Integrating Combined Heat and Power with Biomass Gasification and SOFC-micro Gas Turbine

**Kurzfassung**

Integrierte Wärme- und Stromerzeugung mit Biomassevergasung und SOFC-Mikrogasturbinen


**Authors**

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* Dr. Fryda was awarded the Heinrich-Mandel-Prize 2007 for her work on the utilisation of biomass in decentral generation stations < 1 MW with fluidised bed combustion systems and the application of oxide-ceramic fuel cells.

**Introduction**

There is an increasing trend to develop more efficient biomass-fuelled energy systems. Steam cycles driven by biomass combustion in the range of 5 to 20 MWe currently produce most of the bioelectricity worldwide with electrical efficiencies around 20%. Solid biofuel gasification was recently established as a significant option for power plants based on combined cycles that can achieve efficiencies of above 35% in the range of 20 to 40 MWe [1, 2]. In parallel, R&D moved towards efficient, small-scale CHP systems incorporating gasification and use of product gas in internal combustion engines, micro gas turbines or fuel cells. More specifically, there is a growing interest and promising works on the combination of biomass gasification and Solid Oxide Fuel Cells (SOFCs) [3 to 6], mainly because these can utilise the product gas CO content and have inherent higher tolerances towards several product gas contaminants compared to other fuel cell types.

The paper at hand presents an investigation on the combination of an air-blown fluidised bed biomass gasifier with a high-temperature SOFC and/or MGT in a CHP system of less than 1 MWe, which could respectively operate at two pressure levels, near atmospheric and ~4 bar. The analysis is based on realistic performance estimations by taking into account the functionality of the proposed configurations without overestimations or gross assumptions. For this purpose, accurate models were incorporated into the Aspenplus™ process simulation software for all the integrated unit operations, followed by an exergetic analysis; second law efficiencies for the major process steps and for the overall CHP system were evaluated and discussed, encouraging comparisons with existing exergetic analyses on biomass gasification [7] and SOFCs [8 to 14].

**System Description and Modelling Aspects**

Three possible configurations for a Comb.
Steady state flow sheet models of the three configurations were developed in AspenPlus™ process simulation software together with special FORTRAN blocks for the modelling of gasifier and SOFC sections. The Redlich-Kwong-Soave cubic equation of state was employed for the gas properties estimation [17]. The pressure drop across each unit operation was assumed 1%. Heat exchangers were allowed a significant minimum temperature difference of \(\Delta T > 50 \text{ K}\) in counter-current mode. This is a rather conservative assumption if higher efficiencies are demanded but further reducing \(\Delta T\)s would result in costly heat exchangers. Heat losses in the gasifier, the HRSG and the heat exchangers were assumed \(\geq 2\%\) of the heat transferred.

### Methodology for the Exergetic Analysis

Steady state flow sheet results were used to perform an exergy analysis on each subprocess, i.e. work potential of material and heat streams at any point in a series of energy conversion devices were evaluated in exergy terms. Exergy is the maximum possible amount of work that can be obtained from a material or heat stream that eventually equilibrates with the reference environment, which consists of reference components and is characterised by absence of pressure and temperature gradients.

The exergy of a material stream is given as the sum of molar physical and chemical exergy:

\[
E = N (e_{ph} + e_{ch}) \text{ (W)} \tag{1}
\]

The molar physical exergy of a material stream is evaluated using the data on physical properties, temperature \(T\), pressure \(p\), enthalpy \(h\) and entropy \(s\), calculated by AspenPlus™ and its properties in reference environmental conditions \((T_o = 298.15 \text{ K}, p_o = 1.013 \text{ bar})\) using the following expression:

\[
e_{ph} = (h - h_o) - T_o (s - s_o) \text{ (J mol}^{-1}) \tag{2}
\]

The molar chemical exergy is obtained when the components of the energy carrier are converted to reference compounds and diffuse into the environment:

\[
e_{ch} = \sum_i x_i e_{ci} + R T_o \sum_i x_i \ln x_i \text{ (J mol}^{-1}) \tag{3}
\]

where \(x_i\) is the mole fraction and \(e_{ci}\) the standard molar chemical exergy \((\text{J mol}^{-1})\) of each component \(i\), assuming a reference atmospheric composition given by \(Kotar\) [18].

