in Fig. 2.

- Gasification subsystem

The sized biomass is dried to reduce its water content from about 50% (by weight) to 20% before being fed to the gasifier at 100 °C. The gasifier operating temperature is assumed to about be 600 °C. The gasifier uses air pre-heated to approximately 400 °C as the oxidising agent.

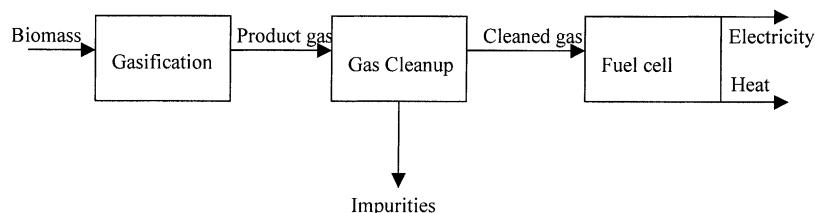


Fig. 1. Overall system configuration.



is maintained at 900 °C by cooling with air supplied in excess of that required for the reaction. The cathode and the anode exit streams are used for heating the respective streams entering the cathode and anode. These exit streams are then introduced into the combustor where the unreacted fuel in the cathode exit stream is oxidised. The flue gas stream is used for district heating.

#### 4.3.2. System 2: hot gas process

The flowsheet for the hot gas process is shown in Fig. 3.

- Gasification subsystem

The drying stage in the hot gas gasification subsystem is similar to that for the cold process. The gasifier operating temperature is assumed to be about 900 °C, using air pre-heated to about 400 °C for oxidation and fluidisation.

- Cleanup subsystem

The hot gas exiting the gasifier at about 900 °C enters a cyclone for particle particulate removal. The gas stream leaving the cyclone is cooled in a heat exchanger to about 450 °C before entering the ceramic filter, to allow for condensation of the alkali compounds. The incoming stream to the filter heats up the gas leaving the ceramic filter, before it enters the SOFC at 850 °C.

- Fuel cell subsystem

The hot gas cleanup process retains the steam from the gasification section so that less fresh steam needs to be added to the fuel cell than in the cold gas cleanup process. The gas stream from the cleanup section is introduced along with make-up steam to the anode side of the SOFC. The SOFC operates in the same way as for the cold gas process.

## 5. System and equipment modelling

### 5.1. Introduction

In this section, a steady-state one-dimensional model is presented for predicting the efficiency of the overall system for the two options considered. In addition, the model developed is used for basic cost analysis.

### 5.2. Hierarchical sub-model decomposition

gPROMS [6] is a general process modelling system with proven capabilities for the steady-state and dynamic simulation and optimisation of highly complex processes. It can also be used for parameter estimation. gPROMS allows for the development of generic models in isolation that can simply be defined in terms of explicitly declared variables and equations. Once these models are developed and tested, they can be connected in an appropriate way to construct the flowsheet model, in a process called hierarchical sub-model decomposition.

### 5.3. System modelling

The system model is decomposed into three subsystems: gasification, cleanup, and fuel cell.

#### 5.3.1. System 1: cold gas process

The subsystems in the process contain the following units.

- Gasification: dryer, blower and co-current fixed bed gasifier.
- Cleanup: cyclone, heat exchanger, filter and wet precipitator.
- Fuel cell: SOFC, two heat exchangers, blower and combustor.

#### 5.3.2. System 2: hot gas process

The subsystems in the process contain the following units.

- Gasification: dryer, blower and fluidised bed gasifier.
- Cleanup: cyclone, heat exchanger and ceramic filter.
- Fuel cell: SOFC, heat exchanger, blower and combustor.

#### 5.3.3. Overall model

The unit models in each subsystem are connected to form subsystem models, and these models connected to obtain the overall models for both processes. The models are then used to obtain performance measures. In particular, the overall system model is used to determine the electrical and overall efficiency, by considering the biomass fuel input and the output power and heat.

The heat available from the system,  $Q_{DH}$ , is given by

$$Q_{DH} = F_e C_{pe} \times (T_{Ce} - 373) \quad (1)$$

where  $F_e$  is the flow-rate of the combustor exit stream,  $C_{pe}$  is the heat capacity of the stream in  $\text{kJ mol}^{-1} \text{K}^{-1}$  and  $T_{Ce}$  its temperature in K.

