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A comparative study of two SOFC based cogeneration systems fed by municipal solid waste by means of either the gasifier or digester

Original

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Nomenclature

AD	Anaerobic digester	<i>Subscript and abbreviations</i>	
AHX	Air heat exchanger	0	dead state
D-SOFC	Digester coupled SOFC	1,2,3,...	state points
G-SOFC	Gasifier coupled SOFC		
SOFC	Solid oxide fuel cell	<i>Greek Symbols</i>	
HRSG	Heat recovery steam generator	η_{th}	thermal efficiency
EES	Engineering Equation Solver		
FHX	Fuel heat exchanger	η_{II}	exergy efficiency
FuelB	Fuel blower	η_I	energy efficiency
AirB	Air blower	η_{is}	isentropic efficiency
P	pump		
K	equilibrium constant		
ΔG	Change in Gibbs function		
S	entropy		
h	enthalpy		
\dot{W}	Power		
\dot{Q}	Heating load		

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27

28

29 1. Introduction

30 Since the world's energy consumption is forecast to rise remarkably during the next two
31 decades, no one can deny the importance of renewable energy sources. In addition, continued
32 demand for fossil fuels means the world will not be able to reduce greenhouse gases in the
33 atmosphere. In this regard, implementing the power plants utilizing the renewable energy sources
34 such as geothermal, solar, biofuel, biomass and so on, is supposed to be of governments' and
35 researchers' interest.

36 Recent developments in solid oxide fuel cell (SOFC) technology have increased interest in the
37 application toward electricity generation particularly the distributed one. In addition to the
38 advantages of SOFC system over conventional power generation methods, heat from SOFC
39 exhaust can be recovered for combined heat and power (CHP) operations to improve the overall
40 system efficiency.

41 Undoubtedly, the hydrogen is the genuine fuel for the stack; however, considering its cost and
42 availability, using the natural gas could be more viable, practical and economical in addition to
43 the reforming process of the fuel. Nevertheless, feeding the SOFC with fuels such as biogas,
44 biofuels, syngas and alcohols has become of great interest as they could be suitable alternatives
45 for the natural gas. There are numerous solid waste gasification facilities operating or under
46 construction around the world. Gasification has several advantages over traditional combustion
47 processes for municipal solid waste (MSW) treatment. It takes place in a low oxygen
48 environment that limits the formation of dioxins and of large quantities of SO_x and NO_x .
49 Furthermore, it requires just a fraction of the stoichiometric amount of oxygen necessary for
50 combustion (partial oxidation). As a result, the volume of process gas is low, requiring smaller
51 and less expensive gas cleaning equipment. The lower gas volume also means higher partial

52 pressure of contaminants in the off-gas, which favors more complete adsorption and particulate
53 capture. Finally, gasification generates a fuel gas that can be integrated with combined cycle
54 turbines, reciprocating engines and, potentially, with fuel cells converting fuel energy to
55 electricity more efficiently than conventional steam boilers[1,2]. Renewability and less CO₂
56 emission are the greatest advantages of commonly used biomass fuels such as paper, wood,
57 waste straw, saw dust, paddy husk, MSW, etc. [3–6]. Three processes are usually involved in
58 thermochemical conversion of biomass into usable fuels for power plants: combustion,
59 gasification and pyrolysis [7,8]. The gasification process proves to be an efficient technique as it
60 transforms biomass into easily usable fuels to be used for electricity generation [9–11]. There are
61 different types of gasifiers based on the types of moving fluid and solid inside[12]. From the
62 availability view point however, 75% percent of the gasifiers are downdraft, 20% is circulating
63 fluidized bed, 2.5% updraft and 2.5% the other kinds [13].

64 A 200kW SOFC combined heat and power system has been developed by Omosun et al. [14]
65 to evaluate the system efficiency and its cost analysis. Two different options were investigated;
66 one of them involved cold gas cleaning and the other used hot gas cleaning. The results revealed
67 that system efficiency for the hot process is higher than that for the cold process due to the better
68 heat management in the cleaning process and higher gasification temperature. Despite the capital
69 cost for the hot process is marginally higher, income earned from selling the extra heat produced
70 may justify the additional cost. Singh et al. [15] analyzed the carbon deposition in a solid oxide
71 fuel cell (SOFC) fueled by a biomass. It is reported that carbon deposition decreased to zero as
72 the operating conditions were varied to get a fuel mixture with higher water content (about 15%).
73 The conjunction of biomass gasification with solid oxide fuel cells is investigated by Athanasiou
74 et al. [16] and the results showed that the electrical efficiency of the integrated gasification-

75 SOFC-steam turbine is about 68% higher than that for the conventional gasification-steam
76 turbine system. A direct internal reforming-SOFC (IR-SOFC) operating with syngas was
77 modeled thermodynamically by Colpan et al [17]. The results showed that recirculation ratio
78 does not have a significant effect for low current density conditions. However, at higher current
79 densities, increasing the recirculation ratio decreases the output power and electrical efficiency
80 of the cell. Jang et al. [18] studied a direct power generation from waste coffee grounds
81 in a biomass fed fuel cell. Results show that biomass type has crucial effect on cell performance.
82 They used waste coffee ground (WCG) as biomass with the benefit of not needing any pre-
83 reformer in the system. It is reported that at cell operating temperature of 900°C the system
84 shows maximum power density twice than that of the carbon black. Pieratti et al. [19]
85 investigated experimentally and theoretically the syngas suitability for solid oxide fuel cell
86 applications. Considering environmental aspects of biomass combustion, steam gasification is
87 used to produce syngas. Experimental data is used to generate and calibrate a 2D theoretical
88 equilibrium model. **It is found that the thermodynamic approach is a simple engineering useful**
89 **tool to obtain reliable results of the gasification model. It is declared that considering the syngas**
90 **composition and energy content, the obtained syngas is a suitable fuel for fuel cells.**
91 Nevertheless, the gas cleaning is still one of the main critical issues. In particular the tar and the
92 H₂S in the gas can rapidly decrease the life of the fuel cells. A new small cogeneration system
93 consisting of a fluidized bed gasifier, coupled to a SOFC and a micro gas turbine is proposed by
94 Di Carlo et al. [20]. Results disclosed that the best case occurs with a temperature of the cathode
95 gas of 800 °C and moisture of 10%, in this case the fuel utilization could be set equal to 0.79 and
96 the electrical efficiency of the overall system is 48%. Focusing on the anode gas recycling,
97 Lorenzo and Fragiaco [21] analyzed the performance of syngas fed SOFC power plant from

98 the viewpoint of thermodynamics. It is reported that there is an optimum value for anode gas
99 recycle ratio in which the thermal efficiency is maximized. An integrated SOFC and biomass
100 (wood) gasification system using air, enriched air, and steam as gasification agent is investigated
101 by Jia et al. [22]. It is found that when using air or oxygen-enriched air as gasification agent, the
102 gasifier reactor caused the greatest exergy destruction while for steam gasification power
103 systems the largest exergy destruction lies in air heat exchanger. Also, for the efficient CHP case
104 the exergy efficiency is calculated to be 36%. In another work, Jia et al. [23] studied the effects
105 of various parameters such as moisture content in biomass, equivalence ratio and mass flow rate
106 of dry biomass on the overall performance of SOFC based CHP system. It is reported that char in
107 the biomass tends to be converted with decreasing of moisture content and increasing of
108 equivalence ratio due to higher temperature in reduction zone of gasifier. Kartha et al. [24]
109 studied a small-scale biomass fuel cell/gas turbine power systems for rural areas. A downdraft
110 gasifier is used to produce syngas because it is reported to be commercialized and simple in
111 design which has very little amounts of tar (the excessive production of which needs a separate
112 reactor for tar removal and dissipates amounts of flue gas energy) and has simple method of gas
113 cleaning. Also, the efficiency of the studied SOFC-GT was found to be 43.4%. Federico
114 Ghirardo et al. [25] studied heat recovery options for fuel cells. It is found that about 181 kW of
115 heat can be recovered in an ORC to produce 35 kW of electricity. The overall efficiency
116 increases from 44% to 49% when the recovery system is used and the cost of energy drops from
117 25 c\$/kWh (isolated SOFC system) to 22 c\$/kWh.

118 Even though continued progress has been made with other alternative treatment technologies
119 (gasification, pyrolysis, plasma, biological drying, etc.), these technologies have by far not seen
120 the same widespread implementation that anaerobic digestion has been able to achieve. In

121 Europe alone, 244 installations dealing with the organic fraction of municipal solid waste as a
122 significant portion of the feedstock have been constructed or are permitted to be constructed.
123 Feeding SOFCs by biogas is investigated by researchers in the recent years. The performance of
124 biogas-fed solid oxide fuel cell system utilizing different reforming agents (steam, air and
125 combined air/steam) was investigated by Piroonlerkgul et al. [26]. It is observed that for the
126 steam-fed SOFC, there is an optimal amount of steam which provides a maximum power
127 density. However, for the air-fed SOFC system, the power density always decreases with the
128 increase of amount of air due to the dilution effect of nitrogen in air. Three configurations of
129 biogas fed solid oxide fuel cell micro-combined heat and power (micro-CHP) systems are
130 studied with a particular emphasis on the application for single-family detached dwellings by
131 Farhad et al. [27]. Tjaden et al. [28] investigated a small scale biogas-SOFC plant in 2014.
132 Results show that the maximum electrical efficiency is calculated 56.55% which is 15% higher
133 than that of the combustion engines fueled by biogas. Also, the advantages and disadvantages of
134 different reforming process such as steam reforming and auto thermal reforming are reported.
135 The results revealed that the cell design voltage is higher than the cell voltage at which the
136 minimum number of cells is obtained for the SOFC stack. Also, the maximum electrical
137 efficiency of 42.7% is obtained for one of the configurations. Producing biogas from biomass
138 and then feeding either the SOFC or internal combustion engine is studied by Santarelli et al.
139 [29]. Optimization results revealed that produced electrical energy for the SOFC is higher than
140 that for the internal combustion engine where the consumed thermal energy is the same for both
141 systems. Papurello et al. [30] studied the performance of a biogas-fed SOFC power plant
142 experimentally. A 500 W_{el} SOFC stack was installed at a biomass digester pilot plant and was
143 fed with real biogas for more than 400 hours, after which a stable voltage was achieved under

144 partial oxidation reforming conditions. The fuel utilization was approximately 55% and the
145 electrical efficiency was close to 34%. The biogas which is obtained from organic waste
146 collection from local municipal areas via digester consists of methane and carbon dioxide
147 concentrations ranging from 60-70% and 30-40% vol., respectively. Trendewicz and Braun [31]
148 analyzed a biogas-fueled solid oxide fuel cell (SOFC) system for producing heat and power from
149 the view point of techno-economic. They estimated that the baseline cost of electricity for the
150 small, the medium, and the large plants is 0.079 \$/kWh, 0.058 \$/kWh and 0.05 \$/kWh,
151 respectively. Gandiglio et al. [32] proposed a model to analyze the integration of waste water
152 treatment (WWT) biogas and solid oxide fuel cell considering both the internal and external
153 reforming. The influence of fuel utilization, internal reforming, biogas composition and steam-
154 to-carbon ratio on both the SOFC and overall plant performance is investigated. It is observed
155 that an increase in the methane concentration of biogas would increase the electrical efficiency of
156 the plant slightly. Siefert and Litster [33] investigated the performance of a biogas-fed SOFC
157 from the viewpoint of economics. Their interesting result may be the one revealing that the
158 anaerobic digestion-SOFC system is significantly more economic than the systems in which the
159 biogas is sent to internal combustion engines or micro gas turbines. An exergoeconomic analysis
160 of biogas-fed SOFC power systems focusing on the (anode/cathode) gas recycling investigated
161 by Mehr et al. [34]. It is reported that, the solid oxide fuel cell system with anode and cathode
162 recycling is superior to the other configurations and its efficiency is calculated as 46.09% being
163 6.81% higher than that of the simple solid oxide fuel cell fed by natural gas. The unit product
164 cost of the solid oxide fuel cell system with anode and cathode gas recycling is calculated as
165 19.07\$/GJ which is about 35% lower than the corresponding value for the simple natural gas fed
166 solid oxide fuel cell system.

167 As discussed above, there are two common ways to produce gas from the biomass, using
168 gasifier or digester. In the present work, a downdraft gasifier and a thermophilic digester are
169 used to produce syngas and biogas from the municipal solid waste, respectively. The produced
170 gases are supposed to drive the SOFC system to produce power. As the compositions of biogas
171 and syngas are quite different, and also the exhaust of SOFC is hot enough, the CHP system
172 would be an interesting choice to utilize the hot exhaust of SOFC stack to produce heating. Also,
173 it would be interesting to reveal which system is more efficient and more economical with the
174 almost same working conditions. Modeling of the systems is performed in EES software and the
175 performance of systems is compared with each other from the thermodynamics and thermo-
176 economics viewpoints.

177 2. System description and assumptions

178 2.1 Systems configuration

179 Schematic diagram of the proposed cogeneration system based on the digester unit system is
180 illustrated in Fig.1a. The system consists of a biogas production unit, a SOFC with anode
181 recycling and heat recovery steam generator (HRSG). Biomass is sent to the digester at
182 environmental conditions and as **the digester is considered thermophilic one the temperature** of
183 biogas exiting the digester will be higher than environment temperature. Then the biogas is sent
184 to the cleanup unit, in which H_2S is coldly removed and then the clean biogas is sent to the
185 blower. The biogas and air are preheated through the fuel heat exchanger and air heat exchanger,
186 respectively, after being pressurized with the help of fuel and air blowers (the pressure is just to
187 make a flow of fuel and air and compensate the pressure drops within the system). The heated air
188 is sent to the cathode of the stack. On the other hand, the biogas is sent to the anode after mixing

189 with the recycling of anode stream (state15). The mixed stream experiences the internal
190 reforming process which brings hydrogen-rich products participating in the electrochemical
191 reaction inside the fuel cell stack. An inverter is used to convert the DC power generated by the
192 stack into grid quality electricity. The electrochemical reaction generates thermal energy a part of
193 which is used to deliver the required heat of the internal reforming reaction, another part is
194 employed to heat up the cell products and residual reactants, and the remaining small amount is
195 transferred to the environment as a heat loss (in the present modeling heat loss is neglected).
196 After the finishing of electrochemical reaction in the SOFC stack, the excess air out of the
197 cathode and the unreacted fuel out of the anode combust completely in an after-burner to
198 generate the combustion gas under high temperature. The exhaust gas from the afterburner is
199 sequentially used to preheat the fuel and air, respectively. The exhaust (state 15) is still hot
200 enough to be utilized to produce hot steam. Therefore, a HRSG unit is considered to be coupled
201 with SOFC system. For the maintenance of the digester temperature in the thermophiles
202 condition, the produce hot water can be used to compensate the heat demand of digester. Fig. 1b
203 depicts the use of syngas produced in the gasification process to produce power and heating by
204 means of SOFC power plant and HRSG unit. The description of the process for the system is
205 almost the same as described for the Fig. 1a. The air from environment is brought to the gasifier
206 along with the biomass in the atmospheric pressure. The equilibrium model is presumed in
207 modeling the gasifier. The equilibrium modeling assumes that all the reactions are in
208 thermodynamic equilibrium and that the pyrolysis product (gas) which is product of interest
209 burns and achieves equilibrium in the reduction zone before leaving the downdraft gasifier.
210 Meanwhile, the high temperature syngas produced in gasifier first used to preheat the air before
211 sending it to the mixer to be mixed with anode gas recycle.

