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Novel Concept for Evaporative Cooling of Fuel Cells: an Experimental Study Based on Neutron Imaging

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Abstract

Polymer Electrolyte Fuel Cells (PEFC), although a promising technology for carbon free production of electricity, are penalized by system complexity, partly due to cooling and humidifying systems. These systems are necessary to avoid the heating up and drying of the membrane, which stop the electrochemical reaction. Here we present an evaporative cooling concept for PEFC developed at Paul Scherrer Institute (PSI). Unlike other concepts, our approach does not require any additional layer in the cell structure. Water flows through dedicated anode flowfields channels, parallel to the gas channels, and is distributed over the cell area thanks to a modified gas diffusion layer (GDL). A synthesis method developed at PSI transforms some portions of the GDL into hydrophilic patterns, which wick the water from the supply channels at low capillary pressure. These hydrophilic areas, parallel and equally spaced, define pathways for liquid water separated from the gases, which avoids flooding. A test cell was built to investigate both water transport with the help of neutron radiography and heat transport thanks to integrated heat flux sensors. Here, we will present how the evaporation can be controlled by the mass flow rates, temperatures, pressures of gases and the geometry of the hydrophilic lines

Keywords: Evaporative Cooling, Humidification, Mass and Heat Transfer, Neutron Imaging, Polymer Electrolyte Fuel Cell (PEFC)

1 Introduction

Polymer electrolyte fuel cells (PEFC) are a promising technology for automotive applications, as they produce carbon-free electricity. However they are still penalized by their cost and their complexity, partly due to the cooling and humidifying systems. These systems are needed because heat released by the electrochemical reaction, which is of the same order of magnitude as the electrical power generated (Zhang and Kandlikar [1]), can heat up and dry the membrane, effectively stopping the reaction. As a result, cooling of fuel cells is a widely studied topic (Zhang and Kandlikar [1], Kandlikar and Lu [2], Hoogers [3]). While most systems rely on convective cooling thanks to a liquid or a gas stream between two adjacent flowfields (Owejan et al. [4], Chen et al. [5], Chen et al. [6], Schmidt et al. [7]) this type of scheme strongly increases the volume of the PEFC stack, since it requires additional space for the coolant. Furthermore, it requires a humidification system on top of the cooling system (Barbir [8]). As a result, evaporative cooling has received a lot of attention (Meyers et al. [9], Goebel [10], Choi et al. [11], Fly and Thring [12]): by evaporating water within the fuel cell, it simultaneously provides cooling and humidification, which removes the need for costly external humidifiers. Additionally, thanks to the high latent heat of vaporization of water, a smaller mass flow rate of liquid water than with a convection-cooled system is needed (Meyers et al. [9]). Last, a numerical study on the stack level showed that evaporative cooling systems could have the innate ability of regulating themselves without active control (Fly and Thring [12]). Several authors have come up with different solutions. Several authors have used direct water injection by nebulizing water into the gas streams (Choi et al. [11], Adcock et al. [13]). This method is extremely simple but is penalized by the high energy consumption for nebulization (Zhang and Kandlikar [1]) and the lack of control on where the evaporation takes place. This last problem can be solved by another approach, the porous water transport plates (WTP) developed by UTC Power capable of removing excess liquid water from the gas channels and bringing it where the gas stream is not saturated to evaporate (Meyers et al. [9],

Weber and Darling [14]). This scheme, however, relies on special expensive WTP and requires an additional space to carry the liquid water between two adjacent cells. In this article, we present an evaporative cooling concept that aims at eliminating the need for a dedicated space for cooling water and can be engineered without strong changes to the fuel cell design (see Figure 1). We then focus on the evaporation power that can be achieved with our concept and propose a simple 1D model to predict it, that we validate with our experimental results.