The total power and heat available from the system,  $Q_T$ , is given by

$$Q_T = Q_{DH} + 200 \text{ kW}_e \quad (2)$$

Using these quantities, the system efficiency,  $\eta_{sys}$ , is defined as the ratio of useful power and heat available from the system to the biomass fuel input to the system

$$\eta_{sys} = \frac{Q_T - Q_D}{M_{biomass} \times \text{LHV}} \quad (3)$$

where  $Q_D$  is the total power demand on the system in kW,  $M_{biomass}$  is the flow-rate of the biomass in  $\text{kg s}^{-1}$  and LHV is the low heating value of the biomass in  $\text{kJ kg}^{-1}$ .

Similarly, the electrical efficiency,  $\eta_{elec}$ , is defined as the ratio of useful power available from the system to the biomass fuel input to the system

$$\eta_{elec} = \frac{200 - Q_D}{M_{biomass} \times \text{LHV}} \quad (4)$$



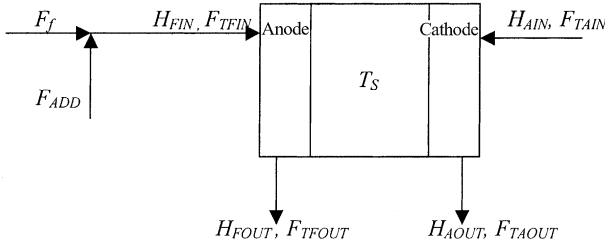


Fig. 4. Diagram of SOFC.

#### 5.4. Equipment modelling

The models derived for the equipment in both processes are presented in this section. It is assumed throughout that pressure drops and heat losses are negligible.

##### 5.4.1. Fuel cell

In this section, simple mass and energy balances for a planar internal-reforming SOFC (Fig. 4) are developed. The main assumptions made in modelling the SOFC are:

- Only hydrogen is oxidised in the SOFC. Methane is reformed to carbon monoxide, which is then shifted to carbon dioxide and hydrogen in the presence of steam.
- The inlet and exit temperatures of fuel and air streams are known. There is a temperature gradient of 100 °C across the SOFC.
- The temperature of the fuel cell structure,  $T_S$ , is homogeneous and midway between the inlet and exit temperatures.
- The fuel utilisation factor is fixed.
- The operating voltage is known and constant over the cell.
- There are six components in the fuel stream entering and exiting the SOFC, defined by the set  $NCF = \{H_2, CO, CO_2, CH_4, H_2O, N_2\}$ . There are two components in the air stream, defined by the set  $NCA = \{O_2, N_2\}$ .

The mass balances for the fuel cell subsystem are first derived. The power rating,  $P_{we}$ , of the fuel cell (200 kW<sub>e</sub>) is the product of the cell operating voltage  $V$ , the total current  $i$  and the inverter efficiency  $\eta_{IN}$

$$P_{we} = iV\eta_{IN} \quad (5)$$

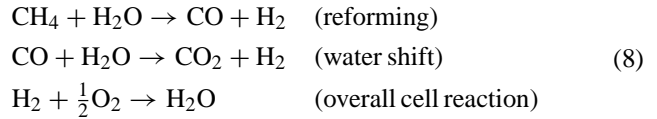
The amount of hydrogen consumed,  $r_b$  (mol s<sup>-1</sup>), is related to the current by Faraday's law

$$r_b = \frac{i}{n \times Fa} \quad (6)$$

For a known fuel utilisation factor,  $U_f$ , the amount of hydrogen supplied,  $H_{2sup}$  (mol s<sup>-1</sup>), is given by

$$H_{2sup} = \frac{r_b}{U_f} \quad (7)$$

The reactions taking place in the SOFC are:



The flow-rate of the fuel stream,  $F_f$  (mol s<sup>-1</sup>) needed to produce the required amount of hydrogen is

$$F_f = \frac{H_{2sup}}{X_{FC}} \quad (9)$$

where  $X_{FC}$  is the number of moles produced by 1 mol of fuel of known composition. For, a known fuel gas composition  $X_i$ , the individual flow-rates are:

$$F_f(i) = F_f \times X(i) \quad \text{for all } i \in NCF \quad (10)$$

An amount of steam equivalent to twice the amount needed for the reforming and water shift reactions is supplied in order to avoid carbon deposition. According to equation [8], the flow-rate of steam needed,  $F_{H_2O}$  (mol s<sup>-1</sup>) is thus

$$F_{H_2O} = (F_f(CO) + F_f(CH_4) \times 2) \times 2 \quad (11)$$

The flow-rate of additional steam supplied,  $F_{ADD}$  (mol s<sup>-1</sup>) is thus given by

$$F_{ADD} = F_{H_2O} - F_f(H_2O) \quad (12)$$

Total flow-rate of the fuel stream entering the SOFC,  $F_{Tfin}$  (mol s<sup>-1</sup>) is

$$F_{Tfin} = F_f + F_{H_2O} \quad (13)$$

For known conversions for the reforming and water shift reactions, the component flow-rates of the fuel exit stream are given as

$$F_{fout}(H_2) = H_{2sup} \times (1 - U_f) \quad (14)$$

$$F_{fout}(CO) = 0 \quad (15)$$

$$F_{fout}(CO_2) = F_f(CO) + F_f(CO_2) + F_f(CH_4) \quad (16)$$

$$F_{fout}(CH_4) = 0 \quad (17)$$

$$F_{fout}(H_2O) = (F_{H_2O} \times 0.5) + (H_{2sup} \times U_f) \quad (18)$$

$$F_{fout}(N_2) = F_f(N_2) \quad (19)$$

For the exit air stream, given known inlet composition, the flow-rates are:

$$F_{aout}(O_2) = F_{Tain} \times 0.21 - 0.5 \times U_f \times H_{2sup} \quad (20)$$

$$F_{aout}(N_2) = F_{Tain} \times 0.79 \quad (21)$$

The energy balance for the fuel cell is now obtained. Considering adiabatic conditions, the enthalpy change between the inlet and outlet stream should be equal to the work  $W$  produced by the fuel cell. Though heat is generated, no heat term is involved in the energy balance, because the air supplied for cooling and reaction has been incorporated in the

enthalpy terms (expressed in kW). The flow-rate of the air stream is obtained from a solution of the energy balance.

The balance around the SOFC is given by

$$(H_{\text{fin}} + H_{\text{ain}}) - (H_{\text{fout}} + H_{\text{aout}}) = W \quad (22)$$

where  $H$  denotes enthalpy (kW), the subscript ‘f’ refers to the fuel stream, the subscript ‘a’ to the air stream, the subscript ‘in’ to the fuel cell inlet and the subscript ‘out’ the fuel cell outlet. Moreover, the enthalpy of a stream is given by

$$H = \sum (F_i \times \Delta H_{\text{ft}}^{\circ}) + \sum F_i (h_i^T - h_i^{T_{\text{ref}}}) \quad (23)$$

where the summations are over all components  $i$  in the stream,  $\Delta H_{\text{ft}}^{\circ}$  denotes the standard heat of formation of component  $i$  (kJ mol<sup>-1</sup>) at temperature  $T_{\text{ref}}$  (K) and where

$$h_i^T - h_i^{T_{\text{ref}}} = \int_{T_{\text{ref}}}^{T_s} C_{pi} dT \quad (24)$$

where  $C_{pi}$  is the heat capacity of component  $i$  (kJ mol<sup>-1</sup> K<sup>-1</sup>).

#### 5.4.2. Combustor

A model is now developed to determine the flow-rate and temperature of the combustor exit stream, given known values for the inlet streams (Fig. 5). The key assumptions in the model are as follows:

- There are seven components in the combustor exit stream, denoted by the set NCE = {H<sub>2</sub>, CO, CO<sub>2</sub>, CH<sub>4</sub>, H<sub>2</sub>O, N<sub>2</sub>, O<sub>2</sub>}.
- Hydrogen is completely oxidised in the combustor.