212 2.2 Assumptions

213 Some meaningful assumptions are considered for the systems modeling as listed below
214 [35,36];

- 215 • The atmospheric air is composed of 79% N₂ and 21% O₂, on a volume basis.
- 216 • Gasification temperature is assumed to be 1073 K and the syngas is in thermodynamic
217 equilibrium
- 218 • A thermophilic digester is used in the modeling
- 219 • Fan work in the digester is negligible
- 220 • All gases are treated as ideal gases and gas leakage from the components and the connecting
221 pipes is negligible.
- 222 • The analysis is carried out under thermodynamic equilibrium and steady state conditions.
- 223 • Changes in kinetic and potential energies are neglected.
- 224 • Temperatures at channel inlets are the same and, similarly, temperatures at the channel exits
225 are the same.
- 226 • The fuel cell is insulated perfectly so that there is no heat interaction with the environment
- 227 • Contact resistances are negligible.
- 228 • Unreacted gases are assumed to be fully oxidized in the afterburner. [37]

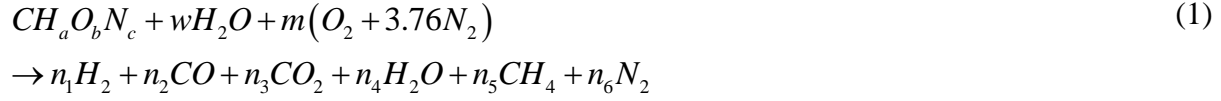
229 The input data for systems' simulation are listed in Table 1.

230 3. Energy analysis

231 3.1 Gasifier modeling

232 The equilibrium model presumed in gasifier modeling assumes that all the reactions are in
233 thermodynamic equilibrium and that the pyrolysis product burns and achieves equilibrium in the

234 reduction zone before leaving the downdraft gasifier [36]. The reactions in the reduction zone
 235 are as stated in the literature [36,38,39].



236 For a biomass, the global gasification reaction is as follows [40]:

237 The equilibrium constants for methane formation reaction and for the water-gas reaction (shift

$$K_1 = \frac{n_5}{n_1^2} \left(\frac{P/P_{ref}}{n_{tot}} \right)^{-1} \quad (2)$$

$$K_2 = \frac{n_1n_3}{n_2n_4} \left(\frac{P/P_{ref}}{n_{tot}} \right)^0 \quad (3)$$

238 reaction) are given by Eqs. 2 and 3, respectively [38]:

239 where n_1 to n_5 are the number of moles in the gasification products in eq. 1. and K_1 and K_2 are

240 the equilibrium constants which can be related to the change in the Gibbs function as follows

241 [38]:

$$\frac{-\Delta G_1^0}{\bar{R}T_g} = \ln K_1 \quad (4)$$

242

$$\frac{-\Delta G_2^0}{\bar{R}T_g} = \ln K_2 \quad (5)$$

243 Where $-\Delta G_1^0$ and $-\Delta G_2^0$ are the change in the Gibbs free function of methane formation equation
244 and shift reaction respectively[36].

245 In Eq. 1, $CH_aO_bN_c$ denotes the biomass, w is the biomass moisture content and m is the kmol of
246 oxygen per kmol of biomass. The parameter m is actually the air fuel ratio and can be determined
247 if the gasification temperature is known (or vice versa). The coefficients n_1 to n_6 are determined
248 through applying the mass balance for H , C , O and N . Based on the ultimate analysis of the
249 MSW as given in Table 2.

250 The moisture content per mole of biomass can be expressed in terms of mass based moisture
251 content (MC) as follows [36]:

$$w = \frac{M_{biomass} MC}{18(1-MC)} \quad (6)$$

252 where,

253 $MC = (\text{mass of water/mass of wet biomass}) \times 100$

254 Assuming an adiabatic gasification at a temperature of 1073 K, the energy balance equation, as

$$\begin{aligned} \bar{h}_{f-MSW}^0 + w(\bar{h}_{f-H_2O}^0) &= n_1(\bar{h}_{f-H_2}^0 + \Delta\bar{h}_{H_2}) + n_2(\bar{h}_{f-CO}^0 + \Delta\bar{h}_{CO}) \\ &+ n_3(\bar{h}_{f-CO_2}^0 + \Delta\bar{h}_{CO_2}) + n_4(\bar{h}_{f-H_2O}^0 + \Delta\bar{h}_{H_2O}) + n_5(\bar{h}_{f-CH_4}^0 + \Delta\bar{h}_{CH_4}) + n_6(\bar{h}_{f-N_2}^0 + \Delta\bar{h}_{N_2}) \end{aligned} \quad (7)$$

255 indicated below, is solved to find the air fuel ratio.

256 In order to validate the gasification model, the experimental and theoretical data reported in

257 the literature are used for gasification temperature of 1100K and moisture content of 16%. The
258 comparison is shown in Table 3. Referring to Table 3, the sum of hydrogen and carbon
259 monoxide content percentages predicted by the present model is 36.43%, which agrees with the
260 experimental data (35.4%) reported by Jayeh [41]. It is found that the obtained results for the
261 present model agree well with those reported by Jarungthammachote [42] as well.

262 In order to validate the simulation results of SOFC, the available experimental data reported
263 by Tao et al. [43] is used. Table 4 compares the cell voltage and power density obtained in the
264 present model developed by the authors and those reported by Tao et al. [43]. The comparison
265 shows a good agreement between them.

266 3.2 Digester modeling

267 In modeling of digester, a thermophilic anaerobic digester with a temperature of 55°C is used.
268 Note that, it is necessary that temperature fluctuations do not occur in the digester to maintain the
269 microbial activity.

270 With knowledge of chemical composition of MSW outlined as ultimate analysis in Table 2
271 the biogas composition can be predicted following the method proposed by Baswell and Hatfield
272 [44]. The global reaction occurring in the digester to produce biogas is:

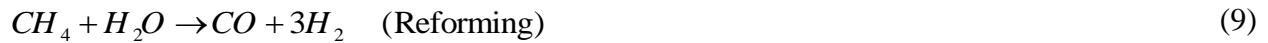


273 In the present work, it is assumed that 0.95 of the OFMSW is volatile. A well designed
274 digester aims to destruct at minimum 0.7 of the volatile solid. Following the procedure presented
275 by Murphy[45] the methane and carbon dioxide mass is found. Knowing that the number of
276 molecules in a unit volume under standard conditions is the same for all gases (This volume is
277 22.412m³/kg) one may find that the 58% of the biogas is methane and 42% of biogas is carbon

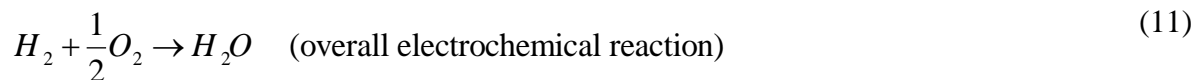
278 dioxide in volume.

279 3.3 SOFC with anode recycle model

280 Energy analysis and molar balance for the SOFC with anode and cathode recycle is used in
281 the present work. [46][46]The biogas is converted into a hydrogen-rich synthesis gas inside the
282 fuel cell by making use of internal reforming and shifting reactions [35]. Also syngas (with
283 hydrogen gas) which does not need any reforming and less methane and carbon monoxide gas is
284 reformed inside the SOFC with the same processes as for biogas. The use of an internal reformer
285 also reduces the dependence of the fuel cell on a cooling system. The chemical reactions in the
286 cells are as follows; [47]

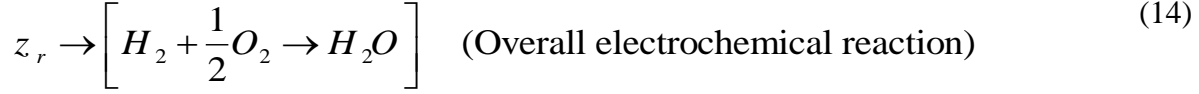
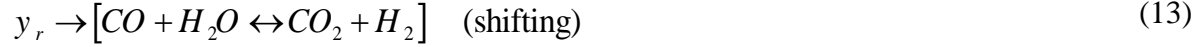
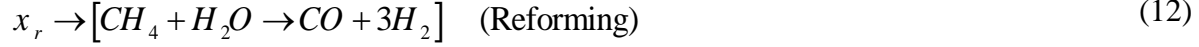


287 Equation 9 is reforming reaction and equation 10 is shifting reaction, carbon monoxide in the
288 reforming reaction, reacts with the water which is brought to the SOFC by means of either
289 recycling or external means such as pump, to produce hydrogen. Hydrogen from shifting



290 reaction is used in the electrochemical reaction as below:

291 The molar conversion rates for reforming, shifting and electrochemical reactions are
292 considered to be x_r , y_r , and z_r , respectively. Therefore, rates of consumption and production of
293 the components can be achieved by the following model:



294 z_r , could be found with the help of current density, Faraday constant, cell number, and active

$$z_r = \frac{j N_{FC} A_a}{2F} \quad (15)$$

295 surface area, as followed by equation (16)

296 Applying mass balance equations along with considering equations 13-16 for the mixing units
 297 and the whole SOFC model the flowing gas compositions may be achieved. In order to solve the
 298 system of equations, 3 more equations are needed to complete the system of the equations.
 299 Looking again in the equilibrium reactions of shifting and reforming, the equilibrium constants

$$\ln K_s = -\frac{\Delta \bar{g}_s^o}{\bar{R}T_{FC,e}} = \ln \left[\frac{(\dot{n}_{CO_2,11} + y_r)(\dot{n}_{H_2,11} + 3x_r + y_r - z_r)}{(\dot{n}_{CO,11} + x_r - y_r)(\dot{n}_{H_2O,11} - x_r - y_r + z_r)} \right] \quad (16)$$

$$\ln K_R = -\frac{\Delta \bar{g}_R^o}{\bar{R}T_{FC,e}} = \ln \left[\frac{(\dot{n}_{CO,11} + x_r - y_r)(\dot{n}_{H_2,11} + 3x_r + y_r - z_r)^3 \left(\frac{P_{12}}{P_{ref}} \right)^2}{(\dot{n}_{CH_4,11} + x_r)(\dot{n}_{H_2O,11} - x_r - y_r + z_r) \times \dot{n}_{12}^2 \left(\frac{P_{ref}}{P_{ref}} \right)} \right] \quad (17)$$

300 can be written as follows respectively:

301 Where, \bar{R} and $T_{FC,e}$ are the **universal gas constant and** the temperature at the exit of the
 302 SOFC, respectively. Also, $\Delta \bar{g}^o$ is the change in the Gibbs free function of shifting and reforming

303 reactions[35].

304 The last equation needed for solving the system equations is obtained using the energy
305 balance for the whole stack. Neglecting the heat loss from the stack, the energy balance can be

$$\dot{W}_{FC,stack} = \sum_k \dot{n}_{k,12} \bar{h}_{k,12} + \sum_L \dot{n}_{L,4} \bar{h}_{L,4} - \sum_m \dot{n}_{m,11} \bar{h}_{m,11} - \sum_n \dot{n}_{n,3} \bar{h}_{n,3} \quad (18)$$

306 written as follows;

307 Where, k , L , m and n are the corresponding gas compositions in each states (e.g. gas
308 composition at state 12 (L) is CO_2 , CO , H_2O , CH_4 , N_2 and H_2). On the other hand, the work rate

$$\dot{W}_{FC,stack} = N_{FC} j A_a V_c \quad (19)$$

309 produced by the SOFC stack $\dot{W}_{FC,stack}$ can be expressed as:

$$V_c = V_N - V_{loss} \quad (20)$$

310 Where cell voltage is defined as:

311 Here, V_N is the Nernst voltage and V_{loss} the voltage loss, which is the sum of three separate
312 voltage losses; Ohmic, Activation and Concentration losses:

$$V_{loss} = V_{ohm} + V_{act} + V_{conc} \quad (21)$$

313 Looking again in the comprehensive analysis of the cell voltage and voltage losses it is found
314 that voltages strongly depend on molar fractions, pressures, electrolyte types and cell operating

315 temperature that need to be verified very precisely. For the sake of brevity detailed voltage
 316 modeling used in the present work has been presented in Appendix A.

317 3.4 Heat recovery steam generator analysis

318 One of the determinant parts of the proposed systems is HRSG, which affects the final system
 319 specifications and has crucial effect on determining whether to choose gasifier or digester. Gas
 320 flow from the AHX is hot enough that can be utilized to generate saturated steam in HRSG. As
 321 shown in Fig. 1a and Fig. 1b the HRSG involves economizer (Eco) and evaporator (Eva) parts.
 322 Water is pressurized by a pump (P) to meet the required steam pressure for the HRSG.