2 Scientific Approach

The evaporative cooling concept for PEFC developed at Paul Scherrer Institute (PSI) does not require any additional layer in the cell structure. Water flows through dedicated channels of the anode flowfield, parallel to the gas channels, and is distributed over the cell area with the help of a modified gas diffusion layer (GDL), so that no additional space is necessary for water transport, and standard bipolar plates can be used. The modified GDL is central to our concept: following a procedure described in Boillat et al. [15], Forner-Cuenca et al. [16] and [17], the wettability of this GDL can be selectively tuned: some parts are made hydrophilic, while others remain hydrophobic. For the evaporative cooling concept, the hydrophilic parts are parallel and equally-spaced lines that can wick the water from the supply channels at very low capillary pressure (near 0 mbars), while the hydrophobic lines remain free of water for capillary pressures as high as 40 mbars (Forner-Cuenca et al. [18]). The modified GDL is placed into the fuel cell so that the hydrophilic lines are orthogonal to the flowfield's channels. The water from the supply channel imbibes the lines, which brings it in contact with the flowing gas. It evaporates there thanks to the convection of the gas flow, providing cooling and humidification. Meanwhile the hydrophobic lines provide free pathways for the gases, thus preventing flooding of the cell (Figure 1). The aim of this article is to characterize the cooling power of our concept and determine how the evaporation can be controlled by the

mass flow rates, temperatures of gases, and the modified GDL itself. In order to achieve this aim, the test cell had to fulfill several requirements. First, in order to investigate the heat transport in the cell, we had to be able to measure heat fluxes precisely, and to control the boundary conditions of the flowfields. Second, because we are evaporating water within the fuel cell, a mass balance of the cell would not bring enough relevant information. Neutron Radiography (NR) is the only method which allows us to visualize the water distribution in the cell through all the different dense materials that we are using. In the next paragraph, we discuss how these requirements are met.

3 Experimental

3.1 Thermal Test Cell and Instrumentation

Heat flux measurements, although not often performed, can measure the thermal signature of the cell's reaction (Burheim et al. [19]) and can provide information on heat transfer within a fuel cell (Thomas et al. [20]). In our case, they enable us to determine the heat consumed by the evaporation and how it evolves with various parameters. Investigating the heat transport in the cell requires precise isothermal boundary conditions on the flowfields. This is achieved by monitoring the flowfields temperatures with resistive thermometers located on the current collectors (CC). These temperatures are used as a feedback for PID-regulated Peltier thermoelectric coolers (TEC), which ensure precise and fast regulation of the flowfields' temperatures. Heat flux measurements are made with Captec® tangential-gradient heat flux sensors (HFS). These sensors are placed between the TEC and the CC, and deliver a voltage proportional to the net incident heat flux, with a sensitivity of around $2 \mu\text{V W}^{-1} \text{ m}^2$. Because these sensors are very thin (around 0.4 mm) and thermally conductive, their thermal resistance is very limited, which reduces the impact of the sensor on the measured heat flux. Figure 2 shows how all these elements are assembled together between an aluminum compression body and the cell itself.

3.2 Post-processing of the Heat Flux Measurement

Figure 3 shows how the test cell is divided into three segments. While only the middle segment corresponds to the active area, the two additional segments are also equipped with HFS and Peltier TECs. The purpose of segment 1, upstream of the active area, is to define precise boundary conditions for the inlet gas and water. Since neither the gas nor the water are precisely at the flow field temperatures before reaching the test cell, this segment ensures that equilibrium between the fluids and the endplates is reached before entering the active area. It also ensures that the gas boundary layer is established, and the heat transfer coefficient (*HTC*) and mass transfer coefficient (*MTC*) will be near constant. Segment 3 helps control that there is no significant drop in the gas temperature due to the evaporation in the cell area. To account for the heat losses during operation, the heat fluxes of all six segments are recorded while the cell temperature is 80 °C, with neither gas nor liquid flow. The results are noted $\phi_{losses\ An,Ca}$ and are subtracted from all measurements:

$$\phi_{An,Ca} = \phi_{HFS\ An_2,Ca_2} - \phi_{losses\ An,Ca} \quad (1)$$

Note that for the rest of this article, all heat fluxes will be corrected so as to represent the heat flux over the surface of the active area of the cell S_{AA} , and not over the surface of the heat flux sensors S_{HFS} . This was done in order to compare to the heat flux generated by the electrochemical reaction:

$$\phi = \phi \cdot \frac{S_{HFS}}{S_{AA}} \quad (2)$$

3.3 Fuel Cell Description

Although we do not present *in operando* results in this paper, the cell was mounted so as to be fully operable. The various elements are:

- (i) The flowfields, 1 mm-thick stainless steel plates with a 5 µm-thick gold coating. The anode flowfield has one channel dedicated to the transport of liquid water, surrounded

by five gas channels on each side. The cathode flowfield has eleven gas channels. All channels have a width of 1 mm and a depth of 0.5 mm. The ribs are 1 mm wide.

- (ii) Anode GDL: the patterned GDL is placed only on the anode. Originally a Toray TGP-H-060 carbon paper with a thickness of 250 μm , it is first coated with Fluorinated Ethylene Propylene (FEP). It is then irradiated with an electron beam at LEONI Studer (Däniken, Switzerland). Thanks to an aluminum mask, some parts remain not activated. When the GDL is finally put into a solution of N-vinylformamide (NVF), the NVF polymerizes on the activated zones, thus creating the hydrophilic lines, whereas the non-irradiated parts keep their FEP content (70 % in weight) and remain hydrophobic. The hydrophilic lines are 500 μm large and the separation between two lines is 930 μm .
- (iii) Anode Micro-Porous Layer (MPL): a home-made MPL based on a Carbel® MP30Z (W.L. Gore & Associates) is used between the patterned GDL and the catalyst coated membrane. It is spray-coated with an ink preparation containing carbon black and FEP, to add a thin hydrophobic microporous layer and improve the interface with the GDL.
- (iv) The catalyst coated membrane (CCM) is a Primea® 5710 (W.L. Gore & Associates)
- (v) Cathode GDL and MPL: The GDL on the cathode side is a commercial SGL 24BC GDL with an integrated MPL.
- (vi) The active area of the cell is a 22 mm-wide and 20 mm-long rectangle
- (vii) PEN and PTFE gaskets ensure the tightness of the cell.

Since only the evaporation is studied here, the gases are H_2 on the anode and N_2 on the cathode. Flowfields' temperatures and inlet temperatures for the gases and for the cooling water are 80 °C. Pressures are equal to 2 bars absolute, and the gases inlet relative humidity is 0 %. Gases on the anode and on the cathode are in counter-flow. Gas and water on the anode flow in the same direction. It must be noted that the dimensions of the active area of our cell is very small in order to isolate the relevant effects. However, it is necessary to keep gas velocities representative of a longer cell, which means high stoichiometric ratios (more than

10 on the anode, 20 on the cathode), and high mass flow rates. High mass flow rates, together with the small dimensions of the active area, are favorable conditions for evaporation, since the mass flow rates are less likely to be fully saturated, but are unfavorable to humidification, since the water is less likely to condense and accumulate in the gas channels.

3.4 Neutron Radiography

Neutron Radiography (NR) was performed at the ICON beamline of the Swiss Spallation Source (SINQ) of the Paul Scherrer Institute (Villigen, Switzerland). More details on this beamline can be found in Kaestner et al. [21]. A neutron sensitive scintillator screen (6LiF/ZnS, thickness 50 μm) was used perpendicular to the beam axis. Images were recorded with a cooled CCD camera (Andor iKon-L, 2048 $\times$ 2048 pixels). The resulting pixel size was 50 μm and the effective resolution including blurring was approximately 100 μm . The neutron beam intensity is attenuated by any given material in accordance with the Beer-Lambert law:

$$I = I_0 \cdot e^{-\mu_{\text{material}} \cdot \delta_{\text{material}}} \quad (3)$$