The mass balances for the combustor are first derived. The un-reacted hydrogen in the fuel stream is oxidised by oxygen in the combustor to produce water



The component flow-rates for the combustor exit stream are given by

$$F_e(\text{H}_2) = 0 \quad (26)$$

$$F_e(\text{CO}) = 0 \quad (27)$$

$$F_e(\text{CO}_2) = F_f(\text{CO}_2) \quad (28)$$

$$F_e(\text{CH}_4) = 0 \quad (29)$$

$$F_e(\text{H}_2\text{O}) = F_f(\text{H}_2\text{O}) + F_f(\text{H}_2) \quad (30)$$

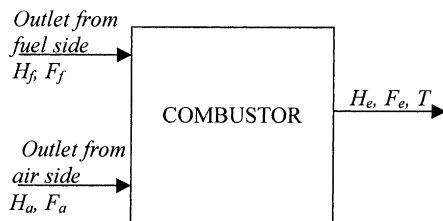


Fig. 5. Diagram of combustor.

$$F_e(\text{N}_2) = F_f(\text{N}_2) + F_a(\text{N}_2) \quad (31)$$

$$F_e(\text{O}_2) = F_a(\text{O}_2) - 0.5 \times F_f(\text{H}_2) \quad (32)$$

Given all flow-rates, the temperature of the exit stream can be calculated from the energy balance

$$H_f + H_a = H_e \quad (33)$$

where  $H_j$  denotes the enthalpy of stream  $j$  (kW).

#### 5.4.3. Gasifier

The gasifier model determines the flow-rates of the wet biomass and air stream for a known product gas composition. The main assumptions are:

- The gasifier operates at a known temperature and at atmospheric pressure.
- Tars, particulates and alkali compounds are the only impurities produced during gasification.

The mass balances are first derived. The flow-rate of carbon (mol s<sup>-1</sup>) entering the gasifier equals that exiting the gasifier

$$F_{\text{CIN}} = F_{\text{COUT}} \quad (34)$$

The flow out of the gasifier is a function of the total carbon content of the producer gas (CO, CO<sub>2</sub>, CH<sub>4</sub>,) and the carbon conversion efficiency of the gasifier  $\eta_{\text{CG}}$

$$F_{\text{COUT}} = \frac{F_{\text{PG}}(\text{CO}) + F_{\text{PG}}(\text{CO}_2) + F_{\text{PG}}(\text{CH}_4)}{\eta_{\text{CG}}} \quad (35)$$

The flow-rate of the dry biomass,  $M_{\text{tbf}}$  (g s<sup>-1</sup>), is

$$M_{\text{tbf}} = \frac{F_{\text{CIN}} \times \text{RMM}_{\text{C}}}{X_{\text{bc}}} \quad (36)$$

where  $\text{RMM}_{\text{C}}$  is the relative molecular mass of carbon in g mol<sup>-1</sup> and  $X_{\text{bc}}$  is the mass fraction of carbon in the dry biomass.

The flow-rate of air needed for gasification,  $F_{\text{Tair}}$  (mol s<sup>-1</sup>), for a known air–biomass ratio,  $W_{\text{ab}}$ , is given by

$$F_{\text{Tair}} = \frac{W_{\text{ab}} \times M_{\text{tbf}}}{\text{RMM}_{\text{air}}} \quad (37)$$

where  $\text{RMM}_{\text{air}}$  is the relative molecular mass of air in g mol<sup>-1</sup>.

#### 5.4.4. Gas cleaning equipment

The models for the gas cleaning equipment (cyclone, filter and wet electrostatic precipitator) are all similar. Given known inlet impurity concentrations  $C_{\text{IN}}$  in g m<sup>-3</sup>, and equipment cleaning efficiencies obtained from the literature, the impurity concentrations in the exit stream can be calculated. The cleaning efficiency represents the fraction of each impurity that the equipment removes from a gas stream entering it. For example, a ceramic filter with 99% efficiency removes 99% of the alkali compounds and particulate from the gas stream flowing through it. The modelling assumptions are as follows:

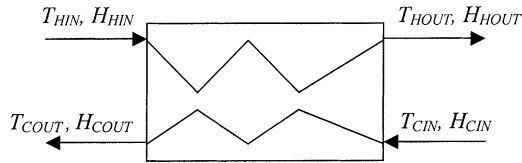


Fig. 6. Diagram of heat exchanger model.

- The flow-rate of the producer gas passes through the cleaning section unchanged.
- The heat capacity of the producer gas is constant throughout the cleaning section

The mass balances for the gas cleaning units over all impurities  $i$  are first derived

$$C_{IN}(i) - C_{IMP}(i) = C_{OUT}(i) \quad (38)$$

$$C_{IMP}(i) = \eta_{CI} \times C_{IN}(i) \quad (39)$$

where  $\eta_{CI}$  is the equipment cleaning efficiency.