The chemical exergy of the solid fuel was calculated 17689 kJ kg\(^{-1}\) with the help of the statistical correlation \(\beta\), proposed by \(Szargut\) [19]:

\[
e_{ch, fuel} = (LHV_{\text{fuel, dry}} + w_{\text{H}_{2}O}) \beta + \left( e_{o} - H_{o}^\circ z_i \right) \text{ (kJ kg}^{-1}) \tag{4}
\]

Table 1 presents the proximate and ultimate analysis of the biomass fuel (dried olive kernel). The oxygen to carbon mass fraction of the solid fuel is calculated between 0.667 \((x_{O_2} / z_c)\) < 2.67, and the formula for wood is applied [19], which gives \(\beta = 1.1182\).

The exergy of a heat stream \(Q\) is given with the help of the Carnot factor: \(E_X^Q = Q (1 - T_o / T)\), where \(T\) is the temperature at which \(Q\) is available. Exergy of power output, \(E_X\), equals power itself.

AspenPlus™ flow sheet calculations provide values for mole flows \(N\) and mole fractions \(x_i\) of all streams, as well as their physical properties \((h, s, T, p)\). The evaluation of the molar reference enthalpy and entropy \((h_o, s_o)\) of every material stream was obtained by a duplicate of each stream expanded and cooled to reference conditions.

For the modeling of a realistic system, heat losses were introduced in several unit operations. These losses do not increase the

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**Figure 1. Flow sheet diagram of the biomass gasification CHP with SOFC and/or MGT.**

The flow sheet diagram of the biomass gasification CHP with SOFC and/or MGT.
streams’ entropy but contribute to exergy loss out of the system. Additionally, exergy losses due to mechanical and electrical inefficiencies were considered; these deteriorate power outputs from equipment such as turbomachinery and the inverter. Non-ideal conditions and real gas equations were used for the simulation of each unit operation and exergy losses due to mixing were accounted; these cannot be avoided and contribute to entropy increase and therefore exergy destruction within the boundary of the system due to dissipation (irreversible exergy destruction). Both exergy losses and exergy destruction have been summed up under the term ‘irreversibilities’ IR, and an exergy balance for a control volume is expressed:

\[
\sum E^e + \sum E^f = \sum E^r + \sum E^w + \text{IR} \quad (W)
\]  

**Modelling of the Combined Heat and Power System**

**Modelling Biomass Air Gasification**

The Gibbs free energy minimisation method for the C-H-O atom blend of the biomass fuel and oxidant mixture was applied for predicting the thermodynamic equilibrium composition of product gas major components: H₂, CO, CH₄, CO₂, H₂O, N₂, as well as char, and oxidant mixture was applied for predicting the thermodynamic equilibrium composition of the biomass fuel, as Equivalence Ratio. Therefore, non-ideal conditions and real gas equations were used for the simulation of each unit operation and exergy losses due to mixing were accounted; these cannot be avoided and contribute to entropy increase and therefore exergy destruction within the boundary of the system due to dissipation (irreversible exergy destruction). Both exergy losses and exergy destruction have been summed up under the term ‘irreversibilities’ IR, and an exergy balance for a control volume is expressed:

\[
\sum E^e + \sum E^f = \sum E^r + \sum E^w + \text{IR} \quad (W)
\]

**Table 1. Biomass fuel data.**

<table>
<thead>
<tr>
<th>Proximate analysis</th>
<th>Ultimate analysis (%/w dry basis)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volatiles (%/w dry)</td>
<td>72.64</td>
</tr>
<tr>
<td>Fixed carbon (%/w dry)</td>
<td>24.78</td>
</tr>
<tr>
<td>Moisture (%/w)</td>
<td>10.0</td>
</tr>
<tr>
<td>Heating values</td>
<td></td>
</tr>
<tr>
<td>HHV (kJ/kg dry)</td>
<td>18 900</td>
</tr>
<tr>
<td>LHV</td>
<td>15 567</td>
</tr>
</tbody>
</table>

The gasifier fuel input is expressed considering the amount required for stoichiometric combustion of the fuel, as Equivalence Ratio:

\[
\text{ER} = \frac{\text{Air input (kg s}^{-1}\text{)}}{\text{Stoichiometric air (kg s}^{-1})} \quad (6)
\]

