

# Study of a small heat and power PEM fuel cell system generator

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## Abstract

A micro-cogenerator based on a natural gas reformer and a PEMFC is studied in its entirety, pointing out the links between different sub-systems. The study is conducted within the EPACOP project, which aims at testing PEMFC systems on user sites to evaluate development and acceptance of this technology for small stationary applications. Five units were installed from November 2002 to May 2003 and have been operated until now, in real life conditions. They deliver up to 4 kW of AC power and about 6 kW of heat.

Center for Energy and Processes (CEP), one of the scientific partners, processes and analyses the experimental data from the five units, running in different regions of France. This database and the study of the flowsheet enable to propose changes to enhance the efficiency of the system composed of a steam reforming, a shift and a preferential oxidation reactor, a fuel cell stack and heat exchangers. The steady state modelling and optimisation of the system is done with Thermoptim®, a software developed within CEP for applied thermodynamics.

At constant power, main targets are to decrease natural gas consumption, to increase heat recovery and to improve the water balance. This study is made using the pinch point analysis, at full load and partial load.

Main results of this study are different system configurations that allow improvement of gross electrical and thermal efficiency and enable to obtain a positive water balance.

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## 1. Introduction

Small cogeneration PEMFC systems are gaining interest among power facilities and governmental organisations, especially in Japan and North America. In Europe, domestic gas and electricity suppliers, but also boiler manufacturers, are testing, adapting and trying to improve these devices in order to assess their ability to stick to efficiency, reliability and cost targets that meet European electricity market needs. These systems are fed with natural gas (NG) and deliver 1–10 kW of AC low voltage (LV) power and they are usually connected to a LV grid. Stand-alone systems are fed with propane.

This study is conducted in Center for Energy and Processes (CEP), in the frame of a French research project, named EPACOP,<sup>1</sup> led by Gaz de France, co-funded by ADEME, with collaboration of three CNRS laboratories in Nancy: LEMTA, GREEN and LSGC.

CEP has a 10 year experience in PEMFC stacks evaluation and PEMFC systems analysis. Since the development of its fuel cell test bench, it has worked on several French projects (CAR-BUPAC and SAPAREF) and European projects (FEVER, PMFP and PVFCSYS).

### 1.1. State of the art of NG-fed systems: what is a small cogeneration PEMFC system made of?

The system studied is mainly composed of a fuel processor, a stack and an electric compartment that contains converters, possibly batteries and the operation control system. These three sub-systems are usually put in three different compartments. As seen in Fig. 1, a fourth sub-system, commonly called “heat recovery sub-system” or “thermal management sub-system” links the two first sub-systems. It is not an additional compartment because water-cooling circuits and heat exchangers are deeply overlapped with fuel processor and stack sub-systems.

The zoom shows the fuel processor, which is most often made of a reforming reactor, one or two water shift reactors (shift) and a preferential oxidation reactor (Prox).

For stationary applications, the reforming reactor is a steam reforming reactor (SRR) or an autothermal reformer (ATR).

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<sup>1</sup> Acronym of “Expérimentation de Piles À Combustible de petite taille sur sites Opérationnels” (testing small fuel cells on customer sites).

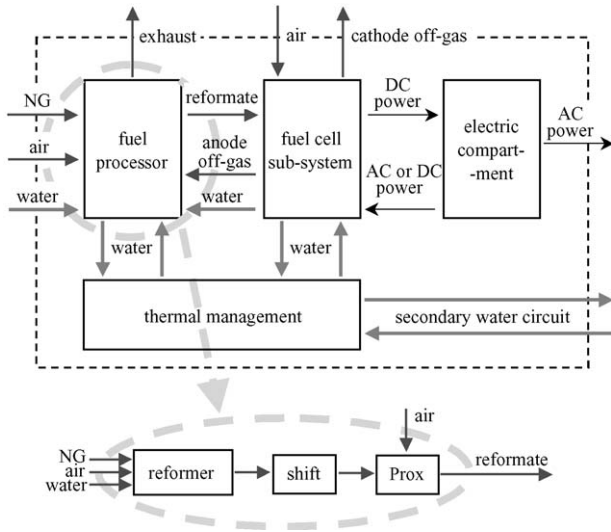


Fig. 1. Functional diagram of a small NG-fed cogeneration PEMFC system.