323 Applying the energy balance as well as the pinch point temperature difference for the whole
 324 HRSG system the mass flow rate of steam and temperature of exhaust can be determined. Energy
 325 balance for the economizer and the evaporator parts are as follows:

$$\sum_j n_j \left(\bar{h}_{f_j}^0 + \Delta \bar{h} \right)_{exhaust\ gas,\ in} + n_{water,\ in} \bar{h}_{water,\ in} = \sum_j n_j \left(\bar{h}_{f_j}^0 + \Delta \bar{h} \right)_{exhaust\ gas,\ out} + n_{water,\ out} \bar{h}_{water,\ out} \quad (22)$$

326 Where the amount of heat used for generating hot steam is defined as;

$$\dot{Q}_{Heating} = \dot{m}_{product\ gas} (h_{15} - h_{17}) \quad (23)$$

327 3.5. First law efficiency for proposed system

328 Finally with regarding the above mentioned analysis, the first law efficiency for the power
 329 generation (electrical efficiency) mode and CHP mode could be defined as follows respectively:

$$\eta_{I,P} = \frac{\dot{W}_{FC,\ stack,\ ac} - \dot{W}_{FC} - \dot{W}_{AC} - \dot{W}_{pump}}{\dot{m}_{biomass} LHV_{biomass}} \quad (24)$$

$$\eta_{I,CHP} = \frac{\dot{W}_{FC,stack,ac} - \dot{W}_{FC} - \dot{W}_{AC} - \dot{W}_{pump} + \dot{Q}_{Heating}}{\dot{m}_{biomass} LHV_{biomass}} \quad (25)$$

330 Where \dot{W}_{FC} , \dot{W}_{AC} and \dot{W}_p is the amount of fuel compressor power and air compressor power
 331 required to blow the fuel and air to the SOFC and pump required power respectively.

332 4. Exergy analysis

333 An effective use of energy can be assessed by means of exergy analysis for energy converting
 334 systems. Exergy can be divided into four parts: physical, chemical, kinetic and potential
 335 exergies. For the processes involved in this paper, the latter two are neglected since the changes
 336 in elevation and velocity are negligible [48,49]. Thus, the specific exergy of any stream is
 337 expressed as

$$e = e_{ph} + e_{ch} \quad (26)$$

338 where the physical exergy is defined as [48] :

$$e_{ph,i} = h_i - h_0 - T_0 (s_i - s_0) \quad (27)$$

339 The specific chemical exergy of a solid biomass fuel can be expressed as follows [36]:

$$e_{biomass}^{ch} = \beta LHV_{biomass} \quad (28)$$

340 where the factor β in Eq. 28 denotes the ratio of the chemical exergy to the lower heating value
 341 (LHV) for the organic fraction of the biomass. A statistical correlation for the β is provided by
 342 Szargut and Styrylska [36]:

$$\beta = \frac{1.044 + 0.16 \frac{z_H}{z_c} - 0.34493 \frac{z_o}{z_c} \left(1 + 0.0531 \frac{z_H}{z_c} \right)}{1 - 0.4142 \frac{z_o}{z_c}} \quad (29)$$

343 where z_H, z_C, z_O are the weight fractions of the hydrogen, carbon and oxygen in the biomass,
 344 respectively. The chemical exergy of an ideal gas mixture can be expressed as [48]:

$$e_i^{ch} = \sum_i x_i e_{0,i}^{ch} + \bar{R}T_0 \sum_i x_i \ln x_i \quad (30)$$

345 where x_i is the mole fraction of i_{th} component and $e_{0,i}^{ch}$ is the standard chemical exergy of that
 346 component [38]. The exergy balance for a system component is expressed as:

$$\sum \dot{E}_{in} = \sum \dot{E}_{out} + \dot{E}_D \quad (31)$$

347 With the aid of eq. 31 one may find the exergy destruction in system components in order to
 348 define the most probable candidate of optimizing for higher efficiency.

349 Finally, exergy analysis can be completed by defining the efficiency of second law for the power
 350 generation (electrical efficiency) mode and CHP mode as follows:

$$\eta_{II,P} = \frac{\dot{W}_{FC,stack,ac} - \dot{W}_{FC} - \dot{W}_{AC} - \dot{W}_{pump}}{\dot{E}_{in}} \quad (32)$$

$$\eta_{II,CHP} = \frac{\dot{W}_{FC,stack,ac} - \dot{W}_{FC} - \dot{W}_{AC} - \dot{W}_{pump} + (\dot{E}_{22} - \dot{E}_{20})}{\dot{E}_{in}} \quad (33)$$

351 where \dot{E}_{in} is the rate of input exergy and can be defined as below for gasifier coupled SOFC
 352 and digester coupled SOFC respectively:

$$\dot{E}_{in} = e_{biomass}^{ch} + w e_{water}^{ch} + 4.76 m e_{air}^{ch} + \dot{E}_6 + \dot{E}_{19} \quad (34)$$

$$\dot{E}_{in} = e_{biomass}^{ch} + w e_{water}^{ch} + \dot{E}_{19} \quad (35)$$

353 5. Economic analysis

354 5.1. Methodology

355 It seems that investigating the performance of a system without looking at its product cost will
 356 not be accomplished. Integration of economic point of view with thermodynamic seems to be
 357 appealing. Exergoeconomic method firstly **introduced by Tsatsaronis et al. [50]** gets the interest
 358 of researchers who use the method to analyze the thermodynamic systems from the viewpoint of
 359 economic. The purpose of exergoeconomic analysis for a system is to disclose the cost formation
 360 processes and calculate the cost per unit exergy of product streams. To calculate the cost of each
 361 unit exergy stream, the cost balance equation along with the required auxiliary equations are
 362 applied to each component of the systems. For a system component receiving thermal energy and
 363 generating power, the cost balance is written as:

$$\sum \dot{C}_{out,k} + \dot{C}_{w,k} = \sum \dot{C}_{in,k} + \dot{C}_{q,k} + \dot{Z}_{k,PY} \quad (36)$$

$$\dot{C} = c E\dot{x} \quad (37)$$

364 Where, c is the cost per exergy unit and $E\dot{x}$ is the total exergy rate.

365 **In equation (36), $\dot{Z}_{k,PY}$ is** the appropriate charge due to capital investment and operating and
 366 maintenance expenses for each component in a reference year. Note that, the cost of the system
 367 components which are available in an original year is converted from that original time to a same
 368 reference year (year 2013 for present work) with the help of Chemical Engineering Plant Cost
 369 Index (CEPEI).

$$\dot{Z}_k^o = \dot{Z}_k^{o,CI} + \dot{Z}_k^{o,OM} \quad (38)$$

370 Cost at reference year=Original cost \times $\frac{\text{Cost index for the reference year}}{\text{Cost index for the original year}}$

371 Or

$$\dot{Z}_{k,PY}^o = \dot{Z}_k^o \frac{CI_{k,PY}}{CI^o} \quad (39)$$

372

373 The term \dot{Z}_k^o in Eq. (38) can be calculated as [48]:

$$\dot{Z}_k^o = \frac{Z_k CRF \varphi}{N} \quad (40)$$

374 Where φ is the maintenance factor, N is the number of system operating hours in a year and

375 CRF is the Capital Recovery Factor, which can be expressed as [51]:

$$CRF = \frac{i_r (1+i_r)^n}{(1+i_r)^n - 1} \quad (41)$$

376 Here, i_r is the interest rate and n is the system life. The input data used in economic

377 evaluations along with the cost and auxiliary equations for each component of the systems is

378 gathered in Table 5.

379 5.2 Cost evaluation

380 To obtain the cost of all unit exergy streams, the linear system of equations is solved

381 assuming that the cost of unit exergies associated with the input fuel is an input.

382 The exergoeconomic evaluation of the systems is carried out using the thermoeconomic

383 variables, namely, the unit cost of the fuel ($c_{F,k}$), the unit cost of the product ($c_{P,k}$), the cost rate of
384 exergy destruction ($\dot{C}_{D,k}$), the cost rate of exergy loss ($\dot{C}_{L,k}$) and the thermoeconomic factor (f_k).
385 These parameters are calculated using the following relations [48]:

386 6. Results and discussions

387 The effect of decision parameters such as the current density, the temperature difference of
388 SOFC stack on efficiencies of power generating system and CHP system along with net power
389 output, heating capacity and unit product cost of the systems is investigated. Nominal values of
390 current density, where maximum power is achieved, are found to be 0.6 mA/cm² and 0.2
391 mA/cm² for the D-SOFC and G-SOFC systems respectively.

392 Fig. 2 shows the effect of current density on the first and second law efficiencies of the power
393 generating system. Referring to Fig. 2, the current density range is lower for G-SOFC system
394 because of N₂ presence in the gasifier exit, which causes partial pressure of components in stack
395 exit to become less than the D-SOFC system and it affects the cell voltage due to decrease in J_{0a} .
396 Looking again in Fig 2, for constant value of fuel utilization factor both first and second law
397 efficiencies tend to decrease with increasing current density. With increasing 0.6 A/cm² in
398 current density for the D-SOFC system, first and second law efficiencies tend to decrease by an
399 almost 40%. While the reduction is about 30% for the G-SOFC system (with a possible increase
400 of 0.2 A/cm²).

401 Fig. 3a shows the effect of current density on the first and second law efficiencies of the CHP
402 system. The results show that the first law efficiency for the G-SOFC system is in the range of
403 50%- 74% while the efficiency for the D-SOFC system is obtained in the range of 35%-84%.
404 The great difference between the first and second law efficiency values for G-SOFC system is

405 due to the exergy rate within the HRSG system (exergy related to heating).

406 Fig. 3b shows the effect of current density on both the net power output and heating capacity.
407 First important point is that, the net power output is maximized for specific values of current
408 density for both the D-SOFC and G-SOFC systems. **The optimum current density value for G-**
409 **SOFC system is found to be 0.199 A/cm² and the corresponding net power output is calculated**
410 **46.98 kW.** For the D-SOFC system **the optimum current density and net power values are 0.64**
411 **A/cm² and 159.7 kW, respectively.** In addition, it is observed that an increase in the current
412 density increases the heating capacity with an almost same trend of power. **At the optimum**
413 **working points (where the net power output is maximized) the heating capacity values are**
414 **88.3kW and 43.31kW for the G-SOFC and D-SOFC systems, respectively.** Another important
415 aspect of Fig 3b is that although the net power output is higher for the digester based system, the
416 heating capacity is higher for the gasifier system which could be interfered from contemplating
417 in Figs 3a and 3b. Also by an increase in current density, the value of heating capacity increases
418 by up to 83% and 45% for the G-SOFC system and D-SOFC system respectively.

419 Fig. 4 shows the effect of current density on the unit exergy cost of power output (c_w) and hot
420 steam (c_h) for both the D-SOFC and G-SOFC systems. It is revealed that besides the efficiency
421 of D-SOFC is higher it has come at the price of lower unit product cost. Also it is unfolded that
422 for both cases the unit product cost is minimized at some specific point. For the G-SOFC system,
423 the minimum value of unit exergy cost of heating (24.67\$/GJ) occurs at current density of 0.203
424 A/cm² while for the case of D-SOFC system the values are 14.17 \$/GJ and 0.602 A/cm²,
425 respectively. It can be seen that even in the minimum values, the D-SOFC system is cost-
426 efficient by 41.6% and 70.2% for heating unit exergy cost and power unit exergy cost,
427 respectively.

428 Another key parameter having effect on the system performance is the stack temperature
429 difference. Fig. 5 shows the effect of stack temperature difference on second law efficiency of
430 CHP and power generation systems for both the D-SOFC and G-SOFC systems. Second law
431 efficiency for the D-SOFC system is always higher than that for the G-SOFC system but
432 important aspect of this figure is that variation of second law efficiency for power generation
433 mode and CHP mode of G-SOFC tends to decrease after a specific value of temperature. This is
434 actually due to the different composition of inlet gas sent to the anode and consequently different
435 partial pressure of H_2 and H_2O at the anode exit. As the stack temperature difference increases
436 $110^\circ C$, exergetic efficiency for D-SOFC system (in the CHP mode) rises by 52.9% while with
437 an increase of $90^\circ C$ the efficiency of the G-SOFC system increases by 75.3%. In addition, as the
438 temperature difference increases by $80^\circ C$ the second law efficiency of power generation system
439 for the G-SOFC system rises by 25%, the change is an almost the same for the case of D-SOFC
440 with a value of 25.4%.

441 Fig. 6a shows the effect of stack temperature difference on the net power output and heating
442 capacity for the D-SOFC and G-SOFC systems. Comparing the G-SOFC and D-SOFC systems,
443 net power output of G-SOFC system increases by 27.60% with an increase in stack temperature
444 difference and for the D-SOFC system the increase is 25.5%. The trend of heat duty is different
445 as with an increase in temperature of stack results in an increase of exhaust potential of SOFC
446 system which would be utilized in HRSG system. Although the trend seems to be similar for two
447 cases, for the D-SOFC system the percentage of increase is much higher compared to the G-
448 SOFC system.

449 Fig. 6b shows the effect of stack temperature difference on unit exergy costs of heating
450 capacity and power output for both the D-SOFC and G-SOFC systems. **It is revealed that for the**

451 D-SOFC system unit heating product cost is minimized at stack temperature difference of 183 K
452 which corresponds to unit cost of 9.18 \$/GJ while the stack temperature difference and unit cost
453 values are 135.6 K and 16.1 \$/GJ for the G-SOFC system. Also, at the minimum point the D-
454 SOFC system has 75.3% less unit product cost in heating and 91.31% in power unit product cost
455 compared to the G-SOFC system.

456 Finally to give an insight of overall product cost variation with terminal temperature
457 differences of SOFC stack Fig. 6c is presented. It shows that the with a change of about 100 °C in
458 stack temperature difference, the minimum unit product cost of G-SOFC system is more than the
459 maximum unit product cost of D-SOFC system.

460 Fig. 7a and Fig. 7b illustrate the G-SOFC and D-SOFC systems' calculated parameters at
461 nominal operating condition and stack temperature difference equal to 100 °C. According to Fig.
462 7a the second law efficiency for the D-SOFC system is higher with a value of 97.8% in power
463 generation system and 57.7% in CHP system. The difference becomes less in CHP system due to
464 higher heating capacity in G-SOFC system, moreover considering the results illustrated in Fig.
465 7b it is found that the unit product cost for G-SOFC system for power generation system and
466 CHP system is 42.6% and 24.5% more than those for the D-SOFC system respectively. Fig. 7b
467 shows the value of net output power and heating capacity as well as the unit product cost for
468 power and heating for the two proposed systems at the same working conditions. Fig.7b shows
469 that the G-SOFC system has 111 kW less power output and 45.9kW more heating load than the
470 D-SOFC system.