where I_0 and I are the beam intensities respectively upstream and downstream a given material; μ_{material} the absorption coefficient of this material and δ_{material} its thickness. By dividing every image of a fuel cell pixel-wise by a reference image of the same fuel cell in totally dry conditions, the contribution of liquid water to neutron attenuation can be isolated. Using Eq. (3), it is then possible to determine the thickness of liquid water present in a fuel cell. It must be noted that a certain number of corrections must be made on all images in order to account for spatial inhomogeneity of the neutron beam, thermal noise of the camera, neutron scattering, and small displacements of the cell during the experiments. All corrections are described in Boillat [22]. An example of a post-processed image of our PEFC with evaporative cooling concept is shown on Figure 4. The red rectangle represents the actual active area. The dark line in the center is the water supply channel, and the thin lines

perpendicular to it are the water filled hydrophilic lines. The gas channels cannot be seen on the NR image, and have been represented as dotted lines.

4 Results and Discussion

4.1 Influence of the Gas Flow rates

Figure 5 shows the influence of the anode gas flow rate on the total measured heat flux, which is defined as:

$$\phi_{total} = \phi_{An} + \phi_{Ca} \quad (4)$$

For all cases, the behavior with respect to anode gas flow rate is similar: for small values of the anode normal volumetric flow rate, the heat fluxes increase linearly, before reaching a plateau at about 400 NmL min⁻¹. The linear increase corresponds to cases where the anode flow is small enough to be fully saturated with water vapor. On the other hand, for higher values of the flow rate, the flow leaves the active area without being fully saturated, and the heat fluxes do not reach their maximal value. The behavior with respect to the cathode flow rate, presented in Figure 6, shows a much quicker plateau. A cathode flow rate higher than 50 NmL min⁻¹ is large enough to ensure that the gas leaves the segment unsaturated. Last, it must be noted that the heat fluxes obtained for flow rates representative of a long cell behavior (i.e. higher than 400 NmL min⁻¹ on the anode, and 800 NmL min⁻¹ on the cathode) are higher than 1 W cm⁻² and in the same order of magnitude as the heat generated by the reaction during operation. The evaporative cooling concept can therefore produce enough cooling power for an operating cell.

4.2 1D model for the Evaporative Cooling

A simple 1D model was made based on several assumptions, summarized on Figure 7:

- (i) While the evaporation takes place at the hydrophilic lines, part of the water vapor is removed by the anode gas flow, and another part is removed by the cathode flow. For

simplification, we will refer to the part exiting on the anode side as the evaporation rate on the anode side $\dot{m}_{evap\ An}$ and the part exiting on the cathode side as the evaporation rate on the cathode side $\dot{m}_{evap\ Ca}$.

- (ii) The total heat flux is only due to evaporation in the cell:

$$\phi_{total} = \phi_{An} + \phi_{Ca} = (\dot{m}_{evap\ An} + \dot{m}_{evap\ Ca}) \cdot L_{LV} / S_{AA} \quad (5)$$

with L_{LV} the latent heat of vaporization of water. The evaporation on the anode side is driven by the convection from the anode gas flow.

- (iii) In the absence of water production by the chemical reaction, the evaporation on the cathode side is driven by water vapor diffusion through all the anode and cathode layers (GDLs, MPLs and membrane) and convection by the cathode gas flow. Since the diffusion strongly limits the mass transfer on this side, we refer to it as diffusion-driven evaporation in Figure 1 and in the rest of the text.

- (iv) The evaporation rate on the anode side is not influenced by the cathode flow

- (v) The evaporation rate on the cathode side is not influenced by the anode flow

The species conservation on a small element of an anode gas channel gives us:

$$U \cdot S \cdot (\rho_{H_2O}(x + dx) - \rho_{H_2O}(x)) = Surf \cdot MTC \cdot (\rho_{H_2O}(Surf) - \rho_{H_2O}(m)) \quad (6)$$

with U the gas velocity, ρ_{H_2O} the water vapour density, MTC the mass transfer coefficient and S the cross-section of the gas channel. On the right side of Eq. (6), $Surf$ is the contact surface between the anode gas flow and the liquid water in the hydrophilic lines, whereas m corresponds to the middle of the channel.