According to Refs. [1,2], SRR is often preferred to ATR for stationary applications because of a higher efficiency. It is effectively the case of main Japanese manufacturers (Osaka Gas, Tokyo Gas, IHI) who chose this technology. Nevertheless, Plug Power's GenSys™, which has been widely sold in the USA and among the world (with adaptation), produces hydrogen from natural gas using ATR.

The two technologies are mature and have rather close efficiencies. Many definitions of fuel processor efficiency are found in literature, this can be misleading. The definition chosen here is the one of Refs. [2,3], i.e. the ratio of lower heating value (LHV) of H<sub>2</sub> consumed in the stack to the LHV of total inlet NG. With a SRR-based fuel processor, the efficiency varies from 60% [4] to 78% [3] at full load.

The main difference between SRR and ATR is the concentration of hydrogen in dry reformat out of the Prox, 70–80% with SRR [5] and 30–46% with ATR [5,6]. It leads to different maximum utilisation rates of the anode gas. Because of a sharp increase of the anodic contribution of cell activation overvoltage when there is a lack of hydrogen near the membrane, a stack fed with reformat cannot use H<sub>2</sub> at 100%. The more the reformat rich in hydrogen is, the higher the utilisation rate can be. As it can reach 100% with pure hydrogen, 80–85% with reformat from SRR, it seems that it is not higher than 70% with a reformat from ATR [7].

This paper deals with steam reforming only.

A SRR is made of a (usually) tubular catalyst bed, where a mix of steam and NG ("feed mix") produces mainly hydrogen, carbon monoxide and dioxide and a burner where combustion of NG ("NG fuel") and anode off-gas brings heat to support the endothermic reaction in the bed. Exhaust gas is cooled warming-up the feed mix to temperature of reaction (600–900 °C) and boiling water.

The fuel cell sub-system is composed of a stack and its auxiliary equipment. The stack is typically made of 20–150 cells with an active area of 100–1000 cm<sup>2</sup> per cell. It delivers 2–10 kW DC. It is operated at low pressure (less than 200 mbar g) and at a tem-

perature between 50 and 75 °C [8]. The main auxiliary in terms of power consumption is the air compressor. Then, a device to humidify and preheat this inlet air is necessary. Humidification of anode inlet gas may not be necessary because the reformat out of the Prox has a relative humidity between 60 and 100%.

At full load, the whole system has an electrical gross efficiency of 27% [4] to about 35%. Gross efficiency is the ratio of DC power produced by the stack to the LHV of total inlet NG. It is the product of the fuel processor and the PEMFC stack efficiencies. The energetic efficiency of the stack is defined as the ratio of the electric power produced to the LHV of consumed H<sub>2</sub>. It depends on the conception (membrane electrode assembly, design of flow field channels, etc.) and operating conditions (temperature, pressure, humidity and utilisation rate).

Thermal efficiency of the system is defined as the ratio of the heat captured in the secondary water circuit to the LHV of the inlet NG, and has a value between 30 and 60%. A global efficiency can be defined as the sum of electrical and thermal efficiencies.

The fuel processor and the fuel cell sub-systems interact strongly, not only in the direction from fuel processor to fuel cell, but also in the other way, like anode off-gas (sent to the SRR burner), water produced by the cells (sent to the reaction chamber of SRR) and cooling circuit which crosses the two sub-systems (Fig. 1).

The fuel processor consumes water, while the fuel cell produces some. Water balance is the difference between collected water by condensation and needs of water for the fuel processing. It can be positive if enough liquid water is recovered, or negative if not. In the last case, additional water has to be brought to the system.