#### 5.4.5. Heat exchanger

A simple heat exchanger model (Fig. 6) is used to derive the area required for heat transfer for a known heat load  $Q$  (kW). The key assumptions are as follows:

- Heat exchanger efficiency and approach temperature are known.
- Three of the four stage temperatures ( $T_{HIN}$ ,  $T_{HOUT}$ ,  $T_{CIN}$ ,  $T_{COUT}$ ) are fixed
- The mass flow-rates of all streams are known.

The heat load is directly related to the enthalpy change for the hot fluid:

$$Q = H_{HIN} - H_{HOUT} \quad (40)$$

$$Q = U \times \text{Area} \times \theta_m \quad (41)$$

where  $H_{HIN}$  (kW) is the enthalpy of the inlet hot fluid,  $H_{HOUT}$  (kW) is the enthalpy of the outlet hot fluid,  $U$  ( $\text{kW m}^{-2} \text{K}^{-1}$ ) is the overall heat transfer coefficient,  $\theta_m$  is the logarithmic mean temperature difference (K), Area is the heat exchange area ( $\text{m}^2$ ).

#### 5.4.6. Blower

A blower model is derived to determine the power required for blowing in the gasification and fuel cell sections. It is assumed that the discharge pressure is known and that heat losses are negligible. The power required is a function of the total discharge pressure,  $\Delta P_r$ , and the volumetric flow-rate delivered by the blower,  $Q_v$

$$P_{wblower} = Q_v \times \Delta P_r \quad (42)$$

where  $Q_v$  ( $\text{m}^3 \text{s}^{-1}$ ) is given as

$$Q_v = F_{\text{Tair}} \times V_{\text{mol}} \quad (43)$$

where  $V_{\text{mol}}$  is the volume of a mole of air and  $F_{\text{Tair}}$  is the flow-rate of air in  $\text{mol s}^{-1}$ .

#### 5.4.7. Dryer

The dryer model determines the amount of energy needed to reduce the water content of the biomass to an acceptable level of 20% by weight. It is assumed that the wet biomass contains 50% by weight water and enters the dryer at  $25^\circ\text{C}$ , that heat losses are negligible and that thermal equilibrium is achieved in the wet biomass. The dryer mass balance is given by

$$M_{\text{FEED}} = 0.625 M_{\text{IN}} \quad (44)$$

where  $M_{\text{IN}}$  ( $\text{kg s}^{-1}$ ) is the total flow-rate entering the dryer, and  $M_{\text{FEED}}$  is the total flow-rate leaving the dryer and fed to the gasifier ( $\text{kg s}^{-1}$ ). The amount of dry biomass is

$$M_{\text{biomass}} = 0.8 \times M_{\text{FEED}} \quad (45)$$

The dryer energy balance gives an estimate of the heat required by this unit. To raise the water in the wet biomass to  $100^\circ\text{C}$ ,  $Q_{\text{H}_2\text{O}}$  (kW) is needed:

$$Q_{\text{H}_2\text{O}} = 0.5 \times M_{\text{IN}} \times C_{p\text{H}_2\text{O}} \times (373 - 298) \quad (46)$$

where  $C_{p,\text{H}_2\text{O}}$  ( $\text{kJ kg}^{-1} \text{K}^{-1}$ ) is the heat capacity of water.

The heat required to raise the temperature of the dry biomass to  $100^\circ\text{C}$  is

$$Q_{\text{biomass}} = 0.5 \times M_{\text{IN}} \times C_{p\text{biomass}} \times (373 - 298) \quad (47)$$

where  $C_{p\text{biomass}}$  ( $\text{kJ kg}^{-1} \text{K}^{-1}$ ) is the heat capacity of the dry biomass. The heat required to vaporise the water is

$$Q_{\text{vap}} = 0.3 H_{\text{VH}_2\text{O}} M_{\text{IN}} \quad (48)$$

where  $H_{\text{VH}_2\text{O}}$  is the heat of vaporisation of water in  $\text{kJ kg}^{-1}$ . Therefore, the total heat supplied is

$$Q_{\text{tot}} = Q_{\text{H}_2\text{O}} + Q_{\text{biomass}} + Q_{\text{vap}}$$

