



AutoRE

(Deliverable 1.6)

**Preliminary engineering solutions for
integration of membranes**





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Executive summary

This document is submitted in relation to deliverable D1.6, of the Horizon 2020/FCH JU AutoRE project. It describes the effects of the operating conditions on the performance of a membrane-based reformer process and the integration solutions for Pd-based membranes.

In this study, we performed a sensitivity analysis on the fuel processor system based on membrane steam reformer by varying operating parameters. The analysis has been performed in accordance to the deliverable D4.2 and published literature.

Brief overview of the membrane reformers types is given and the preliminary configurations are proposed built around the achievable membrane permeability and the required hydrogen flux.



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1 Introduction and scope of the document

Deliverable 1.6 is a part of Work Package 1 of the AutoRE project. The objective of this task is to define preliminary integration solutions for newly developed components, specifically hydrogen selective membranes.

The system in the study is a 50 kW_{el} cogeneration system (CHP) based on a polymer electrolyte membrane fuel cell (PEM FC). It includes a fuel processor unit where natural gas is converted to hydrogen rich gas by steam reforming reaction followed by water gas shift reaction. Reactions take place in a membrane-based steam reformer which compared to conventional multi stage hydrogen generation plant that includes water gas shift (WGS) reactors and a pressure swing adsorption unit (PSA) allows production of a high purity hydrogen gas in just one step and one reactor. In this way a plant is significantly simplified. Moreover, due to the direct extraction of hydrogen from a reaction zone and shift of the reaction equilibrium to the side of products, more hydrogen is generated.

In section 2 operating parameters assessment is performed including a comparison with the traditional plant. Effects of the operating parameters namely steam to carbon ratio (S/C), air to carbon ratio (A/C), temperature, and pressure, on the fuel processor efficiency are investigated. Hydrogen permeability and membrane surface sensitivity analysis are performed as well. The simulations have been performed for both the conventional and the membrane-based fuel processor plant. The results presented report the optimal operating parameters for the conventional and the membrane-based reformers from a thermo-chemical point of view.

In section 3 membrane reformer configurations are proposed based on the achievable membrane permeability values and the required membrane surface.



2 Working parameters assessment

2.1 Pd-based membranes for hydrogen generation

Hydrogen is most commonly produced by steam reforming of hydrocarbons. Such a technology includes two stages: production of syngas which is a mixture of CO and H₂, and following purification steps during which CO concentration in the syngas is decreased, while the H₂ concentration increases. To meet the requirements of a polymer electrolyte membrane fuel cell (PEM FC), produced hydrogen rich gas should not contain more than 1 ppm of CO. Such a process requires therefore several units: steam reformer, high and low temperature water gas shift reactor, and lastly, CO purification step through either preferential oxidation or pressure swing adsorption (PSA). In a CHP plant based on PEM FC including onsite hydrogen generation, cost of the fuel processing unit can reach 50% of the total plant cost [1]. Membrane reactors could lower the price of the fuel processing unit. Integrating membranes in a steam reformer, water gas shift reactor or as a separate unit (see D3.3 for more information), means that there would no longer be the need for a PSA unit which accounts for a large part of the capital costs, and plant simplification could be achieved. Furthermore, due to the continuous extraction of hydrogen from a reactor, chemical equilibrium is shifted to the product side, thus more hydrogen is produced. The only drawback of membrane reforming is that hydrogen is obtained under lower pressure and there might be a need for a compressor downstream of the fuel processing unit.

Dense palladium membranes have shown very good performance in hydrogen generation processes. Pd-Ag alloyed membranes can operate at high temperatures, i.e. 300 – 700 °C (stability depending on thickness), which makes them suitable for integration both in the steam reformer and water gas shift reactor. The main advantage of membrane integration with steam reformer is that hydrogen is removed directly from the reaction zone, and the subsequent purification steps are not needed [2]. Moreover, membrane reformer application allows for obtaining a higher steam reforming reaction efficiency at a lower temperature compared to the conventional reactors, thereby decreasing both production costs and the possibility of reforming catalyst deactivation by coke deposition [3]. Plant simplification can be obtained by integration of membranes in a water gas shift reactor as well. In this way, the PSA unit is not required anymore, neither is the cooling and dehydration steps, and the membranes are integrated into a milder environment regarding the reaction temperature, which is in the interest of their stability. Furthermore, the hydrogen partial pressure at the retentate side is higher, leading to a larger driving force for the permeation process. However, it has been found that the integration of membranes into WGSR reactor increases fuel processor efficiency only by 4.8 %, whereas in the case of integration into SR reactor the efficiency increases by 11.6%, compared to the conventional plant [2]. For this reason, it has been decided to investigate and show here the assessment of the membrane-based steam reformer.