Once the design of the process is set, key parameters which characterize operating conditions are defined: "NG fuel" to total NG ratio (NG fuel/NG), steam to carbon ratio (S/C), air factor at the burner ( $\lambda$ ), oxygen to carbon monoxide ratio at the inlet of Prox (O<sub>2</sub>/CO), hydrogen utilisation rate in the anode ( $\tau_{H_2}$ ) and oxygen utilisation rate in cathode ( $\tau_{O_2}$ ).

$$\frac{\text{NG fuel}}{\text{NG}} = \frac{F_{\text{NG fuel}}}{F_{\text{NG total}}} \quad (1)$$

$$\frac{S}{C} = \frac{F_{\text{H}_2\text{O}}^{\text{SRR in}}}{\sum_i i F_{\text{C}_i\text{H}_{2i+i}}^{\text{feed}}} \quad (2)$$

$$\lambda = \frac{F_{\text{O}_2}^{\text{burner}}}{F_{\text{O}_2}^{\text{stoichio burner}}} \quad (3)$$

$$\frac{\text{O}_2}{\text{CO}} = \frac{F_{\text{O}_2}^{\text{PROX in}}}{F_{\text{CO}}^{\text{PROX in}}} \quad (4)$$

$$\tau_{\text{H}_2} = \frac{IN_{\text{cell}}}{[2FF_{\text{H}_2}^{\text{anode in}}]} \quad (5)$$

$$\tau_{\text{O}_2} = \frac{IN_{\text{cell}}}{[4FF_{\text{O}_2}^{\text{cathode in}}]} \quad (6)$$

where  $F_{\text{H}_2\text{O}}^{\text{SRR in}}$  is the molar flow rate of  $\text{H}_2\text{O}$  entering SRR,  $\sum iF_{\text{C}_i\text{H}_{2i+1}}^{\text{feed}}$  the equivalent molar flow rate of carbon in NG feed,  $F_{\text{O}_2}^{\text{stoichio burner}}$  the molar flow rate of oxygen in air entering the burner to conduct the combustion at stoichiometric conditions,  $N_{\text{cell}}$  the number of cells of the stack,  $I$  the stack current and  $F$  is the Faraday constant.

Furthermore, the temperature in the burner (adiabatic combustion temperature), the SRR (temperature of reaction  $T^{\text{RC}}$ ), the shift, the Prox and the stack have to be known. Because the pressure is close to the atmospheric one, it has very little influence on stack performance. Yet, it is an important parameter concerning thermochemical equilibrium in SRR reaction chamber and liquid/vapour equilibriums.

## 1.2. Modelling

In order to evaluate what the contributions of different parts of the system are and how they are linked, a steady state model of the whole system (minus the electric compartment) has been made. To do this, Thermoptim<sup>®</sup>, a software developed within CEP to calculate complex thermodynamic cycles, has been used. Elementary components (burner, compressors and heat exchangers) are modelled within Java classes, implemented within the software. Thermoptim<sup>®</sup> links these different elements and processes fluids, thanks to its library of physical and thermodynamic properties of gas and liquids [9]. External classes (i.e. added by the user) have been developed to model the stack, the three reactors of the fuel processor (SRR, shift and Prox) and condensers/separators of a humid gas.

### 1.2.1. Model of steam reforming reactor

This model deals with the whole reactor, composed of a reaction chamber, a burner, a preheater and a boiler (Fig. 2). The reaction chamber of the SRR is modelled taking into account the composition of NG, steam to carbon ratio and physical conditions (pressure and temperature). The following reactions are considered:

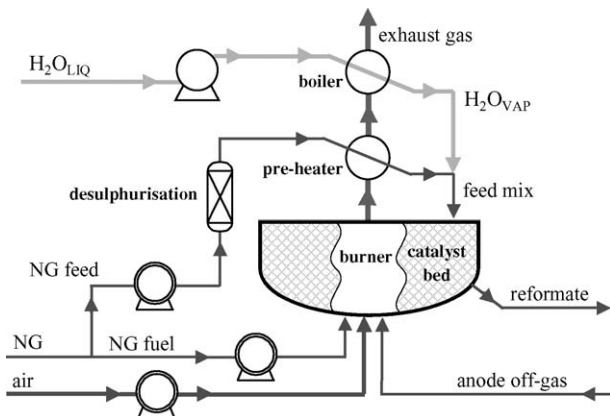
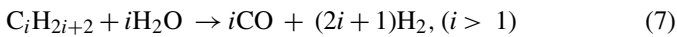


Fig. 2. Steam reforming reactor (SRR) and its direct environment.