The gasification cold gas efficiency neglects the sensible heat of the gas and char produced and is defined as [1]:

\[
\eta_{cg} = \frac{\text{LHV in cold product gas}}{\text{LHV in feedstock}} \quad (7)
\]

In order to evaluate the degree of biomass work potential conservation in the gaseous product fuel, exergy analysis of air gasification is applied for the control volume 1 (CV 1) shown in Figure 1. Based on the general definition of the degree of perfection for a process by Szargut [19], the exergetic efficiency of air gasification is:

\[
\eta_{ex, gas} = \frac{\text{E_{gas}} + \text{E_{char}}}{\text{E_{biomass}} + \text{E_{turb}} + \text{E_{fan}}} \quad (8)
\]

Here \(\text{E_{gas}}\) includes the product gas sensible heat. The gasifier operates autothermally, therefore \(\text{E_{turb}}\) is zero. Since in this application the physical and chemical exergy of char is of no use it was not included in the nominator of equation (8).

**Modelling the SOFC**

The SOFC configuration and its control volume are depicted in Figure 1 (CV 2). A typical tubular cathode supported SOFC similar to the Siemens Westinghouse system [16] was modelled. Both anode and cathode depleted fuel and air are assumed to exit the SOFC stack compartment at 1173 K and are introduced to the post-cell combustor which serves as a final air pre-heater. The flue gas temperature could significantly increase, but for their use in an MGT expander further amount of compressed preheated air is introduced to suppress the turbine inlet temperature (TIT) to its maximum allowed value, assumed 1173 K.

Two operation pressure levels were studied corresponding to the two gasifier operation pressure levels taking into account pressure losses, i.e. \(p_{SOFC} = 1.2\) bar and 3.56 bar. Due to very low methane and hydrocarbons concentrations in the clean product gas, the common internal pre-reformer was not employed here [22]. Also contrary to natural gas fuelled SOFC configurations [23], no recirculation of depleted anode gas is considered, due to high nitrogen content of the fuel gas that would significantly dilute the anode gas. Supplementary steam is added to the product gas before it reaches the catalytic SOFC anode, to ensure carbon deposition-free operation. In the base case calculations, the additional quantity of steam is specified to achieve a Steam to Carbon Ratio (STCR) equal to 2:

\[
\text{STCR} = \frac{n_{H_2O}}{n_{C_{H_2} + n_{CO} + n_{CO_2}}} \quad (9)
\]

Nevertheless the requirement for large amounts of steam raises two problems: finding available water quantities and coping with efficiency penalty deriving from exergy destruction associated with steam production. The level of the latter negative effect was investigated by setting STCR as a parameter ranging from 0.5 (which is the least thermodynamic requirement to avoid carbon deposition) up to the value of 2. Similarly high STCR values have been suggested when partially pre-reformed methane is fed to SOFCs to assure no carbon deposition will occur [23, 24].

The SOFC model was built in AspenPlus™ using available blocks and a calculator with a FORTRAN routine for the electrochemical properties estimation. The electrochemically-reacted oxygen is separated from the cathode and fed to the anode, which is modelled by an RGIBBS reactor model that brings the anode mixture into chemical equilibrium. The low methane content justifies this equilibrium assumption rather than using some methane reforming rate reactions. The fuel utilisation factor of the stack is:

\[
U_I = \frac{n_{H_2, \text{REACT}}}{n_{H_2, \text{in}} + n_{C_{H_2} + n_{CO} + n_{CO_2}}} \quad (10)
\]

where \(n_{in}\) refers to the anode’s fuel species input and \(n_{H_2,\text{REACT}}\) is the H₂ (mol s⁻¹) reacting in the hydrogen electrochemical reaction, which was solely considered:

\[
H_2 + (1/2)O_2 \rightarrow \text{H}_2\text{O} \quad (11)
\]

The output voltage of the cell is:

\[
V = V_{OC} - V_{DH} - V_{ACT} - V_{IR} \quad (12)
\]