Integrating Eq. (6) over the active area gives the following equation for the molar fraction of water vapour Y_{H_2O} :

$$Y_{H_2O} = Y_{sat} \cdot \left[1 - \exp\left(-\frac{\overline{MTC} \cdot \rho_{H_2O} \cdot Surf}{\dot{m}_{gas\ An}}\right) \right] \quad (7)$$

where $\dot{m}_{gas\ An}$ is the anode gas mass flow rate, Y_{sat} the molar fraction of water vapour at saturation, ρ_{H_2} the density of H_2 . $\overline{MTC}$ is the average mass transfer coefficient and is computed as (Incorpera et al. [23]) :

$$\overline{MTC} = \frac{Sh \cdot \mathcal{D}}{D_H} \quad (8)$$

with Sh the Sherwood number, $\mathcal{D}$ the diffusivity of water vapour into H_2 and D_H the hydraulic diameter. The Sherwood number can be computed from the Nusselt number thanks to the heat and mass transfer analogy:

$$Sh = Nu_T \cdot Le_{H_2O}^{1/3} \quad (9)$$

For an established laminar internal flow in a rectangular duct with only one wall cooling (evaporating), the value for the Nusselt number (from Shah [24]) is $\overline{Nu}_T = 3.185$.

N.B: For low values of $\dot{m}_{gas\ An}$, the molar fraction of water vapour $Y_{H_2O}(Surf)$ tends towards Y_{sat} the molar fraction of water vapour at saturation. This value depends only on the temperature and the pressure of the gas flow. As a result, these parameters have a high influence on the evaporation rate.

The evaporation rate over the integration domain is then equal to:

$$\dot{m}_{evap\ An} = \dot{m}_{gas\ An} \cdot \frac{M_{H_2}}{M_{H_2O}} \cdot \frac{Y_{H_2O}(Surf)}{1 - Y_{H_2O}(Surf)} \quad (10)$$

with M_{H_2} , M_{H_2O} the molar masses of hydrogen and water

And the heat flux generated by the evaporation rate on the anode is equal to:

$$\phi_{evap\ An} = \dot{m}_{evap\ An} \cdot L_{LV} / S_{AA} \quad (11)$$

Since the total heat flux measured is only due to the evaporation, we have:

$$\phi_{total} = \phi_{An} + \phi_{Ca} = \phi_{evap\ An} + \phi_{evap\ Ca} \quad (12)$$

However the heat flux measured on the anode ϕ_{An} (resp. cathode) is not equal to the heat flux due to the evaporation rate on the anode $\phi_{evap\ An}$ (resp. cathode).

4.3 Detection of the contact surface *Surf*

From Eq. (7), it is clear that the contact surface between the water in the hydrophilic lines and the anode gas flow, *Surf*, is an important parameter. An algorithm was developed to extract this parameter from the NR images. First, the gray-level images (Figure 8 a) are binarized with the Otsu threshold-detection algorithm (Otsu [25]) and the resulting image is inverted so that the pixels representing the water have a value of 1 (Figure 8 b). Then the image is filtered with several morphological closings and openings (Figure 8 c). A binary image is created to represent the flowfields ribs and channels (Figure 8 d): the ribs and the water channel have a value of 0, the gas channels a value of 1. This image is multiplied pixel-by-pixel to (Figure 8 c) to get a binary image where the pixels with value 1 are the pixels where the water is in contact to the gas flow (Figure 8 e). Last, the pixels of this final image are added and multiplied to the surface of one pixel to obtain the contact surface *Surf*.