Reaction (7) is supposed to be total. Degrees of conversion of (8) and (9) are determined by the equilibrium law (equilibrium constant and partial pressures) for the requested physical conditions, i.e. temperature of reaction ( $T^{\text{RC}}$ ) and pressure. A correction factor,  $\chi$ , can be added to take into account the fact that equilibrium of (8) is not completely reached, due to kinetic limitations. For example, if  $\chi = 90\%$ , the degree of conversion is 0.9 time the value of the one calculated with equilibrium law. Reaction (9) is much faster than (8), so it is assumed to be at equilibrium [1].

The composition of the products at thermochemical equilibrium is checked with a powerful computation engine, T&TWinner [20], which calculates all components present in fixed temperature and pressure conditions with minimization of the free enthalpy method. It is verified that no solid carbon (graphite) nor other components (like methanol) are formed in these conditions and that superior alkanes react effectively totally.

The burner is supposed to be composed by an adiabatic combustion chamber (integrated module of Thermoptim<sup>®</sup>) followed by a heat exchanger with the chamber of reaction.

### 1.2.2. Model of shift reactor

The shift reactor (or the two shift reactors) is modelled considering reaction (9) in an adiabatic catalyst bed, with a known inlet temperature. Thermochemical equilibrium, which depends on temperature (calculated by dichotomy) and composition, is supposed to be reached. Total pressure has no influence on the equilibrium.

### 1.2.3. Model of Prox

The Prox reactor is modelled very simply. Air is added to reformat so that two reactions take place in this reactor:



$\text{O}_2$  has to be in excess compared to CO, i.e.  $\text{O}_2/\text{CO}$  is superior to 0.5. Relation (10) is supposed to be total at 99% and remaining  $\text{O}_2$  reacts totally with hydrogen following reaction (11).

### 1.2.4. Model of stack

The stack model is macroscopic and rather basic. It is based on mass and energy balances, assuming that the faradic efficiency is 100%. Temperature is supposed to be uniform. Pressure is supposed to be close to atmospheric pressure, which means that this model does not take into account the influence of total pressure on fuel cell performance. Heat is entirely evacuated in water-cooling circuit.

Inlets are molar flow rates and compositions of inlet anode and cathode gas, current production ( $I$ ), number of cells ( $N_{\text{cell}}$ ), active area ( $A$ ), temperature of stack ( $T^{\text{stack}}$ ), membrane water transfer coefficient ( $\alpha$ , defined in relation (3) of Ref. [10]) and coefficients of the electrical equation.