The Neumann open circuit cell voltage \(V_{OC}\) was evaluated at a corrected average operating temperature \(T_{SOFC}\), i.e. the average between the mixed anode and cathode inlet flow (~ 900 K) and the outlet of the SOFC at 1173 K:

\[
V_{OC} = \frac{\Delta G^o}{2 \cdot F} - \frac{R \cdot T_{SOFC}}{2 \cdot F} \cdot \ln \frac{p_{H_2}^{\text{stat}} \cdot p_{H_2O}^{\text{stat}}}{p_{H_2}^{\text{in}} \cdot p_{H_2O}^{\text{in}}} \quad (13)
\]

where \(F = 6.023 \times 10^{23} \times 1.602 \times 10^{-19} \text{Cmol}^{-1}\) is the Faraday constant, 2 is the number of e⁻ produced per H₂ mole that reacts through reaction (11) of which the molar Gibbs free energy change is expressed as
$\Delta G = \Delta H - T_{SOFC} \Delta S$, calculated at $T_{SOFC}$ and standard pressure. Finally, $p_{\text{outi}}$ are the SOFC-exit partial pressures of the participating components in reaction (11). Using the above partial pressures and temperature data, a FORTRAN calculator was used for the estimation of the overpotentials due to Ohmic ($\nu_{\text{Ohm}}$, activation ($\nu_{\text{act}}$) and polarisation ($\nu_{\text{pol}}$) losses. This calculator, the details of which were presented elsewhere [14], is based on works from Campanari et al. [24, 25], Chan et al. [26, 27], Costamagna et al. [23], and Selimovic [28].

The SOFC stack’s power output is:

$$P_{\text{SOFC}} = V \cdot I \quad \text{(W)} \quad (14)$$

where the current is evaluated as $I = 2F_{\text{H}_2, \text{REACT}}$. The corresponding current density is $J = I / A_{\text{SOFC}}$ (Am$^{-2}$) where $A_{\text{SOFC}}$ is the active cell surface area (m$^2$).

By specifying the utilisation factor, the anode flow throughput and composition, iterative calculations of the overall energy balance are performed over the SOFC stack control volume to result in the air throughput adjustment to reach an almost adiabatic operation (allowing ~ 5 kWth thermal losses) at the desired SOFC temperature. The electrical efficiency of the stack is:

$$\eta_{\text{SOFC}} = \frac{n_{\text{H}_2} \cdot \text{LHV}_{\text{H}_2} + n_{\text{CO}} \cdot \text{LHV}_{\text{CO}}}{n_{\text{H}_2}}$$

where $n_i$ is the molar input of each gas component and $\text{LHV}_i$ their respective lower heating value, while the exergetic electrical efficiency [19] is:

$$\eta_{\text{ex,SOFC}} = \frac{P_{\text{SOFC}}}{E_{\text{gas}} + E_{\text{sat}}}$$

$$\Delta G = \Delta H - T_{SOFC} \Delta S$, calculated at $T_{SOFC}$ and standard pressure. Finally, $p_{\text{outi}}$ are the SOFC-exit partial pressures of the participating components in reaction (11). Using the above partial pressures and temperature data, a FORTRAN calculator was used for the estimation of the overpotentials due to Ohmic ($\nu_{\text{Ohm}}$, activation ($\nu_{\text{act}}$) and polarisation ($\nu_{\text{pol}}$) losses. This calculator, the details of which were presented elsewhere [14], is based on works from Campanari et al. [24, 25], Chan et al. [26, 27], Costamagna et al. [23], and Selimovic [28].

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$$\eta_{\text{ex,SOFC}} = \frac{P_{\text{SOFC}}}{E_{\text{gas}} + E_{\text{sat}}}$$

Results and Discussion

Gasifier Analysis Results

Figure 2a shows the calculated Equivalence Ratio (ER) and the cold gas efficiency of air gasification vs. gasification temperature. The two gasification pressure levels considered were 1.4 bar and 4 bar. It must be noted that for both operating pressure options the air inlet temperatures are equal, approxi-

![Figure 2. (a) cold gas efficiency (%) and equivalence ratio (ER) of product gas, and (b) hot moist gas composition (% vol dry & N$_2$ free) and product gas exergetic efficiency, for atmospheric and pressurised operation vs. gasifier temperature.](image-url)

Table 2. Input data for peripheral equipment.