4.4 Comparison to Experimental Results

Once *Surf* is known, the heat flux due to the evaporation on the anode $\phi_{evap\ An}$ can be computed and compared to the total measured heat flux ϕ_{total} for various anode and cathode gas flow rates (Please note that although the 1D model uses the mass flow rates, the figures show the evolution of the heat fluxes with respect to the normal volumetric flow rates for practical reasons). As can be seen on Figure 9, there is a good qualitative agreement between the 1D model and the measurements. In fact, the difference between measured and modelled heat fluxes is an offset that is independent on the anode mass flow rate. Based on this last fact and on Eq. (12), it seems reasonable to identify this offset with the heat flux due to the evaporation on the cathode $\phi_{evap\ Ca}$. Once the offset is removed for all cathode mass flow rates, the 1D model is in very good agreement with the measurements, as can be seen on Figure 10. The second step is to determine the heat flux due to the evaporation rate on the cathode side $\phi_{evap\ Ca}$. By applying Eq. (12), $\phi_{evap\ Ca}$ is obtained by subtracting the modelled

$\phi_{evap\ An}$ from the total measured heat flux ϕ_{total} , for all anode mass flow rates and all cathode flow rates. The results are shown on Figure 11. As can be seen, for each cathode mass flow rate, the computed points are very close to one another, which confirms our assumption that the evaporation rate on the cathode side is not influenced by the anode flow rate, and that the 1D model for the evaporation rate on the anode side is pertinent at least for the points tested so far.

A similar 1D model is made for the evaporation rate on the cathode side where:

$$\dot{m}_{evap\ Ca} = \dot{m}_{gas\ Ca} \cdot \frac{M_{N_2}}{M_{H_2O}} \cdot \frac{Y_{H_2O\ Ca}(Surf)}{1 - Y_{H_2O\ Ca}(Surf)} \quad (13)$$

with

$$Y_{H_2O\ Ca}(Surf) = Y_{sat} \cdot \left[1 - \exp\left(-\frac{(\overline{MTCa}^{-1} + C_1^{-1})^{-1} \cdot \rho_{N_2} \cdot Surf}{\dot{m}_{gas\ Ca}}\right) \right] \quad (14)$$

where C_1 is a fitting constant to take into account the diffusion of the water vapour in the various layers of the fuel cell. A least-square fitting method gives $C_1 = 0.071\text{ m s}^{-1}$. The fitted 1D model is also plotted on Figure 11.

4.5 Influence of the contact surface

A last experiment was made to determine the influence of the contact surface on the evaporation rate. Since changing the GDL itself would take too much time, the flow rate of cooling water is increased from 2 to 8g hr⁻¹. Such a small mass flow rate is achieved thanks to a Bronkhorst™ Liqui-Flow mass flow controller. Since these values are lower than the achievable evaporation rate, the cooling water cannot flow through the entire cell and oscillates around an equilibrium position. When the liquid mass flow rate increases, the distance reached by the water front and the number of imbibed lines increase and so does the contact surface between liquid and gas. The total heat flux is plotted with respect to the contact surface on Figure 12 and compared to the 1D fitted model. As can be seen, the trend agrees well with the 1D model. A conclusion that can be made from this last figure is that the

contact surface is an important parameter to influence the evaporation rate in the cell. Since we can choose the lines' geometry and separation, we should be able to have a good control of the cooling power within the cell.

5 Conclusions

A compact evaporative cooling for PEFC was developed using a patterned GDL with hydrophilic lines. A PEFC with this evaporative cooling concept has been investigated with conditions representative of a fuel cell but without chemical reaction to determine the effect of the gas mass flow rates and the geometry of the hydrophilic lines on the evaporative cooling power. Here we showed that:

- (i) with representative gas mass flow rates, the cooling power achieved is of the order of magnitude of the heat generated by a PEFC in typical operation (higher than 1 W cm^{-2})
- (ii) While evaporation takes place on the anode side, the water vapour is carried out of the cell by both the anode and the cathode flows.
- (iii) The cooling power achieved by the evaporation can be controlled with the mass flow rates of the gases, as well as the temperature and the pressure of the gases.
- (iv) The geometry and separation of the hydrophilic lines strongly influence the surface of water in contact with the gas flow and in turn control the evaporation rate and the cooling power.
- (v) Last, a simple 1D model has been established to predict the evaporation rate as a function of the temperature, the anode and cathode mass flow rates, and the contact surface between the water in the hydrophilic lines and the gas mass flow.