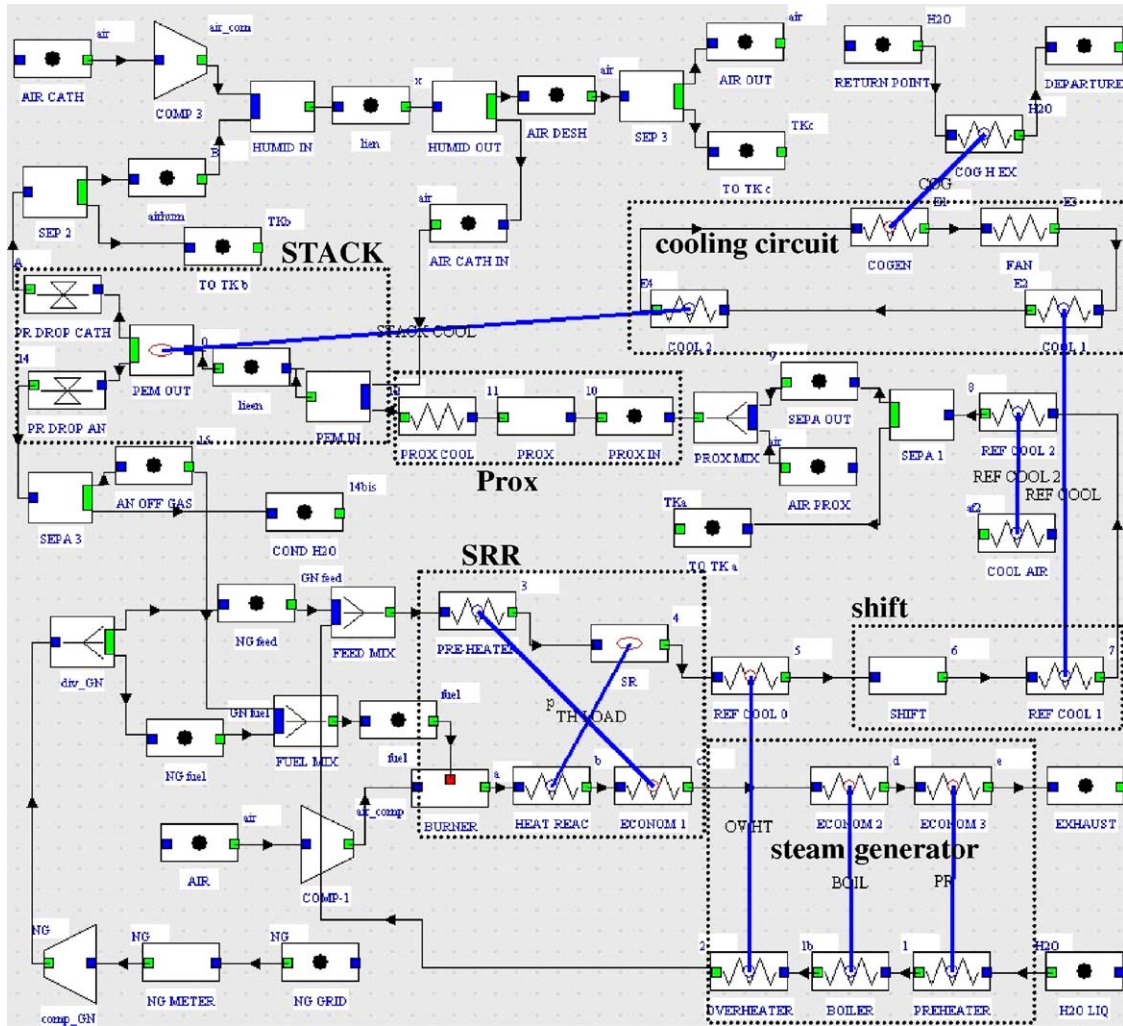


Fig. 3. Thermoptim® diagram flowsheet.

Outlets are stack voltage and power, heat to reject and flow rates and compositions of outlet fluids, which can be biphasic.

$$F_{H_2}^{an\ out} = F_{H_2}^{an\ in}(1 - \tau_{H_2}) \quad (12)$$

$$F_{O_2}^{cat\ out} = F_{O_2}^{cat\ in}(1 - \tau_{O_2}) \quad (13)$$

$$F_{H_2O}^{cat\ out} = F_{H_2O}^{cat\ in} + \tau_{H_2}(1 - \alpha)F_{H_2}^{an\ in} \quad (14)$$

$$F_{H_2O}^{an\ out} = F_{H_2O}^{an\ in} + \tau_{H_2}\alpha F_{H_2}^{an\ in} \quad (15)$$

No air bleed is considered at the entrance of anode compartment.

Electrical equation is taken into account very simply because the purpose of this model is not to study the stack for itself and because the fuel cell is operated in the linear part of its polarisation curve. Thus, the power delivered by the stack can be written as below.

$$P = IV_{stack} = I(V^* - kI) \quad (16)$$

where  $V^*$  and  $k$  are mathematic coefficients, homogenous to a tension and a resistance (but are not Nernst potential and resistance of stack). They are fitted with experimental data and they

depend on pressure, temperature and utilisation rates. In this study, temperature and pressure vary very few and the utilisation rates only depend on intensity.