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Hot steam</th>
<th>Cold steam</th>
<th>Heat losses % of heat transferred</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat exchangers</td>
<td>In</td>
<td>Out</td>
<td>In</td>
</tr>
<tr>
<td>HX1</td>
<td>–</td>
<td>–</td>
<td>430</td>
</tr>
<tr>
<td>HX2</td>
<td>~ 940</td>
<td>or &gt; 1200**</td>
<td>–</td>
</tr>
<tr>
<td>HX3</td>
<td>–</td>
<td>HX2 hot steam outlet</td>
<td>–</td>
</tr>
<tr>
<td>HX4</td>
<td>–</td>
<td>HX2/HX3 hot steam outlet</td>
<td>–</td>
</tr>
<tr>
<td>HX5</td>
<td>–</td>
<td>HX4 hot steam outlet</td>
<td>363</td>
</tr>
<tr>
<td>Turbomachinery</td>
<td>Isentropic Efficiency (%)</td>
<td>Inlet temperature</td>
<td>Outlet pressure (K)</td>
</tr>
<tr>
<td>MGT</td>
<td>84</td>
<td>75</td>
<td>1173 (max) 293</td>
</tr>
<tr>
<td>Inverter Efficiency (%)</td>
<td>95</td>
<td>98</td>
<td></td>
</tr>
</tbody>
</table>

* accounting for losses through the gas cleaning system
** atmospheric cases (i.e. no MGT), temperature varies on Uf.
mately 480 K. After the conversion of available char, the gasification model results in almost similar product gas composition for both pressures (Figure 2b). There is always some methane left even at higher temperatures and ERs because of methane correction. From a thermodynamic point of view, biomass air gasification processes should be accomplished with the minimum air (ER) necessary above the requirement to maximise carbon conversion. Increasing the gasifier temperature and therefore ER has an overall negative effect on the energetic efficiency because major chemical exergy carrier components, i.e. combustibles in the product gas, are minimised (Figure 2b). Nevertheless, kinetic reasons such as advancement of tar reforming reactions, fluidisation limitations or heat losses might impose higher ER values in practice. The gasifier temperature was chosen 1080 K, and the corresponding ER value in both atmospheric and pressurised mode of operation is 0.37. The model predicts a very slight exergetic effectiveness increase in the case of pressurised gasification. A higher moisture fuel would result in a penalty on the gasification efficiency because of dilution of the product gas with water vapour and requirements of higher ERs to sustain autothermal operation.

SOFC Analysis Results

The clean product gas main composition at both pressure levels, before and after humidification up to STCR = 2, are shown in Table 3. For these anode gas compositions, several characteristic curves are drawn for $T_{SOFC} = 1173$ K and $P_{SOFC} = 1.2$ bar/3.56 bar, respectively. Pressurised operation is improved over atmospheric, for a given fuel throughput and utilisation factor, because (a) slightly increased current densities are achieved, (b) the SOFC voltage ($V$) is increased due to increased open circuit voltage ($V_{oc}$), (c) activation overpotential is less while (d) ohmic and cathode overpotentials remain almost constant. Figure 3 shows the cell power output vs. current density for three fuel utilisation ratios and the two pressure levels. SOFC power is increased for small $U_f$ values because the partial pressures of the reactants remain high until exiting the stack. The 100 m² SOFC stack, fed with product gas from the gasifier with biomass throughput of 200 kg/h, operates with current densities $J$ around 4000 A/m²; this region of operation was chosen for the rest of the calculations.

For finalising the configuration of the CHP system the preferred fuel utilisation has to be determined. The stack’s air requirement (and therefore off-gas volume) increases with $U_f$ because of greater cooling load required from the stack. Furthermore, the post-SOFC combustor temperature decreases with increasing $U_f$ because the air/fuel mixture is leaner. Figure 4 shows the combustor temperature without additional air, for the three $U_f$ values in the pressurised operation at the above-mentioned system through-

<table>
<thead>
<tr>
<th>Table 3. Product gas composition before and after steam addition.</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Component</strong></td>
</tr>
<tr>
<td>CH₄</td>
</tr>
<tr>
<td>CO</td>
</tr>
<tr>
<td>CO₂</td>
</tr>
<tr>
<td>H₂</td>
</tr>
<tr>
<td>H₂O</td>
</tr>
<tr>
<td>N₂</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table 4. Base case results for the three configurations studied.</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Biomass throughput (kg/h)</strong></td>
</tr>
<tr>
<td>SOFC stack active area (m²)</td>
</tr>
<tr>
<td>Anode STCR</td>
</tr>
<tr>
<td>Optimised $U_f$</td>
</tr>
<tr>
<td>$U_f$</td>
</tr>
<tr>
<td>$J$ (A/m²)</td>
</tr>
<tr>
<td>$P_{net}$ (kW)</td>
</tr>
<tr>
<td>$P_{PGT} – P_{COMP}$ (%$P_{net}$)</td>
</tr>
<tr>
<td>$V / VOC$ (Volt)</td>
</tr>
<tr>
<td>$\eta_{el} / \eta_{CHP}$ (%)</td>
</tr>
<tr>
<td>$\eta_{el} / \eta_{CHP}$ (%)</td>
</tr>
</tbody>
</table>

Table 4. Base case results for the three configurations studied.