471 One method to illustrate exergy accounting graphically is the Grassman diagram [48]. The
472 width of the arrows entering or leaving the control volume is a quantitative measure of
473 designated parameter. Also for the sake of brevity, values of exergy destruction along with

474 entering and leaving exergy to the whole system are presented in Table 6. According to the Figs.
475 8a and 8b and Table 6, one may conclude that the air heat exchanger and the gasifier are the
476 main sources of irreversibility due to the existence of three sources of irreversibility (temperature
477 difference, chemical reaction along with system friction) within these components in the D-
478 SOFC and G-SOFC systems respectively.

479 Table 7 and Table 8 show the exergoeconomic analysis results for the G-SOFC and D-SOFC
480 systems. Last column of these tables is exergoeconomic factor which a low value of this factor
481 calculated for a major component suggests that cost saving in the entire system *might* be
482 achieved by improving the component efficiency (reducing exergy destruction) even if the
483 capital investment for the component will increase. However, the exergoeconomic factor is not
484 sufficient to explain if a component has to be modified or not. As an example: even if a
485 component has a too low value of exergoeconomic factor (suggesting therefore its substitution
486 with a component of higher performance and higher cost) if the same component elaborates a
487 quantity of fuel which is negligible (and so it has a low value of the so called exergetic factor) it
488 is not worth at all to substitute this component with a better one, as its “exergy role” on the
489 system is simply negligible. The most important components to discuss are the components
490 elaborating a large amount of inlet fuel (so, which have a high value of exergetic factor): only in
491 their case it is interesting to analyze the values of their exergoeconomic factor. Referring to first
492 column of Tables 7 and 8, it can be noted that for the gasifier based system, among the
493 components having higher inlet exergy (SOFC stack, gasifier, after burner and AHX respectively)
494 the SOFC stack and AHX have the highest (88.75%) and lowest (16.02%) exergoeconomic
495 factors, respectively. Therefore, for the case of G-SOFC system, on the one hand, engineers
496 should focus on reducing the investment and operation costs of SOFC stack and on the other

497 hand, they are to reduce the costs associated with exergy destruction for the AHX. The same
498 point can be stated for the case of D-SOFC. Therefore, for the D-SOFC and G-SOFC systems,
499 not only designers should emphasize on decreasing the investment cost of SOFC stack but also
500 reducing the cost associated with exergy destruction within the air heat exchangers should be in
501 priority.

502 Finally at the end, a summary of plant type, fuel type and technology of using biomass to
503 produce gas to feed SOFC power plant in some published works as well as those proposed and
504 obtained in the present work are listed in Table 9. Comparison shows that most of the published
505 works focused on just one technology (gasification to produce syngas or digestion to produce
506 biogas) to utilize the biomass. Also there are few works that comprehensively analyzed the
507 system from the both of thermodynamic and techno economic point of view. However, in the
508 present work, the investigation has been made to fulfill this gap and the thermodynamic and
509 economic analyses are performed to compare two well-known technologies of using biomass in
510 the SOFC system.

511 7. Conclusion

512 A comprehensive thermodynamic and thermoeconomic modeling are performed for two
513 proposed cogeneration systems based on SOFC system fed by municipal solid waste. The inlet
514 fuel for the SOFC is a type of syngas produced by gasification process of municipal solid waste
515 or a biogas produced by digestion process. For the comparison purposes, the systems analyzed in
516 two modes; one-generation system (produce power) and a CHP mode (producing power and
517 heating simultaneously). Parametric studies revealed that stack temperature difference along with
518 current density has crucial effect on systems' performance. Additional conclusions are as

519 follows:

- 520 • D-SOFC system has higher efficiency from the viewpoints of energy and exergy when
521 the systems run to produce power.
- 522 • D-SOFC system has higher second law efficiency in CHP mode; however the first law
523 efficiency for the G-SOFC is higher in this mode.
- 524 • D-SOFC system has more power output but less heating capacity compared to those of
525 the G-SOFC system.
- 526 • For the case of G-SOFC system the gasifier is the main source of irreversibility due to
527 temperature difference and chemical reaction. Meanwhile, air heat exchanger is the
528 second source of irreversibility because of large temperature difference on the both sides
529 of heat exchanger. For the D-SOFC system, air heat exchanger has the most distribution
530 in exergy destruction within the system, the stack and digester are the second and third
531 respectively.

532

533

534

535 **Appendix A**

536 Here, the electrochemical model (for calculating the cell voltage) programming in EES is given
537 in details. The cell voltage can be defined as;

$$V_c = V_N - V_{loss} \quad \text{A}$$

1

538 where, V_N is the Nernst voltage and V_{loss} is the voltage loss which is the sum of three

$$V_{loss} = V_{ohm} + V_{act} + V_{conc} \quad \text{A}$$

2

539 separate voltage losses (ohmic, activation, and concentration losses):

$$V_N = -\frac{\Delta\bar{g}^o}{2F} + \frac{\bar{R}T_{FC,e}}{2F} \ln \left(\frac{a_{H_2}^{Anode,exit} \sqrt{a_{O_2}^{Cathode,exit}}}{a_{H_2O}^{Anode,exit}} \right) \quad \text{A3}$$

540 The Nernst voltage which is accounted as the ideal voltage can be expressed as;

541 In equation (A3), the Gibbs energy difference is related to the overall electrochemical

542 reaction. To determine the actual cell voltage, the voltage losses should be calculated. To

543 calculate the Ohmic loss the following formula is used [52];

$$V_{ohm} = (R_{Int} + \rho_{an}L_{an} + \rho_{cat}L_{cat} + \rho_{ely}L_{ely})j \quad \text{A4}$$

544 where, ρ , L and R_{Int} denote electrical resistivity of a cell component, thickness of a cell

545 component and interconnection resistivity, respectively (See Table A.1).

546 The activation polarization is the sum of those defined for both the anode and cathode as

547 follows;

$$V_{act} = V_{act,a} + V_{act,c} \quad A5$$

$$V_{act,e} = \frac{\bar{R}T_{FC,e}}{F} (\sinh^{-1}(\frac{j}{2j_{0a}})) \quad A6$$

$$V_{act,e} = \frac{\bar{R}T_{FC,e}}{F} (\sinh^{-1}(\frac{j}{2j_{0c}})) \quad A7$$

548 Where j_0 is the exchange current density. Eqs. (A8) and (A9) are used to evaluate the values
549 of the exchange current density for the anode and the cathode, (see variables in Table A.2),

$$j_{0,a} = \gamma_{an} \left(\frac{RT}{2F} \right) e^{\left(-\frac{E_{a,an}}{RT} \right)} \quad A8$$

$$j_{0,c} = \gamma_{cat} \left(\frac{RT}{2F} \right) e^{\left(-\frac{E_{a,cat}}{RT} \right)} \quad A9$$

550 respectively [52].

551 Concentration loss is sum of the losses related to gas concentration occurring in the anode

$$V_{conc} = V_{conc,a} + V_{conc,c} \quad A10$$

552 and cathode.

$$V_{conc,an} = \frac{RT}{2F} \ln \left(\frac{P_{H_2} P_{H_2O,TPB}}{P_{H_2O} P_{H_2,TPB}} \right) \quad A11$$

553 Where

$$V_{conc,cat} = \frac{RT}{4F} \log \left(\frac{P_{O_2}}{P_{O_2,TPB}} \right) \quad A12$$

554 And

555 where the subscript *TPB* denotes the three-phase boundary. To calculate the pressure at the

$$P_{H_2O,TPB} = P_{H_2O,an} + j \frac{RT L_{an}}{2 F D_{an,H_2}^{eff}} \quad A13$$

556 reaction sites, the following equations have been used [52,53]:

$$P_{H_2,TPB} = P_{H_2,an} - j \frac{RT L_{an}}{2 F D_{an,H_2O}^{eff}} \quad A14$$

$$P_{O_2,TPB} = P_{cat} - (P_{cat} - P_{O_2,cat}) \exp\left(j \frac{RT L_{cat}}{4F D_{O_2}^{eff} p_{cat}}\right) \quad A15$$

557

558 where, $D_{H_2}^{eff}$, $D_{H_2O}^{eff}$ and $D_{O_2}^{eff}$ are the effective gaseous diffusivity through the anode (for H₂),

559 anode (for H₂O) and the cathode (for O₂), respectively. The effective gaseous diffusivity can be

$$\frac{1}{D_{an,H_2}^{eff}} = \frac{\varepsilon_{an}}{\tau_{an}} \left(\frac{1}{D_{H_2,K}} + \frac{1}{D_{H_2,H_2O}} \right) \quad A16$$

$$\frac{1}{D_{an,H_2O}^{eff}} = \frac{\varepsilon_{an}}{\tau_{an}} \left(\frac{1}{D_{H_2O,K}} + \frac{1}{D_{H_2O,H_2}} \right) \quad A17$$

$$\frac{1}{D_{cat,O_2}^{eff}} = \frac{\varepsilon_{cat}}{\tau_{cat}} \left(\frac{1}{D_{O_2,K}} + \frac{1}{D_{O_2,N_2}} \right) \quad A18$$

560 calculated as [52,53];

561 Where the porosity (ε) and tortuosity (τ) of electrode materials are estimated to be 0.48 and

562 5.4, respectively. To calculate the effective gaseous diffusivity, combined ordinary and Knudsen

563 diffusion should be defined and calculated using the following equations as[52];

$$D_{H_2,K} = 97r_{pore,an} \sqrt{\frac{T}{M_{H_2}}} \quad A19$$

$$D_{H_2O,K} = 97r_{pore,an} \sqrt{\frac{T}{M_{H_2O}}} \quad A20$$

$$D_{O_2,K} = 97r_{pore,cat} \sqrt{\frac{T}{M_{O_2}}} \quad A21$$

$$D_{H_2,H_2O} = \frac{1.43 \times 10^{-7} T^{1.75}}{\sqrt{M_{H_2,H_2O}} (V_{H_2}^{1/3} + V_{H_2O}^{1/3})^2 P} \quad A22$$

$$D_{O_2,N_2} = \frac{1.43 \times 10^{-7} T^{1.75}}{\sqrt{M_{O_2,N_2}} (V_{O_2}^{1/3} + V_{N_2}^{1/3})^2 P} \quad A$$

23

564

565 Where M is molecular weight of species, V represents diffusion volume of species. Meanwhile,

566 pore radius value (r_{pore}) is estimated to be $0.5 \mu m$.

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810 **Figures' caption**

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812 Figure 1 schematic diagram of CHP system based on a) digester coupled SOFC, b) gasifier coupled
813 SOFC

814 Figure 2 Effect of current density on first and second law efficiencies of power generating system

815 Figure 3 Effect of current density on a) first and second law efficiencies of CHP system b) net power
816 output and heating value of CHP system

817 Figure 4 effect of current density unit product cost of heat and power

818 Figure 5 Effect of stack temperature difference on second law efficiency of power generating and CHP
819 system

820 Figure 6 effect of stack temperature difference on a) net power output and heating capacity b) unit
821 product cost of heating and power c) total product cost of digester coupled SOFC and gasifier coupled
822 SOFC

823 Figure 7 Comparison of a) first and second law efficiencies of systems as well as the total product cost
824 b) net output power and heating load of systems as well as the their product costs

825 Figure 8 Grassman diagram of the a) digester coupled CHP system based on SOFC (D-SOFC) b)
826 gasifier coupled CHP system based on SOFC (G-SOFC)

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835 **Tables' Caption**

836 Table 1. Input data for the SOFC systems [34,37,45,55]

837 Table 2 Ultimate analysis of MSW [39]

838 Table 2. The comparison of the component percentages in the producer gas obtained from the gasification
839 in the present work and those reported in the literature, for wood with a moisture content of 16% and for a
840 gasification temperature of 1100K.

841 Table 3. Comparison of results obtained from the present work with the experimental values reported by
842 Tao et al. [43]

843 Table 4 Input data * and cost and auxiliary equations for each component [11,33]

844 Table 6. Exergy and exergy destruction rates for two proposed SOFC systems

845 Table 7 Exergoeconomic analysis results for the gasifier coupled SOFC

846 Table 8 Exergoeconomic analysis results for the Digester coupled SOFC

847 Table A.1. Material Resistivity used for ohmic voltage loss estimation [52]

848 Table A.2. Parameters correspond to anode and cathode sides material [52]

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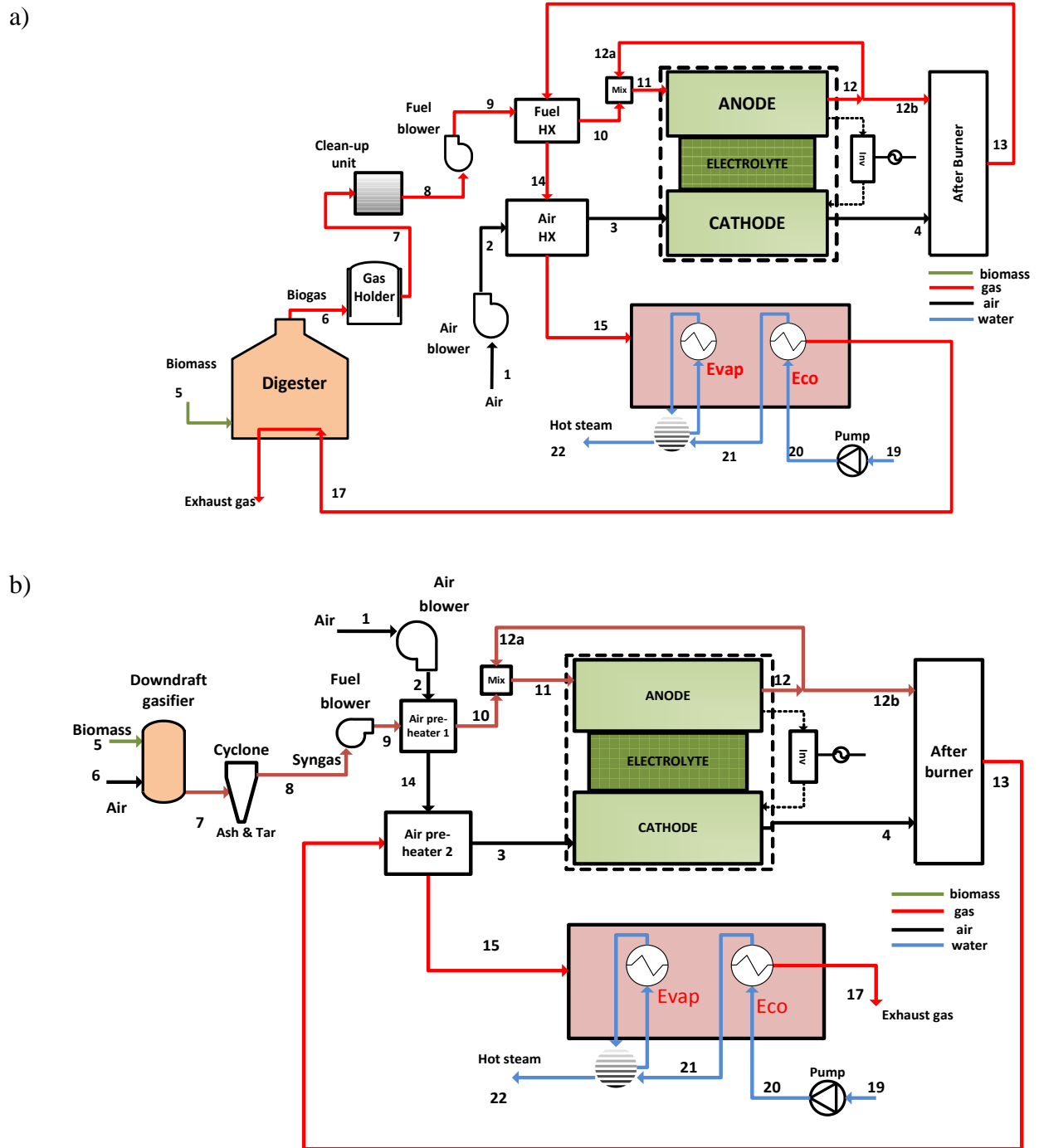