Other electrical equations can be found [11–13] and are used to study the influence of different parameters on the stack electrical performance.

### 1.2.5. Integration in the whole system

All components are linked within diagram interface, as shown in Fig. 3. In the following of this article, this architecture is considered as the reference system.

### 1.3. Experimental data and fittings of parameters

Units from H Power,<sup>2</sup> installed and operated by Gas de France have been widely metered in order to assess precisely their operating conditions. Results given hereafter are from the unit of a town hall in the district of Dunkerque, a day of November 2003. It was fed with NG from Groningue, which

<sup>2</sup> In November 2002, Plug Power Inc. acquired H Power Corp.



Table 1  
Experimental data

Current demand (A)	30	40	60	80
$V_{\text{stack}}$ (V)	86.5	84.0	78.2	71.5
Electric DC power of stack (kW)	2.6	3.4	4.7	5.7
Heat from stack in cooling circuit	2.4	3.3	5.0	7.0
Heat in secondary water circuit (kW)	4.0	5.0	6.0	8.0
NG flow rate (NL min <sup>-1</sup> )	22.2	26.2	31.5	39.3
NG feed flow rate (NL min <sup>-1</sup> )	14.3	16.8	20.0	26.0
Gross electrical efficiency (%)	21.6	23.5	27.5	26.7
Thermal efficiency (%)	33	35	35	37
Global (gross) efficiency (%)	55	59	63	64
Steam to carbon ratio, S/C	9.9	9.0	8.3	7.3
Abs. pressure gas inlet anode (bar)	1.08	1.09	1.09	1.12
Temperature of burner (°C)	850	875	905	950
Temperature of reaction chamber, $T^{\text{RC}}$ (°C)	650	650	650	650

has a rather small LHV (9.1 kWh(Nm<sup>-3</sup>)) because of a relatively important concentration of N<sub>2</sub> (about 10%). That day, the unit ran on its four main levels of gross DC power (from 2.6 to 5.8 kW) which are controlled by the current demand of 30–80 A.

Main performance and operational conditions are given in Table 1. Most of these figures are directly measured, the others (efficiencies, heat from stack in cooling circuit and S/C) are obtained by simple calculations from measured ones.

As said before, modelling is not used to study the stack itself but the whole system. Electric power is not an outlet but an inlet and stack electrical efficiency is set to fit this measured power. Flow rates, S/C ratio, temperatures and pressures in Table 1 are also inlets of the model.

The utilisation rate of hydrogen is calculated from (5), with  $N_{\text{cell}} = 120$  and a molar flow rate of H<sub>2</sub> coming from the Prox.

Gas analysis has been conducted in several points of the system by the LSGC.<sup>3</sup> It permitted to fit  $\chi$ , the correction factor defined in Section 1.2.1, so that calculated composition of reformate out of SRR sticks to measured one. This gas analysis also confirmed the fact that thermodynamic equilibrium is reached in the shift reactor.

It can be noticed that heat released in water-cooling circuit calculated with model of stack (Table 2b) is close to the one measured (Table 1).

Water balance of the whole system shows a deficit which is rather important. City water is needed and a de-ioniser is used. This means that an additional device consumes power from the stack and water. Thus, a way of improvement is to get a positive water balance. Water condensed from anode off-gas is not collected, nevertheless, simply collecting this water is not sufficient to achieve a positive water balance.