Even with very high $U_f = 0.85$ some additional air is required to suppress the combustor temperature down to the maximum allowed turbine inlet temperature (TIT) ($1173$ K). Figure 4 also shows the required additional preheated air as percentage of the primary SOFC air. Higher $U_f$ (above 0.85) were not considered because of risk of SOFC anode oxidation. No additional air option was taken into account in the atmospheric option where no MGT-TIT limitation is posed.

Figure 5 shows the comparison of SOFC exergetic efficiencies vs. SOFC power output at different fuel utilisation ratios for the atmospheric and pressurised operation. A first remark is that SOFC stack efficiencies are considerably lower at atmospheric operation. Furthermore it is obvious that maximising the $U_f$ does not necessarily result in higher stack efficiencies. Nevertheless this wrong assumption is very commonly taken for granted in SOFC power cycles presentations which are based on gross assumptions about the SOFC behaviour. Higher efficiencies are gained by lowering $U_f$ at higher electric demands from an SOFC stack of a given active surface area, and the $U_f$ choice has to be based on the examination of the overall CHP efficiency.

CHP System Analysis Results

After correcting the power outputs of the SOFC, $P_{SOFC}$, and MGT, $P_{GT}$, with mechanical and inverter efficiencies respectively, and subtracting the power for compressors operation, $P_{COMP}$, the energetic electrical efficiency of the system is defined as:

$$\eta_{el} = \frac{P_{SOFC} + P_{GT} - P_{COMP}}{(input\ biomass)_{LHV}}$$

while the power and thermal energetic efficiency of the system is defined as:

$$\eta_{CHP} = \frac{P_{SOFC} + P_{GT} - P_{COMP} + Q_{useful}}{(input\ biomass)_{LHV}}$$

The system exergetic efficiency for electricity production is:

$$\eta_{el,ix} = \frac{P_{SOFC} + P_{GT} - P_{COMP}}{E_{biomass} + E_{air}}$$

and the combined electrical and thermal (CHP) exergetic efficiency is:
The electrical exergetic efficiencies of the three examined configurations for a range of fuel utilisation ratios versus biomass throughput are shown in Figure 6. For a great range of fuel throughputs, and in specific around the base case of 1000 kWth fuel energy input (i.e. 200 kg biomass/h) the MGT option has higher electrical efficiencies compared to the atmospheric SOFC system, despite its optimised $U_f$ for that region of operation (at $U_f = 0.75$). The atmospheric SOFC suffers from low effectiveness to produce power together with great exergy losses associated with nil power extraction from the high temperature SOFC flue gases. The larger air compression power consumption of the SOFC-MGT system is more than offset by increased SOFC performance and additional power produced from the MGT resulting in $\eta_{ex,el} \geq 35.6\%$ at biomass exergy input 1000 kW. For the three systems studied and the base case of 200 kg biomass/h, the efficiency ratios (i.e. equations 17 to 20) and important results are presented in Table 4.

The pressurised MGT configuration offers increased thermal output but at low temperatures (Figure 7c), thus not reflected in the $\eta_{ex,CHP}$. Part of the flue gas energy of the SOFC configurations is consumed to produce steam for product gas moistening, therefore, deteriorates the quality of available heat (Figure 7a, b). The influence of the required STCR on the electrical and CHP system exergetic efficiencies was studied parametrically. In order to present the exergy destruction associated with steam production, two exergy loss ratios, $\zeta_{steam}$ and $\zeta_{compr}$ were defined:

$$\zeta_{steam} = \frac{\text{IR}_{\text{steam}}}{E_{\text{IN}}}$$

where $\text{IR}_{\text{steam}}$ is the cumulative irreversibility associated to steam production and its mixing with the syngas prior to entering the anode, and was calculated from a combined exergy balance applied to control volumes CV3 and CV4 in Figure 1:

$$E_{\text{water}} + E_{\text{flue gas}} + E_{\text{syngas}} = E_{\text{steam}} + E_{\text{gas OUT}} + \text{IR}_{\text{steam}}$$
The second exergy loss ratio is associated with fuel cell compressor power consumption, because it is indirectly affected from the level of steam addition since the later changes the cooling demands of the stack:

\[ \zeta_{\text{compr}} = \frac{P_{\text{compr}}}{E_{\text{in}}} \]  