Figure 1 schematic diagram of CHP system based on a) digester coupled SOFC, b) gasifier coupled SOFC

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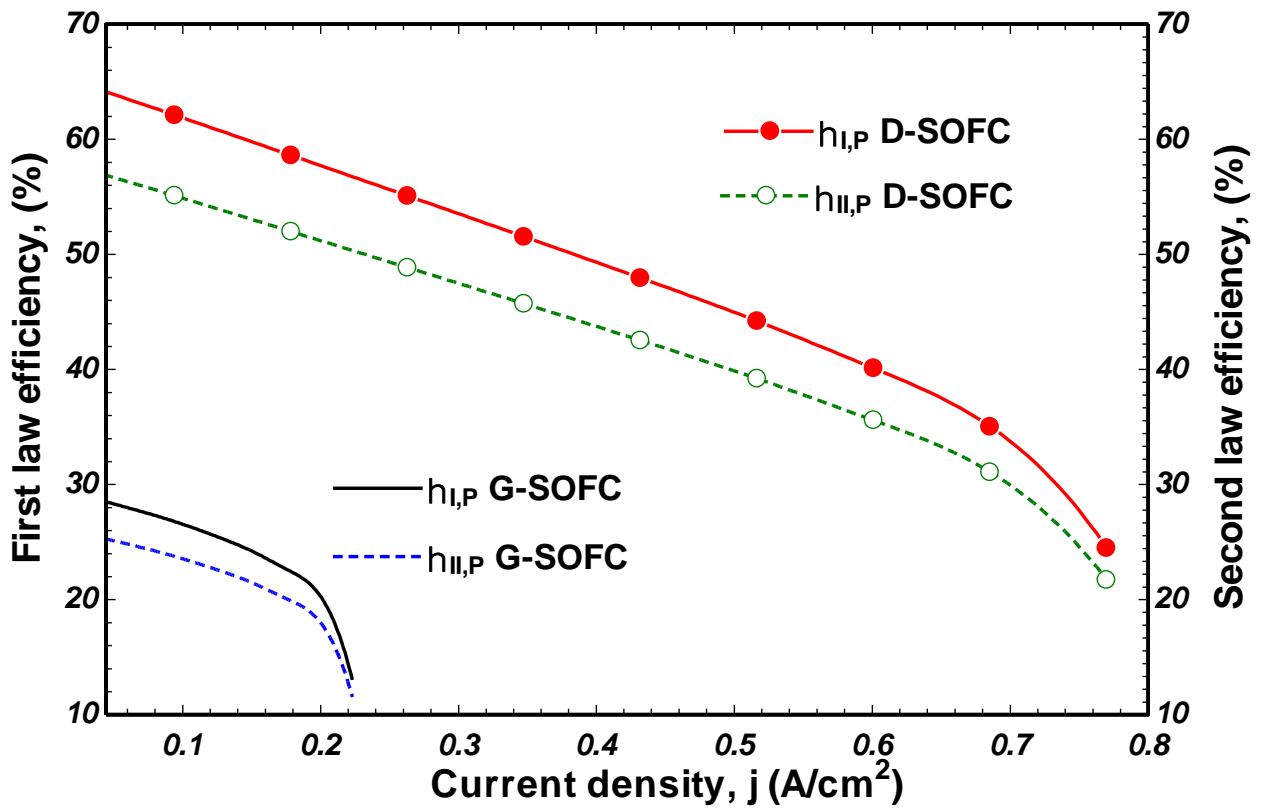
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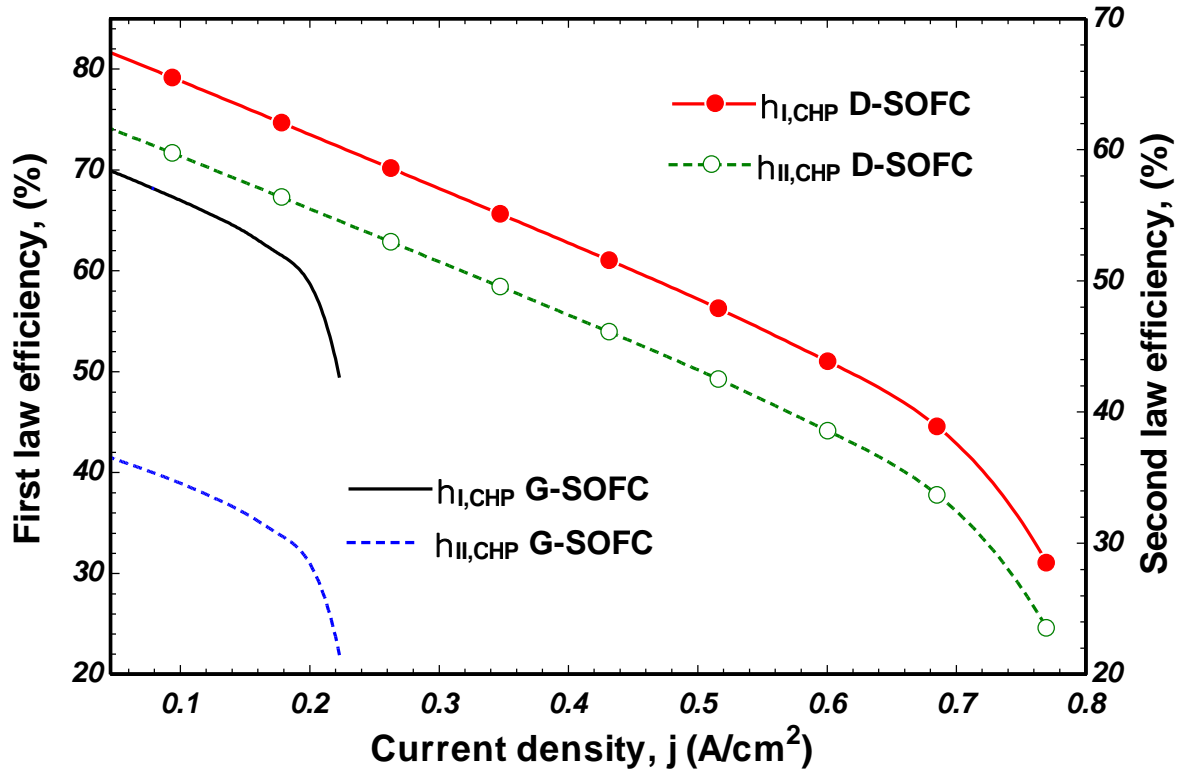
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Figure 2 Effect of current density on first and second law efficiencies of power generating system

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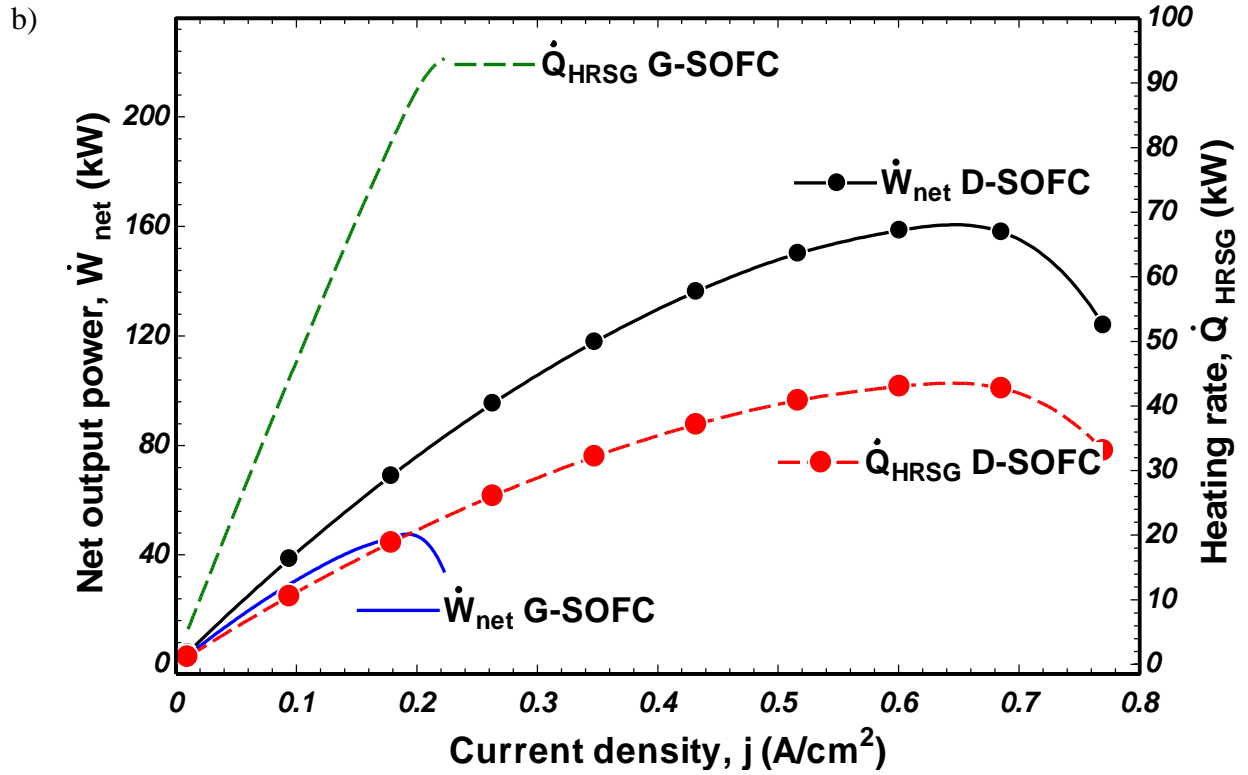


Figure 3 Effect of current density on a) first and second law efficiencies of CHP system b) net power output and heating value of CHP system

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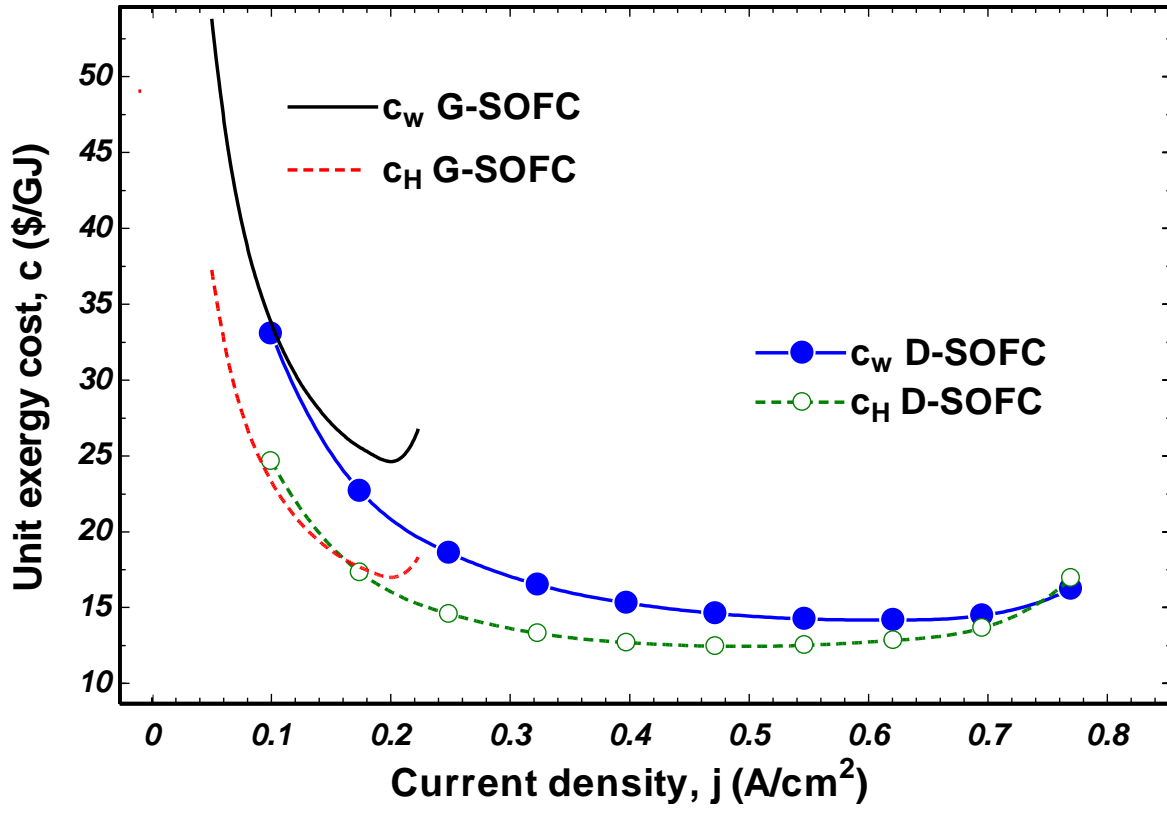