At part load ( $I = 30$  A), the value of  $-0.42$  mL s<sup>-1</sup> was measured. The relative error of the calculated value is 15%. The membrane water transfer coefficient may have been underestimated. With a value of 50%, this error drops to 6%. For redesigning, water from both anode off-gas and cathode off-gas will be

Table 2  
Assumed parameters of model (a) and calculated performance (b)

	30	40	60	80	<i>a</i>
Current demand (A)	30	40	60	80	
$\chi$ in SRR (%)	100	100	98	98	
$\lambda$ (burner)	3.0	2.7	2.7	2.4	
Stack efficiency (%)	57.5	55.8	52.0	47.5	
Membrane water transfer coefficient (%)	40	40	40	40	
<b>b</b>					
Conversion of CH <sub>4</sub> (%)	99.5	99	97	95.5	
Fuel processing efficiency (%)	37.5	42.2	52.7	56.2	
Utilisation rates of hydrogen (%)	52.0	57.9	73.8	75.5	
Heat from stack in cooling circuit	2.5	3.4	5.5	7.8	
Water balance (mL s <sup>-1</sup> )	-0.35	-0.41	-0.44	-0.57	

collected so that the influence of assumed value of membrane water transfer coefficient will be minor.

When power increases, the fuel processor efficiency increases, the stack efficiency decreases and the product of these two efficiencies, the gross electrical efficiency, is maximum at 75% of full load. This is mainly due to the fact that the fuel processor is designed to be run at full power. For example, steam to carbon ratio (S/C), which is already pretty high at full power, is even higher at part load. This leads to very good methane conversion rates but to bad fuel processor efficiencies.

#### 1.4. Improvements and redesigning

In comparison with the units of the project, improvements in terms of efficiency can be done. This part focuses on gross electrical efficiency, thermal efficiency and water balance. Electric power is the product that makes the difference compared to water heating appliances. Heat is a by-product and it is much cheaper than electric power (even if in some systems, the monitoring is done following heat demand). Furthermore, electricity can easily be sold to a local grid, heat cannot. For these reasons, optimisation has been conducted considering in this order, electrical gross efficiency, water balance and thermal efficiency. In fact, water balance and thermal efficiency are linked because the temperature of the cold point of the system, which is often the return point of secondary water circuit, has the same effects on these two aspects: the more this temperature is low, the more heat can be recovered and the more water can be condensed.

Many publications deal with thermodynamic analysis of steam reforming applied to a fuel cell [6,14], some others study the issue of kinetics [15,16]. Fewer analyse the system as a whole, with interactions between the three sub-systems: fuel processing, fuel cell and water-cooling circuit. This type of approach was adopted by [7], but concerning a bigger system (200 kW). In that publication, separated data from an ATR-based fuel processor manufacturer and from a PEMFC supplier were used. Furthermore, the objectives of this author were to maximise heat and liquid water recovery (and to simplify control and design) and did not tackle the issue of maximisation of gross electrical efficiency.

The improvement of gross electrical efficiency can be achieved by increasing fuel processing efficiency without

<sup>3</sup> Laboratoire des Sciences du Génie Chimique de Nancy (Laboratory of Chemical Engineering Sciences).

decreasing stack efficiency. To do so, design aspects and operating conditions (defined in (1)–(6)) have to be considered.

For units of the project, “NG fuel” flow rate is regulated with an ON/OFF solenoid valve with the temperature of reaction chamber and the airflow rate constant. Air factor is calculated for the case when anode off-gas and “NG fuel” are sent to burner. When “NG fuel” is off, the air factor is huge and the temperature of the combustion chamber drops rapidly. A continuous regulation enables a better fit of the air factor, thus, a smaller “NG fuel” flow rate and better controlled emissions of pollutants (CO, NO<sub>x</sub>).

Using fewer heat power is possible concerning heating of reaction chamber but it would not be sufficient to vaporize process water. Thus, the steam to carbon ratio has to be reduced between 2 and 4, bearing in mind the potentiality of solid carbon formation in catalyst bed.

Furthermore, in the Prox, oxygen to carbon monoxide ratio is pretty important (10–30). This causes a drop of H<sub>2</sub> concentration in the reformat of 5–10%. The optimum value, 0.5, is a theoretical limit, thus, not reachable. Nevertheless, Ref. [3] claims O<sub>2</sub>/CO = 1.5, this value is chosen for redesigning.