## 6. Results and discussion

### 6.1. Results

In this section, the results obtained from modelling are presented for both systems studied. Table 2 shows the composition of the producer gas used as input to the model. Tables 3 and 4 summarise the results obtained from the simulation of both processes assuming a 50% fuel utilisation factor. Table 5 shows an estimate of the capital cost for the

Table 2  
Gas composition (wet) after gasification

Component	vol. %
H <sub>2</sub>	17
CO	13
CO <sub>2</sub>	11
CH <sub>4</sub>	4
H <sub>2</sub> O	15
N <sub>2</sub>	40



Table 3  
Results for gas cleanup section

Variable	Symbol (units)	Cold process	Hot process
Level of impurity entering SOFC	$C_{\text{outfilter}}$ ( $\text{g N m}^{-3}$ )	$5 \times 10^{-5}$ (alkali), $2.5 \times 10^{-5}$ (particulate), 0.005 (tars)	0.05 (alkali), 0.05 (particulate), 5 (tars)
Area of heat exchanger	Area ( $\text{m}^2$ )	33	47

Table 4  
Efficiency estimation

	Specific power or demand (kW)	
	Cold process	Hot process
AC power output SOFR	200.0	200.0
Heat output from district heating	77.1	196.8
Total output	277.1	396.8
Dryer electric demand	10.8	9.7
Blower electric demand (SOFC)	50.7	50.5
Blower electric demand (gasifier)	1.4	1.8
Pre-heated gasifier air electric demand	14.7	18.1
Total electric demand	77.6	80.1
Net electricity production	122.1	119.9
Net total power and heat production	199.5	316.7
LHV dry biomass	588.2	531.5
Electrical efficiency (%)	20.8	22.6
System efficiency (%)	33.9	59.6

two systems. The capital cost includes the total purchase equipment cost and estimates of other costs for functions like piping, instrumentation, control, assembly, installation and commissioning [13]. The capital cost of the hot process is found to be approximately 7% higher than that of the cold process.

It is interesting to consider whether the extra cost of the hot process is justified by considering potential revenue from the sale of excess heat generated. Tables 6 and 7 show the

Table 5  
Cost calculations for both 200 kW<sub>e</sub> systems studied

Equipment/process	Cost (£k)	
	Cold process	Hot process
Fuel handling and preparation [1]	11.6	11.6
Gasifier [1]	39.2	90.6
Blowers	8.9	13.8
Cyclone [1]	3.1	3.1
Filter	9.6	45.5
Wet precipitator [6]	68.7	
Heat exchangers [12]	28.0	27.0
Piping, pumps, instrumentation, control and other peripheral equipment	32.5	32.5
Assembly, installation and commissioning	15.5	15.5
SOFC (£500 kW <sub>e</sub> <sup>-1</sup> )	100.0	100.0
Combustor [8]	6.1	6.1
Total system cost	323.1	345.8
Total unit capacity cost (£ kW <sub>e</sub> <sup>-1</sup> )	2.6	2.9
Total unit capacity cost (£ kW <sub>T</sub> <sup>-1</sup> )	1.6	1.1

Table 6  
Heat production for the hot and cold process

Heat produced by hot process (kW)	197
Heat produced by cold process (kW)	77
Difference (kW)	120

Table 7  
Capital cost differences

Capital cost of hot process (£)	346000
Capital cost of cold process (£)	323000
Difference (£)	23000
Annualised difference (£)	2900

Note: A discount rate of 10% and plant life cycle of 15 years were assumed for calculating the annualised difference.

differences in heat production and capital cost for the two systems considered. The additional heat available from the hot process is approximately 524 MWh per year. The income earned from selling this heat is estimated between £6800 and £9900 per annum, assuming a revenue range of £0.013–0.019 kWh<sup>-1</sup> for heat [4]. Under these assumptions, the additional heat produced by the hot process system appears to justify the higher capital cost (Table 8).

## 6.2. Discussion

### 6.2.1. Efficiency

As expected, the hot process has a higher overall system efficiency (60%) than the cold process (34%) (Table 4). This

Table 8  
Labelling for Figs. 2 and 3

Notation	Equipment/process	
	Cold process (Fig. 2)	Hot process (Fig. 3)
1	Drying stage	Drying stage
2	Co-current gasifier	CFB gasifier
3	Blower	Blower
4	Cyclone	Cyclone
5	Heat exchangers	Heat exchangers
6	Filter	Ceramic filter
7	Wet electrostatic precipitator including quenching process	
8	SOFC	SOFC
9	District heating system	District heating system
10	Combustor	Combustor
11	Impurities	Impurities
A	Anode side of SOFC	Anode side of SOFC
B	Biomass	Biomass
C	Cathode side of SOFC	Cathode side of SOFC
E	Exhaust gas	Exhaust gas

is a result of the superior heat management in the gas cleaning section of the hot process and the higher exit temperature of its gasifier product.