Figure 4 effect of current density unit product cost of heat and power

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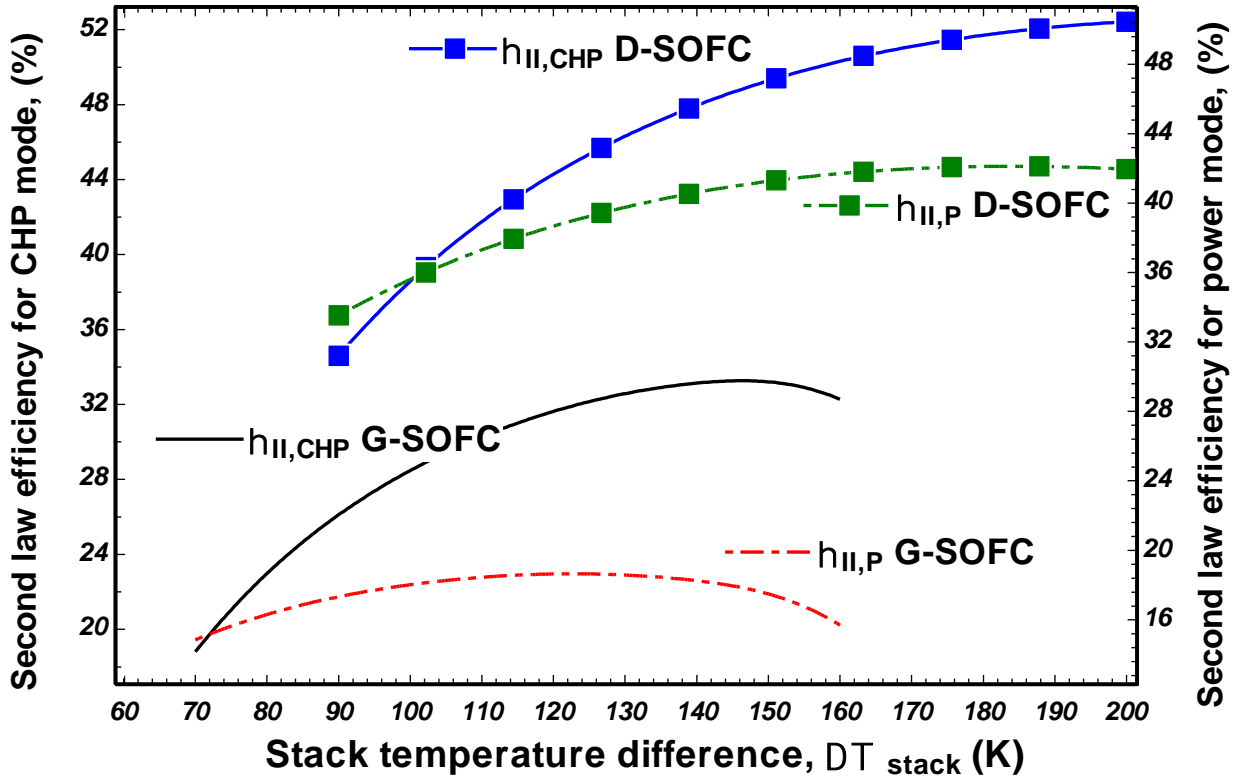
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892 Figure 5 Effect of stack temperature difference on second law efficiency of power generating and CHP
 893 system

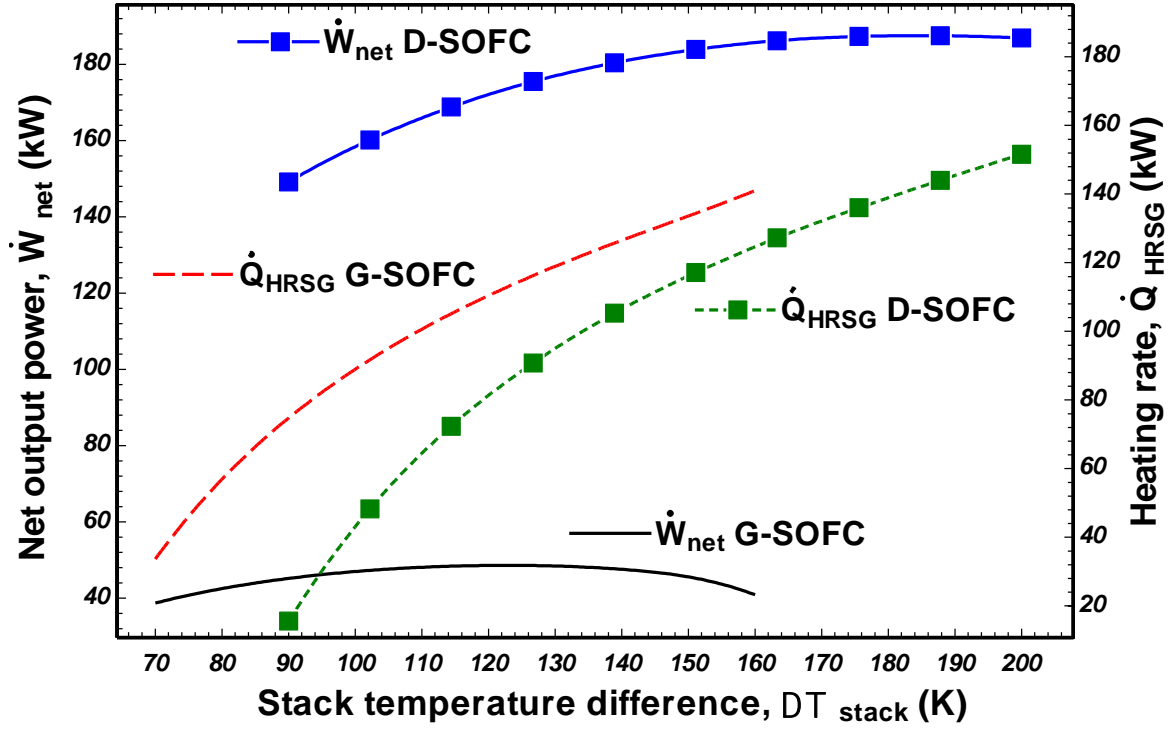
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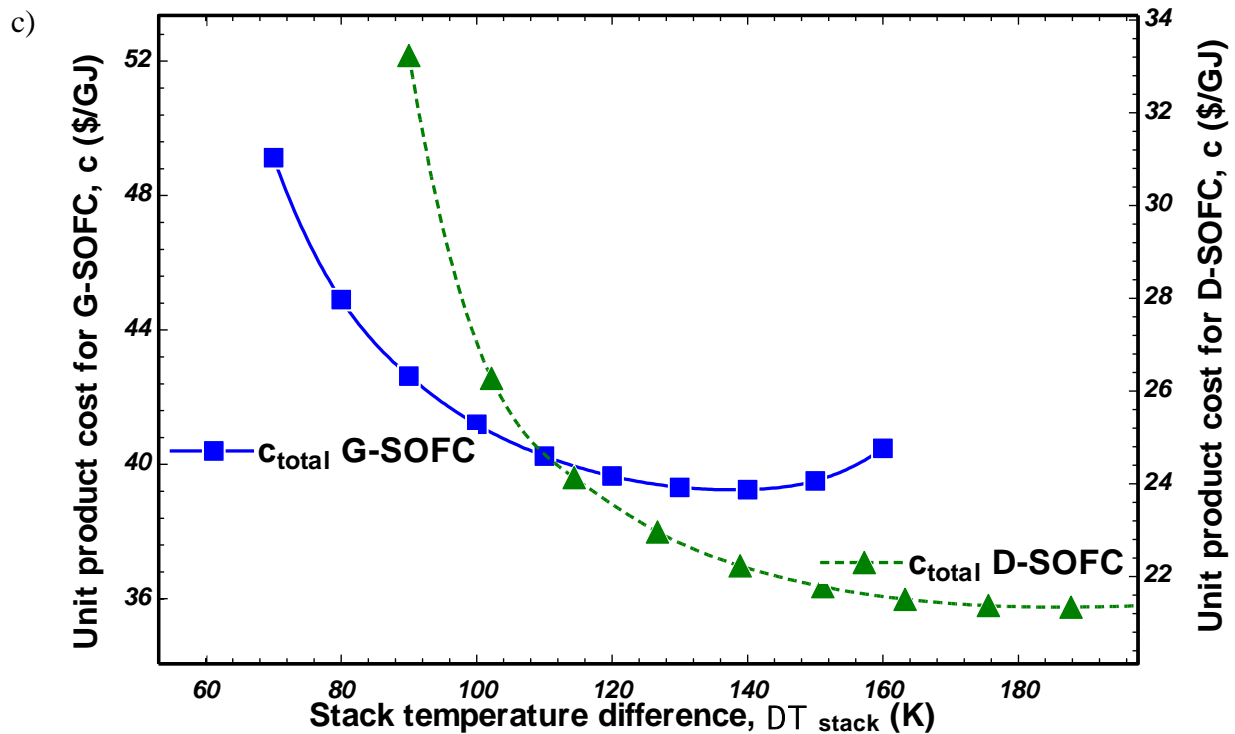
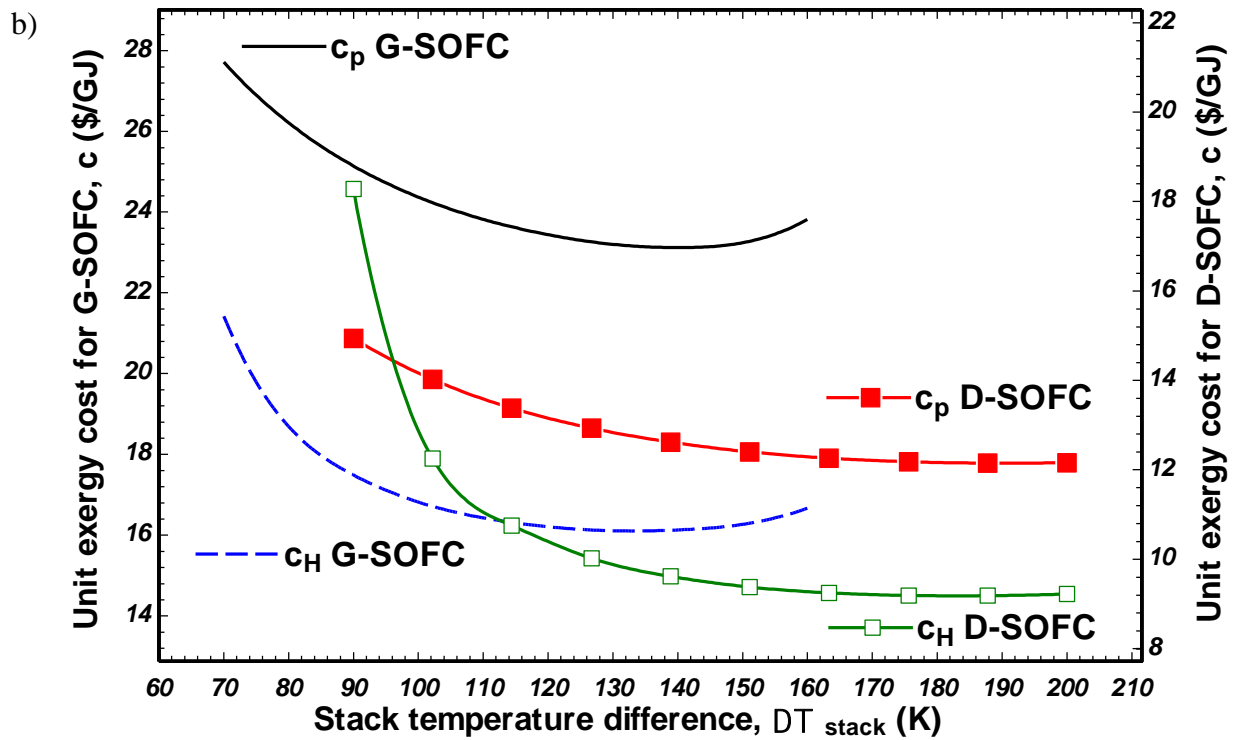


Figure 6 effect of stack temperature difference on a) net power output and heating capacity b) unit product cost of heating and power c) total product cost of digester coupled SOFC and gasifier coupled SOFC