This work already enables us to make a few comments on the implications of our findings for the practical implementation of such a scheme on a working stack:

- Since the anode flow gets saturated much faster than the cathode flow, this scheme does not seem possible with a dead-ended anode, and a system to condense the water

vapour on the anode will have to be added. However this should not be a strong limitation.

- On the other hand, the cathode flow will have to remove most of the water vapour in the end. Our 1D model enables us to estimate for which conditions the cooling power will be enough for an entire commercial-size fuel cell. In particular, schemes with higher temperatures, higher mass flow rates and different materials (thinner, more diffusive layers) will be tested and compared to the model. A scheme with the evaporation rate on the cathode could also be tested.
- Although this work was done with isothermal boundary conditions, conditions in a fuel cell stack will be different. To be more representative, we are working to validate our scheme with adiabatic boundary conditions. We wish to test in particular the possible regulating capability of evaporative cooling concept expected by Fly and Thring (Fly and Thring [12]).

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List of Symbols

ϕ_{total}	Total measured heat flux / W cm^{-2}
$\phi_{losses An,Ca}$	Heat losses on the anode (resp. cathode) / W cm^{-2}

$\phi_{An,Ca}$	Heat flux measured on the anode (resp. cathode) and corrected for the losses / $W\ cm^{-2}$
$\phi_{evap\ An,Ca}$	Heat flux due to the evaporation on the anode (resp. cathode) / $W\ cm^{-2}$
$\dot{m}_{gas\ An,Ca}$	Mass flow rate of gas on the anode (resp. cathode) / $kg\ s^{-1}$
$\dot{V}_{gas\ An,Ca}$	Volumetric normal flow rate of gas on the anode (resp. cathode) / $NmL\ min^{-1}$
$\dot{m}_{evap\ An,Ca}$	Evaporation rate on the anode (resp. cathode) / $g\ hr^{-1}$ ($kg\ s^{-1}$ for the equations)
L_{LV}	Latent Heat of vaporization / $kJ\ kg^{-1}$
S	Surface of the flow cross-section / mm^2
$Surf$	Contact surface between the water in the hydrophilic lines and the anode gas flow / mm^2
S_{HFS}	Surface of the HFS / mm^2
S_{AA}	Surface of the Active Area / mm^2
ρ_{H_2O}	Water vapor density / $kg\ m^{-3}$
ρ_{H_2}	Hydrogen density / $kg\ m^{-3}$
U	Gas velocity / $m\ s^{-1}$
$Y_{H_2O\ An,Ca}$	Molar fraction of water vapor in the anode gas stream (resp. cathode)
$Y_{sat\ An,Ca}$	Molar fraction of water vapor in the anode gas stream at saturation (resp. cathode)
$M_{H_2,\ H_2O,\ N_2}$	Molar mass of hydrogen, water or nitrogen / $g\ mol^{-1}$
MTC	Mass Transfer Coefficient
HTC	Heat Transfer Coefficient
Nu_T	Nusselt number
Sh	Sherwood number
Le	Lewis number

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Figure Captions

Figure 1: Concept of evaporative cooling with patterned GDLs

Figure 2: Diagram of the thermal test cell

Figure 3: Thermal test cell and post-processing of the heat flux measurements

Figure 4: Post-processed NR image of the entire thermal test cell

Figure 5: Influence of anode flow rate on total measured heat fluxes

Figure 6: Influence of cathode flow rate on total measured heat fluxes

Figure 7: 1D model assumptions for the evaporation taking place in the test cell

Figure 8: Detection algorithm to determine the surface of the water filled hydrophilic lines in contact with the gas flow

Figure 9: Comparison of the total measured heat fluxes and the 1D model for a cathode flow rate of $1600 \text{ NmL min}^{-1}$

Figure 10: Comparison of the total measured heat fluxes and the 1D model for all cathode flow rates

Figure 11: Comparison of total heat flux due to the evaporation rate on the cathode side and the 1D model for all anode flow rates

Figure 12: Total measured heat flux with respect to the contact surface