Hydrogen permeation through the dense membranes occurs by a solution-diffusion mechanism. Hydrogen molecules dissociate into hydrogen atoms on the retentate side, diffuse through the film and re-associate on the permeate side. Each of the aforementioned steps may control the permeation rate depending on temperature, pressure, feed composition, and membrane thickness [4] [5]. Permeation of hydrogen through a dense palladium membrane is usually described by the Sieverts' law (equation 1) where the hydrogen partial pressure difference between two sides of the membrane is the driving force.

$$j_{H_2} = K_{H_2} (p_{H_2,ret}^n - p_{H_2,perm}^n) \quad \text{Eq 1}$$

K_{H_2} is the hydrogen permeance, $p_{H_2,ret}^n$ and $p_{H_2,perm}^n$ are the hydrogen partial pressure on the retentate and permeate side, respectively [2]. The hydrogen permeance is mostly determined by the temperature and the thickness of the membrane selective layer [2]. At relatively low pressure and membrane thickness higher than 10 μm , the rate-limiting step is the diffusion of hydrogen through the membrane. In this case, the factor n equals 0.5 [4]. Surface and gas transport limitations may result in deviations from the Sieverts' law, when $n > 0.5$.

U.S. DoE target currently set is $8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$ at 350 – 400 °C. Although the state of the art membranes already exceed this value, reaching $10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, they have been investigated applying pure hydrogen feed, so under real operating conditions this value is normally reduced by up to one order of magnitude because of depletion and concentration polarization effects [2].

2.2 Methodology and results

The steady state modeling of the fuel processor plant consistent with the Deliverable 4.2 and other studies found in the literature [2], [5] is performed by the aid of the simulation software Aspen Plus®.

Two different configurations of the fuel processor plant are analyzed, the baseline plant and the membrane steam reformer plant. The baseline plant includes heat integrated wall reactor (HIWAR), water gas shift reactor (WGSR) and PSA unit, while the second plant configuration is simplified by the integration of Pd membranes in the reformer. The latter removes the need for additional CO purification steps; thus the hydrogen rich gas produced in the membrane reformer is directly fed to the fuel cell stack downstream of the fuel processor unit.

The fuel processor efficiency is calculated according to equation 2:

$$\eta_{fp} = \frac{\dot{m}_{H_2} LHV_{H_2}}{\dot{m}_{NG} LHV_{NG}} \quad \text{Eq 2}$$

where \dot{m}_{H_2} and \dot{m}_{NG} are the mass flow of hydrogen and natural gas, respectively, LHV_{H_2} is the lower heating value of hydrogen and equals 120 MJ/kg, whereas LHV_{NG} is the lower heating value of natural gas which equals 46.5 MJ/kg.

If the effect of auxiliaries (natural gas compressor, water pumps, air compressor, and hydrogen compressor in some cases) is considered, total fuel processor efficiency is calculated according to the equation shown below:

$$\eta_{fp}^{tot} = \frac{\dot{m}_{H_2} LHV_{H_2} - W_{aux}^{ref}}{\dot{m}_{NG} LHV_{NG}} \quad \text{Eq 3}$$

where W_{aux}^{ref} is the sum of the fuel processor auxiliaries' electric consumption, calculated using equation 4.

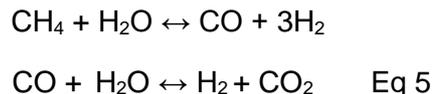
$$W_{aux} = \eta^{is} (h^{out} - h^{in}) \quad \text{Eq 4}$$

The fluid enthalpies are calculated according to Peng Robinson equation of state while the isentropic efficiency values of the pump/compressor, η^{is} , were obtained from Aspen Plus library [2].

The main operating parameters that control methane conversion and hydrogen production are temperature, pressure, and steam to carbon ratio. The fuel processor efficiency has been evaluated and calculated according to the aforementioned equations in different operating conditions for both the configurations. Steam to carbon ration (S/C), air to carbon ration (A/C), temperature, pressure, and hydrogen permeability have been varied in order to find the best operating conditions. The analysis results are shown and discussed below.

2.2.1 Pressure analysis

The strongly endothermic reaction of methane steam reforming ($\Delta H = 206$ kJ/mol), equation 5, which leads to a gas expansion of the product is favored by high temperatures and low pressure.



Increase in the reforming pressure results in a significant methane conversion decrease, therefore, in order to achieve a higher hydrogen molar fraction in the product it is necessary to keep the pressure as low as possible [6]. On the other hand, hydrogen separation efficiency through the selective membranes is determined by the pressure gradient between the permeate and the retentate side, which makes the reformer pressure the most critical parameter in the system design [2][5].

The behavior of the system as function of pressure regarding the fuel processor efficiency of the baseline and the membrane-based fuel processor unit is depicted in Fig. 1 and Fig. 2, respectively. The total fuel processor efficiency is considered here in order to take into account the pressure rise effects in the work required by the auxiliaries. It is assumed that the natural gas comes from the grid at 1 bar, and is compressed at the fuel processor inlet.

Varying the reformer pressure in the range from 5 to 15 bar, (at S/C = 3.73, T = 770 °C, and A/C = 200), total fuel processor efficiency of the baseline plant reaches the highest value at 8 bar, $\eta_{fp}^{tot} = 74.15\%$. The efficiency increases from 5 to 8 bar as the consequence of the improved PSA separation efficiency. The further pressure increase, as expected, is followed by the efficiency decrease prescribed to the hydrogen yield drop as the methane steam reforming reaction gets less favored as the pressure increases.