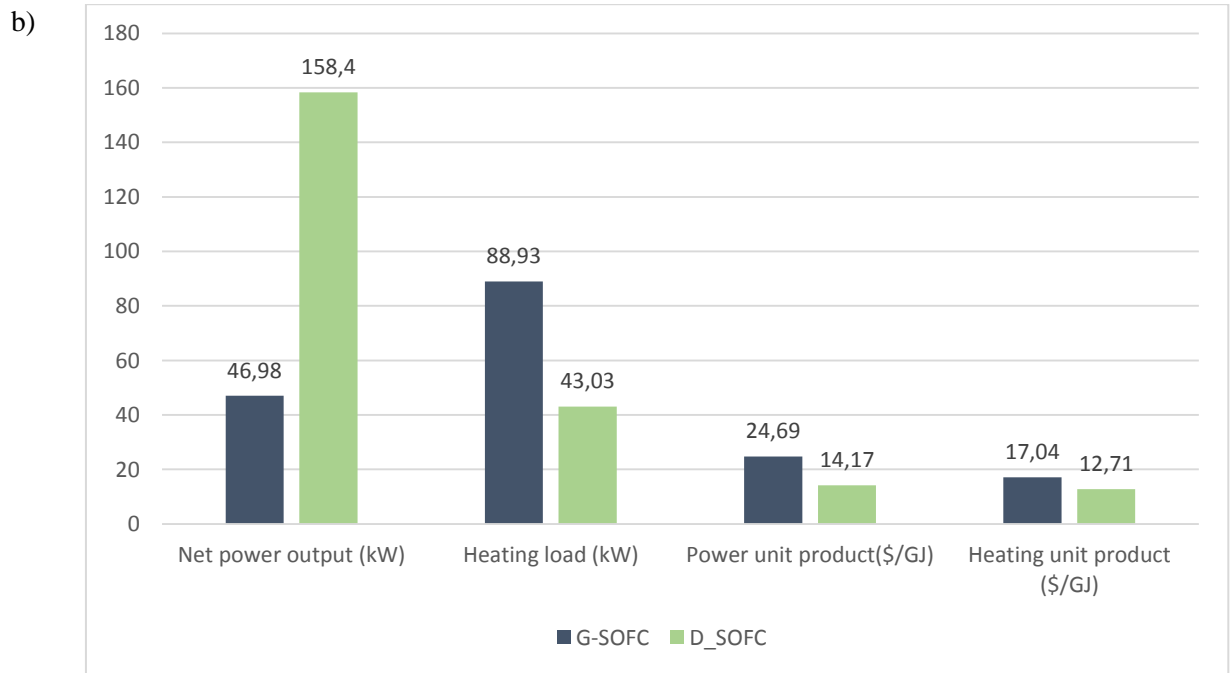
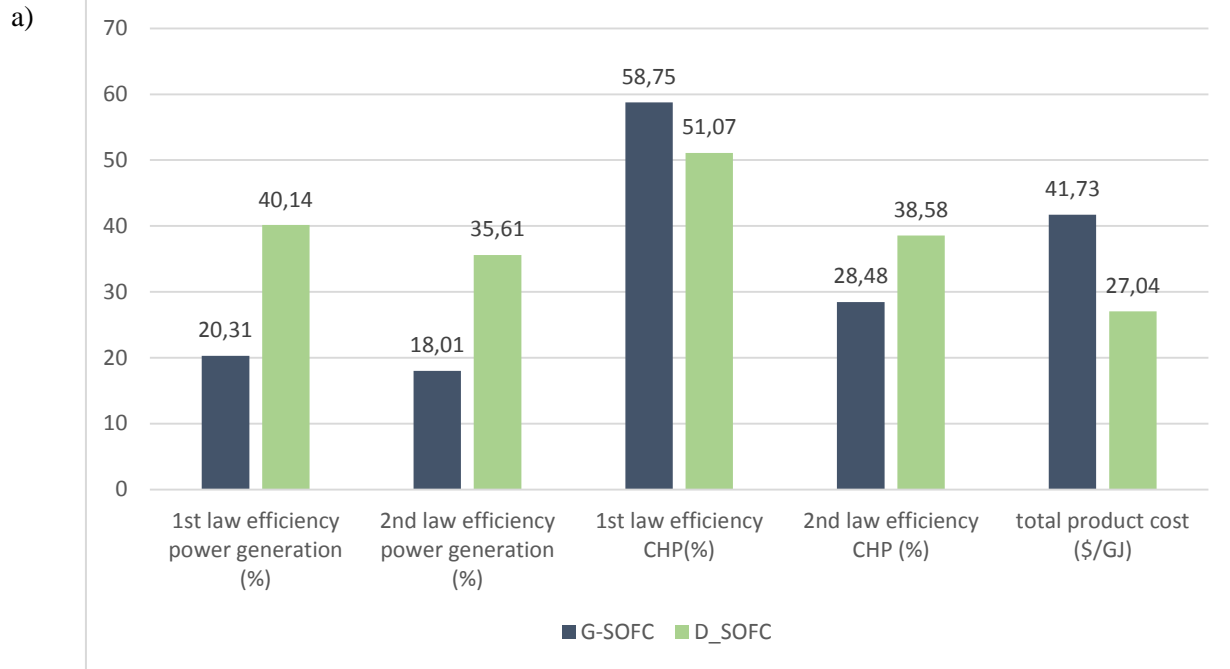
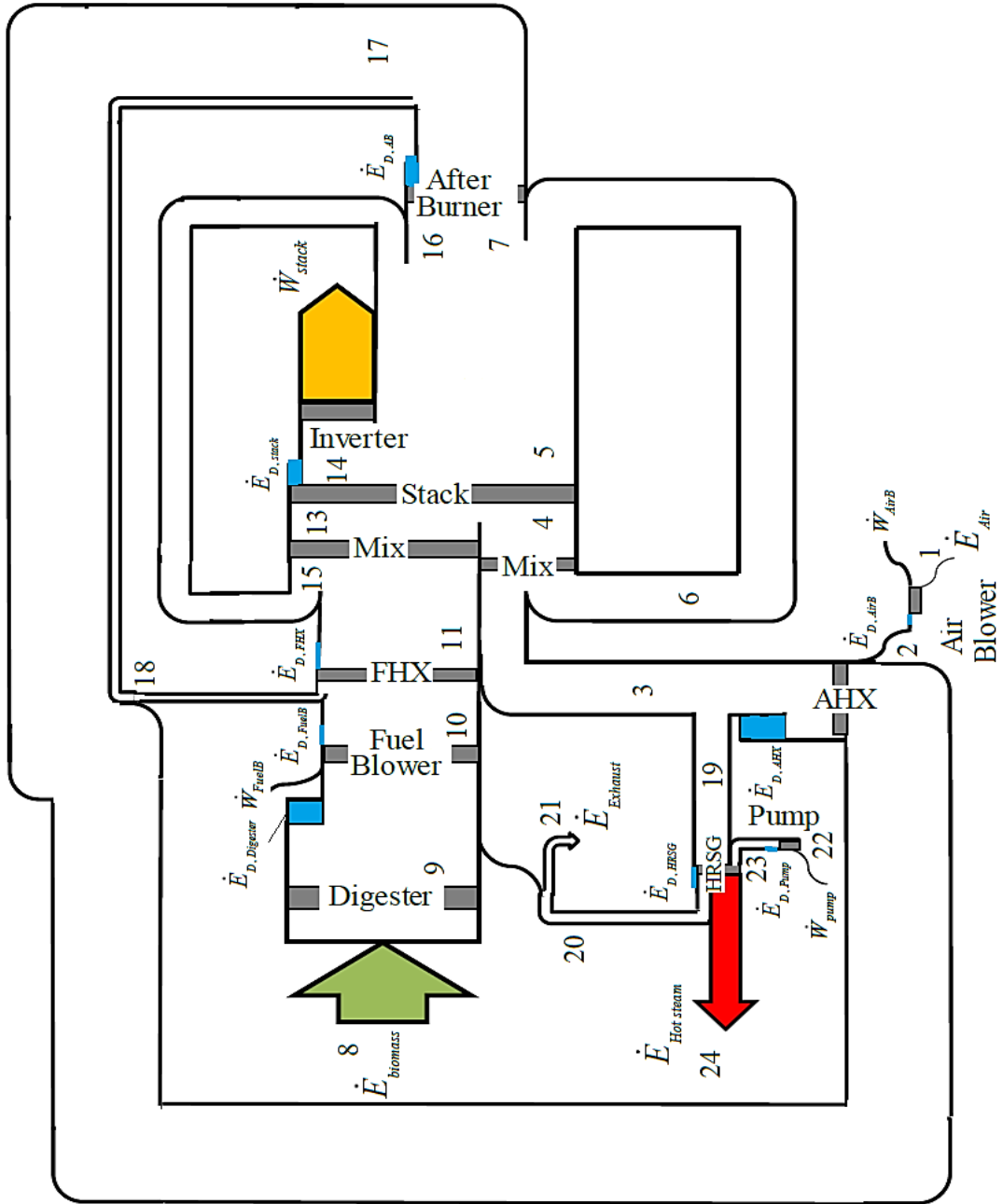


Figure 7 Comparison of a) first and second law efficiencies of systems as well as the total product cost b) net output power and heating load of systems as well as the their product costs

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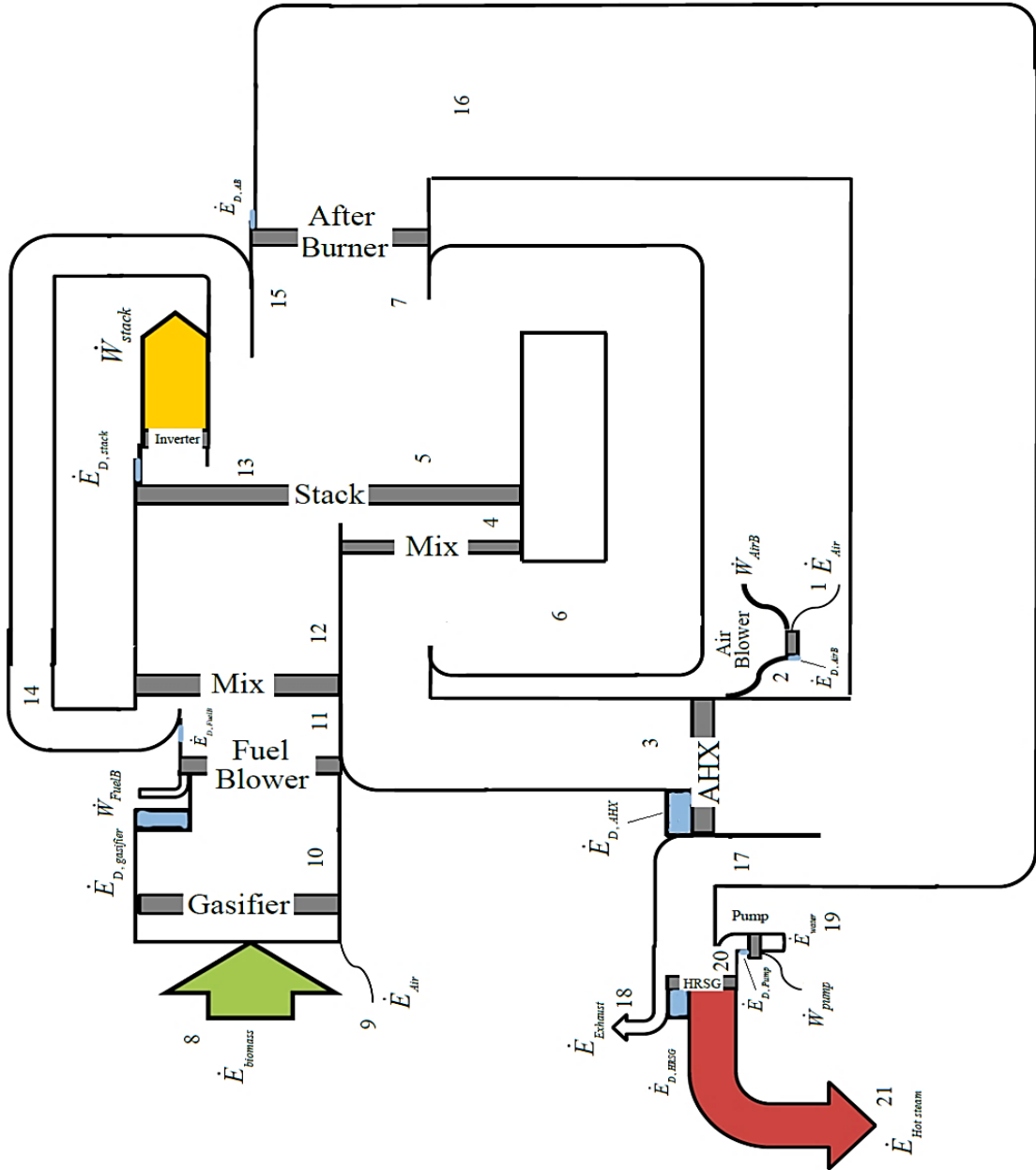


Figure 8 Grassman diagram of the a) digester coupled CHP system based on SOFC (D-SOFC) b) gasifier coupled CHP system based on SOFC (G-SOFC)

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903 **Tables**

Table 1. Input data for the SOFC systems [34,37,45,55]

SOFC system			Unit
	Temperature difference between stack inlet and outlet	100	K
	Fuel utilization factor for whole SOFC	0.80	-
	Active surface area	50	cm²
	DC-AC inverter efficiency	97	%
	Base inlet temperature to SOFC	973.15	K
	Steam to carbon ratio	2	-
	Thickness of anode	500	μm
	Thickness of cathode	50	μm
	Thickness of electrolyte	10	μm
	Fuel blower isentropic efficiency	80	%
	Air blower isentropic efficiency	80	%
	Pump isentropic efficiency	80	%
	Number of cells	11,000	-
	Afterburner combustion efficiency	99	%
	Stack pressure drop	2	%
	Heat exchangers pressure drop	3	%
	Afterburner pressure drop	5	%
Digester			
	Work needed to drive the fan	0	kW
	Exit temperature (Thermophilic digester)	328	K
	Amount of volatile solid	95	%
	Amount of destruction in digester	70	%
Gasifier			
	Gasification temperature	1073.15	K
	Heat loss from gasifier	0	%
	Air inlet temperature	298	K
	Biomass inlet temperature	298	K
HRSG			
	Pinch point temperature difference	15	K
	Steam pressure	10	bar
	Water pump isentropic efficiency	0.75	-

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Table 2 Ultimate analysis of MSW [39]

Biomass	C	H	N	S	O	Ash	Higher heating value (kJ/kmol)
Municipal solid waste	47.6	6	1.2	0.3	32.9	12	433034

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Table 3. The comparison of the component percentages in the producer gas obtained from the gasification in the present work and those reported in the literature, for wood with a moisture content of 16% and for a gasification temperature of 1100K.