(23)

Figure 7. Heat exchangers temperature profile for (a) pressurised SOFC-MGT, (b) atmospheric SOFC and (c) pressurised MGT configuration.

Figure 8a (pressurised system) shows that the electrical efficiency marginally deteriorates with STCR because despite \( \zeta_{\text{steam}} \) grows there is a counterbalancing reduction of the \( \zeta_{\text{compr}} \) since the additional steam cools the stack replacing considerable amounts of air for this purpose. In all cases, the thermal throughput from the MGT and therefore its contribution to power output is constant (line \( P_{\text{GT}} / E_{\text{in}} \) in Figure 8a) and the minor decrease in the electrical efficiency with STCR is attributed to lower partial pressures of reactants within the stack. Similarly, the latter effect is evident in Figure 8b for the atmospheric operation. In both atmospheric and pressurised configurations the effect of the level of additional steam requirement is reflected on the CHP efficiencies (lines \( \eta_{\text{ex,CHP}} \) and \( \zeta_{\text{steam}} \) almost follow parallel trends) because this affects the temperature at which useful heat is available from HX5.

Conclusions

The combination of biomass gasification with SOFC and/or MGT for small-scale CHP was assessed by modelling in AspenPlus™ process simulation software. Two system operation pressures were studied, atmospheric and ~ 4 bar. This small pressure shift does not have significant effect on the product gas composition or on the exergetic efficiency of the gasification process. On the contrary, the pressurised SOFC operation is greatly improved, and with the additional power from an MGT expander, achieves the highest efficiencies \( \eta_{\text{ex,el}} \) \( \geq \) 35 % at J values around 4000 A/m². The \( U_f \) was optimised at 0.85 for the pressurised and at 0.75 at atmospheric SOFC operation for biomass throughputs of around 200 kg/h. The atmospheric SOFC configur-
tion results in considerably lower efficiencies than the simpler pressurised gasification MGT configuration which gives $\eta_{el} \approx 23\%$, and could only surpass this efficiency if very low power densities were employed. Such a system would probably not be economic, i.e. to have the high SOFC-related investment costs without significant revenues from the power production. Through a detailed exergetic parametric analyses it was shown that the increase of additional steam production to achieve a desired STCR does not greatly affect the electrical efficiencies of the CHP systems because it deteriorates the temperature at which off gases are reflected on the combined thermal exergetic efficiency of the SOFC configurations but is negatively reflected on the combined thermal exergetic efficiency scale biomass gasification plants, Bio-

References

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Heinrich-Mandel-Preis für Kraftwerkstechnik

Heinrich-Mandel-Preis für Kraftwerkstechnik der VGB-FORSCHUNGSSTIFTUNG


Die VGB Forschung hat in der Vergangenheit folgende Preisträger ausgezeichnet:

- Dr. Lydia-Emilia Fryda für ihre Arbeiten zur Nutzung von Biomasse in dezentralen Stromerzeugungsanlagen (unter 1 MW) unter Einsatz einer Wirbelstirnflügelvergaser- und von oxid-katalytischen Brennstoffzellen und
- Dr. Jens Hampel für die Entwicklung eines Turbogenerators mit mechanischem Netz-Kopplung zur Wirkungsgradhöhung kleiner Dampfturbinen.

Die beiden Preisträger stellen die ausgezeichneten Arbeiten in dieser Ausgabe der VGB PowerTech vor.

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- Dr. Lydia-Emilia Fryda for her works on the utilisation of biomass in decentral electricity generation plants (< 1 MW) with fluidised bed gasification and oxide ceramic fuel cells and
- Dr. Jens Hampel for the development of a turbo-generator with mechanical link to the electrical grid for increasing the efficiency of smaller steam turbines.

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