As seen in Table 2b, the system studied in the EPACOP project has a very good conversion rate of methane into hydrogen. More generally, in literature, authors always try to reach a “good quality” of reformat concerning methane (1–3% of CH<sub>4</sub> in dry reformat entering the anode). Nevertheless, the methane crosses the anode compartment without interacting with membranes and it is sent to the burner of the SRR. A methane-rich reformat is not “bad” for the system. Furthermore, it entails less heat needed from burner and a smaller flow rate of “NG fuel”. In fact, the decreasing methane conversion rate in SRR (by decreasing S/C) allows to decrease the “NG fuel” to total NG ratio from about 30% to zero, thanks to enough CH<sub>4</sub> and H<sub>2</sub> in anode off-gas.

The model described in Sections 1.2 and 1.3 is run with a different architecture of the reference case concerning the heat management and with other operating conditions. A return temperature of secondary circuit of 30 °C is supposed. This weak temperature allows to condense more water and to obtain a positive water balance. The approach temperature in condensers (minimum temperature difference between the outlet gas and the inlet cold fluid) is supposed to be 11 °C as in Ref. [17]. If this cold point is not available (for example, 45 °C for return point), condensing more water is possible by using fans and by increasing slightly air pressure at cathode.

Concerning thermal management, it is simplified in using the secondary circuit (user’s circuit) directly as a cooling circuit, as it is the case in domestic boilers. The flow rate of this circuit has to be regulated by a water pump so that it is different for different levels of electric load (Table 3).

Reactors and stack have the same design as the reference system but are used slightly differently. Concerning SRR,  $\chi = 80\%$  is a likely value, calculated from reformat composition with a SRR used with S/C = 3,  $T^{\text{RC}} = 650\text{ °C}$  at atmospheric pressure [18]. Concerning the stack,  $\tau_{\text{H}_2} = 85\%$  at full load is an assumption from Ref. [19]. The value of 70% is based on the fact that utilisation rate of hydrogen is lower when the stack runs at part

Table 3  
Parameters assumed and calculated of redesigned system

	Full load	Part load
Electric power of stack (kW)	5.2	2.6
Total NG flow rate (NL min <sup>-1</sup> )	26.0	12.4
“NG fuel” flow rate (NL min <sup>-1</sup> )	0	0
S/C	3	4
$\lambda$	1.2	1.5
Temperature of burner (°C)	1430	1320
$T^{\text{RC}}$ (°C)	685	685
Correction factor in SRR ( $\chi$ ) (%)	80	85
Temperature of shift reactor (°C)	200	200
O <sub>2</sub> /CO	1.5	1.5
$\tau_{\text{H}_2}$ anode (%)	85	70
$\tau_{\text{O}_2}$ cathode (%)	50	50
$T^{\text{stack}}$ (°C)	61	60
Stack efficiency (%)	47.5	55.0
Flow water-cooling circuit (mL s <sup>-1</sup> )	85	38
Performances of system		
Gross electrical efficiency (%)	36.4	38.2
Fuel reforming efficiency (%)	76.6	69.5
Heat recovery efficiency (%)	72	69
Global efficiency (%)	108	107
Water balance (mL s <sup>-1</sup> )	+0.01	-0.04

load. Efficiencies of the stack are the same as the system of reference.

Global gross efficiency is very close to the maximum thermodynamic value because the cooling circuit is optimised and heat losses are supposed to be zero, thus, all the heat is recovered. As in condensation boilers, an efficiency of more than 100% is possible because it is calculated with the LHV; thus, the maximum is 111%, for a system with inlets and outlets at 25 °C.

Unfortunately, the water balance is negative at part load. This is due to a lower utilisation rate of H<sub>2</sub>, which induces anode off-gas and cathode off-gas with smaller concentration of water than at full load.

## 2. Conclusion

Existing cogeneration PEMFC systems are largely perfectible, not only with a thermodynamic analysis but also with a realistic model based on reliable experimental data and likely assumptions. This makes it possible to suggest new operation strategies and new system designs. This study does not take into account improvements that can be made in terms of catalysts and membranes.

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