Constituent	Present model	Experiment [40]	Jarungthammachote equilibrium model [42]
Hydrogen	17.15	15.5	18.04
Carbon monoxide	19.28	19.1	17.86
Methane	0.55	1.1	0.11
Carbon dioxide	10.81	11.4	11.84
Nitrogen	52.21	52.9	52.15

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Table 4. Comparison of results obtained from the present work with the experimental values reported by Tao et al. [43]

Current density (A/m ²)	Cell voltage (V) (Present work)	Cell voltage (V) (Tao et al.)	Error (%)	Power density (W/m ²) (Present work)	Power density (W/m ²) (Tao et al.)	Error (%)
2000	0.742	0.76	-2.368	0.148	0.15	-1.333
3000	0.684	0.68	0.588	0.205	0.21	-2.381
4000	0.634	0.62	2.258	0.253	0.26	-2.692
5000	0.582	0.57	2.105	0.294	0.295	-0.339
6000	0.547	0.52	5.192	0.328	0.315	4.127

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Table 5 Input data* and cost and auxiliary equations for each component [11,33]

Auxiliary equations	Cost equations	Component
$c_{biomass}=2$	$Z_{gasifier}=1600 \left(\dot{m}_{drybiomass} \right)^{0.67}$	Gasifier
$c_{biomass}=2$	$Z_{Digester}=350000 \left(\frac{\dot{V} T}{21000} \right)^{0.75}$	Digester
$\dot{C}_{14}/\dot{E}_{14}=\dot{C}_{33}/\dot{E}_{33}$ $\dot{C}_5/\dot{E}_5=\dot{C}_{33}/\dot{E}_{33}$	$Z_{SOFC}=A_a N_{FC} (2.96 T_{FC,e} - 1907)$	SOFC stack
-	$Z_{AB}=\frac{46.08\dot{m}_7}{(0.955-(P_{17}/P_7))} \left(1+e^{(0.018T_{17}-26.4)} \right)$	Afterburner
$c_1=0$	$Z_{AC}=91562 \left(\dot{W}_{AC}/455 \right)^{0.67}$	Air compressor
$c_{10}=c_F$	$Z_{FC}=91562 \left(\dot{W}_{FC}/455 \right)^{0.67}$	Fuel compressor
$c_{20}=0$	$Z_{P1}=1785 f_n \left(\dot{W}_P/1 \right)^{0.71}$ $f_n=1+(0.2/(1-\eta_p))$	Pump
$\dot{C}_{17}/\dot{E}_{17}=\dot{C}_{18}/\dot{E}_{18}$	$Z_{FHX}=130 (A_{FHX}/0.093)^{0.78}$	FHX
$\dot{C}_{18}/\dot{E}_{18}=\dot{C}_{19}/\dot{E}_{19}$	$Z_{AHX}=390 (A_{AHX}/0.093)^{0.78}$	AHX
$\dot{C}_{19}/\dot{E}_{19}=\dot{C}_{20}/\dot{E}_{20}$	$Z_{HRSG}=6570 \left[\left(\dot{Q}_{eco}/\Delta T_{eco} \right)^{0.8} + \left(\dot{Q}_{eva}/\Delta T_{eva} \right)^{0.8} \right] +$ $21276 \dot{m}_{steam} + 1184.4 \left(\dot{m}_{gas} \right)^{1.2}$	HRSG
-	$Z_{inv}=100000 \left(\dot{W}_{SOFC,DC}/500 \right)^{0.7}$	Inverter

* $i_r=0.12$, $n=20$ years, $\tau=8000h$, T =retention time in digester, $c_F=2\$/GJ$ (biomass)

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Table 6. Exergy and exergy destruction rates for two proposed SOFC systems			
Digester Coupled SOFC		Gasifier Coupled SOFC	
Exergy rate	$\dot{E}X(kW)$	Exergy rate	$\dot{E}X(kW)$
Input exergy	462.6	Input exergy	290.3
Destruction		Destruction	
Digester	41.98	Gasifier	74.42
Air blower	7.018	Air blower	3.591
Fuel blower	0.128	Fuel blower	0.3248
AHX	78.67	AHX1	16.29
FHX	4.716	AHX2	35.3
Stack	42.29	Stack	9.304
AB	30.67	AB	12.95
Mixing unit	2.804	Mixing unit	1.018
HRSB	4.840	HRSB	9.786
Pump	0.049	Pump	0.010
Exergy associated with heating	26.27	Exergy associated with heating	54.29
Exergy associated with power	164.3	Exergy associated with power	49.81
Exergy loss	58.7	Exergy loss	26.93
$U_f=0.80, T_{stack} = 800\text{ }^\circ\text{C}$			

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Table 7 Exergoeconomic analysis results for the gasifier coupled SOFC (G-SOFC)

Components	\dot{E}_r (kW)	$C_{F,k}$ (\$/GJ)	$C_{P,k}$ (\$/GJ)	$\dot{C}_{D,k}$ (\$/h)	$\dot{C}_{L,k}$ (\$/h)	\dot{Z}_k (\$/h)	$\dot{C}_{D,k} + \dot{C}_{L,k} + \dot{Z}_k$ (\$/h)	f_k (%)
SOFC stack	437	16.8	20.05	0.5627	0	4.430	5.00	88.75
Air heat exchanger 1	15.36	4.751	27.87	0.2142	0	0.022	0.236	9.41
Air heat exchanger2	221.9	21.14	26.52	2.974	0	0.567	3.541	16.02
Air blower	15.62	20.05	34.01	0.2592	0	0.345	0.604	57.15
Fuel blower	6.25	20.05	29.92	0.0234	0	0.187	0.210	88.87
After burner	298.8	20.05	21.14	0.9345	0	0.191	1.126	16.99
Gasifier	260.8	2	3.968	0.5358	0	0.839	1.375	61.04
HRSG	37.06	21.14	33.84	0.7448	2.05	0.502	3.296	15.23
pump	0.0405	20.05	67.06	0.0007	0	0.004	0.005	85.78

$U_f=0.80, T_{stack} = 800\text{ }^\circ\text{C}$

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Table 8 Exergoeconomic analysis results for the digester coupled SOFC (D-SOFC)

Components	\dot{E}_r (kW)	$C_{F,k}$ (\$/GJ)	$C_{P,k}$ (\$/GJ)	$\dot{C}_{D,k}$ (\$/h)	$\dot{C}_{L,k}$ (\$/h)	\dot{Z}_k (\$/h)	$\dot{C}_{D,k} + \dot{C}_{L,k} + \dot{Z}_k$ (\$/h)	f_k (%)
SOFC stack	911.8	9.172	11.04	1.397	0	5.834	5.846	76.06
Air heat exchanger	441.8	11.87	15.28	3.361	0	1.130	4.452	24.51
fuel heat exchanger	12.81	11.87	19.44	0.2015	0	0.058	0.220	8.732
Air blower	30.53	11.04	20.73	0.2788	0	0.541	0.820	66.01
Fuel blower	0.753	11.04	33.46	0.0051	0	0.045	0.050	89.88
After burner	561.5	11.04	11.87	1.219	0	0.370	1.589	23.31
digester	444.4	2	2.277	0.3039	0	0.097	0.401	24.22
HRSG	18.04	11.87	25.23	0.2068	2.50	0.428	3.143	13.62
pump	0.014	11.04	64.49	0.0001	0	0.002	0.002	93.11

$U_f=0.80, T_{stack} = 800\text{ }^\circ\text{C}$

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Table 9. Comparison of the published works results with those of the present work

Research	Plant Type; Plant Scale	Fuel; Process; Gas	Thermodynamic analysis		Economic analysis	Year
			1 st law Efficiency (%)	2 nd law Efficiency (%)	Unit Product Cost	
Omosun et al. [14]	Cogeneration (Heat and power);200kW _{el}	Biomass; Fluidized bed gasification; Syngas (17% H ₂ ,13% CO,11% CO ₂ ,4% CH ₄ , 15% H ₂ O,40% N ₂)	Electrical: 22.6% Total: 59.6%	-	2.9k £/kW _e 1.1k £/kW _T	2004
Piroonlerkgul et al. [26]	Mono generation (power); small size (50-60kW)	Biogas (60% CH ₄ ,40% CO ₂)	Electrical: 59%	-	-	2008
Farhad et al. [56]	Cogeneration (Heat and power);small size (~1kW _{el})	Biogas (60.8% CH ₄ ,34.8% CO ₂ , 2.4% N ₂ ,1.2% O ₂ ,0.01% water)	Electrical: 22.6% Total: 59.6%	-	-	2010
Bang-Møller et al. [57]	Cogeneration (Heat and power) Hybrid system; medium size (~300kW)	Wood; two-stage gasification; syngas	Electrical: 58.2% Total: 87.5%	Electrical: 50.4% Total: 53.4%	-	2011
Santarelli et al. [29]	Cogeneration (Heat and power); medium size (scale-up 250kW)	Wheat straw; Two- stage anaerobic digester; biogas (55% CH ₄ ,10% H ₂ ,35% CO ₂)	Electrical: 49.2% Total: 75.1%	-	-	2012
El-Emam et al. [58]	Mono generation (power) Hybrid system; Large size (~30MW)	Coal gasification; syngas(10.6% CO ₂ ,51.6 % CO,0.1% CH ₄ ,35.1% H ₂ ,2.6% N ₂)	Electrical: 38.1%	Electrical: 27%	-	2012
Wongchanapai et al. [59]	Mono generation (power); small scale (5kW)	Wood; two-stage gasification; Syngas(18.7% H ₂ ,21.87 CO,0.22% CH ₄ ,10.51% CO ₂ ,47.3% N ₂)	Electrical: 38.9%	Electrical: 37.4%	-	2012
Campitelli et al. [60]	Mono generation (power); small scale (1-2 kW)	Wood; autothermal gasification; syngas (17.8% H ₂ ,9.4% CO,14 % CO ₂ ,19.8% H ₂ O,0.1% CH ₄)	Electrical: 24.7%	-	-	2013
Arteaga-Pérez et	Cogeneration	Sugar cane/rice husk;	Total:	Total:	-	2013

al. [61]	(Heat and power); medium size (456.5kW)	bubbling fluidized bed gasification; syngas (17.5% H ₂ O, 40.1% N ₂ , 0.6% O ₂ , 6.5% CH ₄ , 13.6% CO, 12.4% CO ₂ , 9.3% H ₂)	55.48%	32.01%		
Morandin et al. [62]	Cogeneration (Heat and power) Hybrid system; small size (~40kW)	Wood; fluidized bed gasifier/Viking gasifier; syngas	Total: 63% (For Viking gasifier case) Total: 58% (For fluidized bed gasifier)	-	15000\$/kW (For Viking gasifier case) 8000 \$/kW (For fluidized bed gasifier)	2013
Trendewicz and Braun [31]	Cogeneration (Heat and power); small, medium and large sizes (330kW, 1530kW, 6140kW)	Waste water; anaerobic digestion; biogas (56.6% CH ₄ , 36.7% CO ₂ , 5.8% H ₂ O)	Electrical: 51.6% Total: 87.5%	-	3584 \$/kW (large size) 3916 \$/kW (medium size) 5780 \$/kW (small size)	2013
Gandiglio et al. [28]	Trigeneration (Heat, power and Algae production); small scale (2kW _{el})	Waste water; anaerobic digester; biogas (60% CH ₄ , 40% CO ₂)	Electrical: 52.56% Total: 85.93%	-	-	2014
Siefert and Litster [33]	Mono generation (power)	Waste water; anaerobic digestion; Biogas (45% CH ₄ , 40% CO ₂ , 15% H ₂ O)	-	Electrical: 58%	3610 \$/kW	2014
Ozcan and Dincer [63]	Trigeneration (Heat, power and cooling); medium size (145.5kW)	ThermoChem bubbling fluidized bed gasifier; syngas (43.3% H ₂ , 9.2% CO, 28% CO ₂ , 5.6% H ₂ O, 4.7% CH ₄)	Electrical: 42.2% Total: 78.8%	Electrical: 36.5% Total: 50.6%	-	2014
Caliandro et al. [64]	Mono generation (power) Hybrid system; small and medium sizes (103kW and 8000kW)	Wood; fast internally circulating fluidized bed gasifier/circulating fluidized bed gasifier/Viking gasifier; syngas	Electrical: 71% (circulating fluidized bed) Electrical: 68.7% (Viking gasifier)	-	0.3 \$/kWh (circulating fluidized bed-medium size) 1.1 \$/kWh (Viking gasifier-small size)	2014
Jia et al. [22]	Cogeneration (Heat and power)	Wood; downdraft gasifier; syngas	Electrical: 42.94%	-	-	2015

	power) Hybrid system; small size (~40kW)		Total: 67.59%			
Curletti et al. [65]	Mono generation (power); large size (1MW _{el})	Waste water; anaerobic digestion; biogas	Electrical: 58.85%	-	2.75 M\$ Net present value	2015
Mehr et al. [34]	Mono generation (power); medium size (~400kW)	Biogas (50-70%CH ₄ ,30%-50% CO ₂)	Electrical: 46.09%	Electrical:43 .92%	19.53 \$/GJ	2015
Lv et al. [66]	Mono generation (power) Hybrid system; small size (182kW)	Wood; gasification; syngas(4.53%CH ₄ ,23.6 4% H ₂ ,13.87%CO,17.9 2%CO ₂ ,40.04%N ₂)	Electrical: 60.78%	-	-	2016
Present work	Cogeneration (Heat and power); small size	Municipal solid waste; downdraft gasifier; Syngas	Electrical: 20.31%	Electrical: 18.01%	24.69\$/GJ	-
			Total: 58.75%	Total: 28.48%		
		Municipal solid waste ;anaerobic digestion; biogas	Electrical: 40.14%	Electrical:35 .61%	14.17\$/GJ	-
			Total: 51.07%	Total: 38.58%		

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Table A.1. Material Resistivity used for ohmic voltage loss estimation [52]

Component	Material	Resistivity	Thickness (mm)
Anode	Ni/YSZ cermet	$\rho_{an}=2.98 \times 10^{-5} \exp\left(\frac{-1392}{T_{FC,e}}\right)$	0.5
Cathode	LSM-YSZ	$\rho_{cat}=8.114 \exp\left(\frac{600}{T_{FC,e}}\right)$	0.05
Electrolyte	YSZ	$\rho_{ely}=2.94 \times 10^{-5} \exp\left(\frac{10350}{T_{FC,e}}\right)$	0.01
Interconnection	Doped LaCrO3	0.0003215	-

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Table A.2. Parameters correspond to anode and cathode sides material [52]

Component	Parameter	Value	Unit
Anode	Pre-exponential factor for anode, γ_{an}	6.54×10^{11}	A/m^2
	Activation energy for anode, $E_{a,an}$	140,000	J/mol
Cathode	Pre-exponential factor for cathode, γ_{ca}	2.35×10^{11}	A/m^2
	Activation energy for cathode, $E_{a,cat}$	137,000	J/mol

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