Though both processes produce the same amount of electricity, the carbon conversion for the fluidised bed gasifier used in the hot process is higher than that of the co-current gasifier used for the cold process. Thus, the cold process uses more biomass than the hot process in producing electricity, leading to a lower electrical efficiency.

The efficiencies of both processes are significantly affected by the electric demand of the blower used for cooling of the fuel cell.

### 6.2.2. Cost

Table 5 shows that the hot process is marginally more expensive than the cold process. This is a result of the high estimated cost for the fluidised bed gasifier and the hot ceramic filter. The use of a tar cracker would have further increased the cost for the hot process, while the use of alternative wet cleaning equipment, as opposed to the wet electrostatic precipitator, would have reduced the cost for the cold process.

The unit capacity cost per kW<sub>e</sub> produced is higher for the hot process, as expected because of its higher capital cost. However, the heat obtained from the hot process is much higher than that from the cold process. Thus, the unit capacity cost in terms of total (electrical and thermal) power generated is higher for the cold process.

### 6.2.3. Impurity levels in fuel gas stream entering SOFC

Wet gas cleaning used for the cold option is a developed and reliable process, and the levels of impurities produced are lower than those for the hot process. The use of a second ceramic filter for the hot process would reduce alkali compounds and particulates to levels similar to those for the cold process, but at an additional cost to the system. The same holds for the use of a tar cracker in the reduction of tars to levels similar to the cold process.

### 6.2.4. Overall system performance

It is difficult to compare the two processes without knowing the actual tolerance limits of a SOFC to impurities. It is assumed that only three impurities are produced during gasification. This is a simplification, as other impurities such as HCl and NH<sub>3</sub> may be produced. Wet gas cleaning can deal with most of these impurities, unlike hot gas cleaning. This favours the cold process. Hot gas cleanup systems also need to be proven reliable [14].

## 7. Conclusions

A simple steady-state model of a 200 kW<sub>e</sub> SOFC combined heat and power system has been developed to determine system efficiency and for cost analysis. Two different options were evaluated, a process involving cold gas cleaning and one involving hot gas cleaning.

Results obtained from the simulation show that system efficiency for the hot process is higher than that for the cold process because of better heat management in the cleaning process and higher gasification temperature. Though the capital cost for the hot process is marginally higher, income earned from selling the extra heat produced may justify the additional cost.

The concentration of impurities in the gas stream entering the SOFC for the cold process is estimated to be lower than for the hot process. Though there are no confirmed tolerance limits for SOFCs, the levels for the cold process are within acceptable ranges for other equipment like internal combustion engines [7]. Those for the hot process are higher, and may cause problems in the operation of downstream equipment. For the hot process to produce a gas with similar concentration levels, additional equipment would be needed. This would increase the capital cost and the economic viability of the system would need to be verified based on potential electricity and heat revenues. Also, the equipment for gas cleanup at high temperatures is not proven technology.

The major factor affecting the efficiency of both options was the high power requirement of the blower needed to supply cooling air to the SOFC. The flow-rate of the cooling air supplied by the blower is a function of the temperature gradient across the SOFC. For this paper, the gradient assumed was 100 °C and the power demand of the blower represents over 60% of the total power demand for both processes. For a gradient of 150 °C, this value is about 50%. The use of more efficient cooling methods for the SOFC could therefore have a significant beneficial impact on these systems.

The findings of this paper suggest that hot gas cleaning is the preferred option to cold gas cleaning for the integration of biomass gasification with a SOFC in terms of system efficiency. However, further analysis is required on some issues (level of impurities in the fuel gas stream, cost and equipment reliability) relating to hot gas cleaning to substantiate this claim.

## Acknowledgements

Financial support from the UK Engineering and Physical Sciences Research Council (EPSRC) through grants GR/N34789 and GR1R09282 is gratefully acknowledged.

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