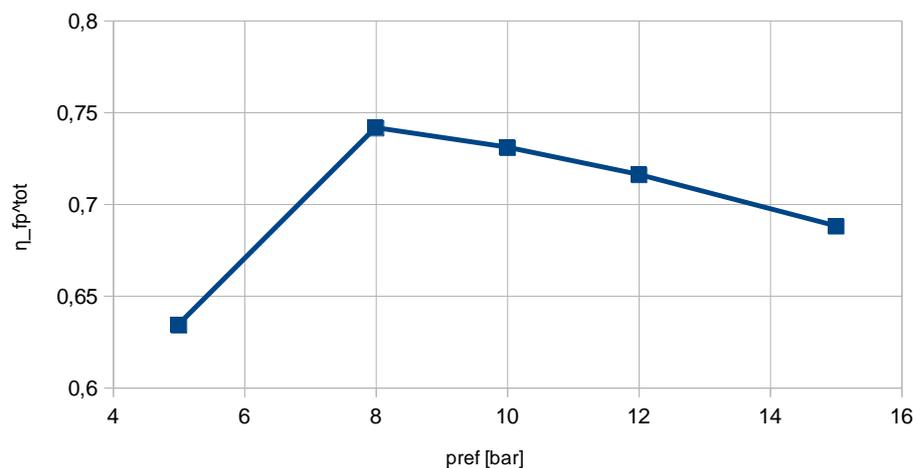


Figure 1. Total fuel processor efficiency of the baseline fuel processor plant as a function of the reforming pressure.

In contrast, increasing the membrane reformer pressure results in ever increasing values of the reformer efficiency, from 69.77 % at 10 bar to 74.08 % at 50 bar, due to the constant increase of the hydrogen permeation process driving force. After a sharp performance increase from 10 to 15 bar, further pressure increase is followed with a reduced performance increase, as can be seen in Fig. 2. Compared to the baseline plant, the efficiency of the membrane-based reformer plant is higher at the whole pressure range investigated, except at 5 bar when the pressure at the retentate side was too low to obtain satisfactory hydrogen pressure difference.

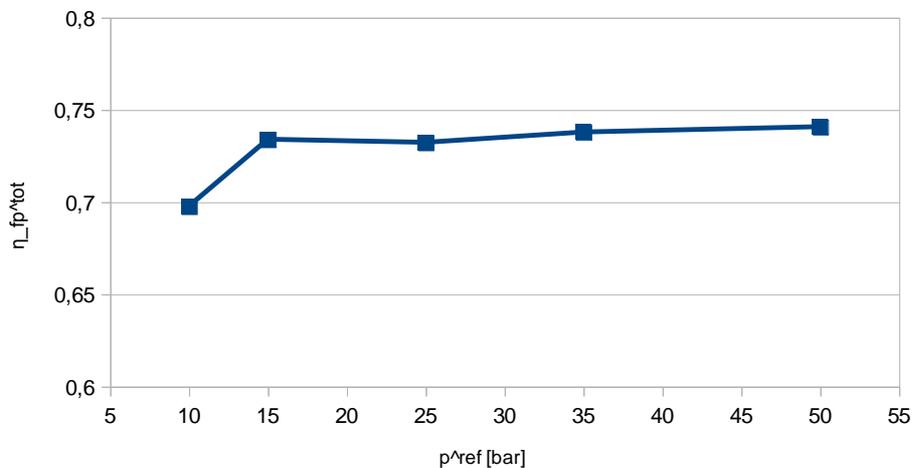


Figure 2. Total fuel processor efficiency of the membrane-based reformer as a function of the reforming pressure.

The increase in pressure from 35 bar to 50 bar results in the marginal performance improvement; the fuel processor efficiency increases by only 0.39 % (from 73.79 % to 74.08 %), which would not compensate for the increased plant costs.

In order to reduce the cost of the system operated at lower pressure, a membrane characterized by higher hydrogen permeance value could be utilized. For this reason, surface and permeability sensitivity analysis have been performed. Hydrogen permeance, K_{H_2} , U.S. DoE target currently set is $8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$ while state of the art membranes can already reach values of $10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$ at 350 – 400 °C. Since these values are obtained using pure H_2 as the feed gas and therefore the permeance under real operating conditions would be somewhat decreased [2], additional to these values in this analysis the permeance one order of magnitude lower than the DOE target has been taken into account. The results shown in Fig. 3 represent the effect of the membrane surface and permeance on the total fuel processor efficiency when membranes with permeance values $8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$ and $8.5 \times 10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$ are applied. It could be seen how the augmentation of the membrane surface and the hydrogen permeance of the membrane causes the performance increase but also that there is a saturation in the increase at 0.2 m^2 for both the membranes. It has been also found that there is a linear dependence between the hydrogen permeance and the membrane surface, as shown in Table 1.

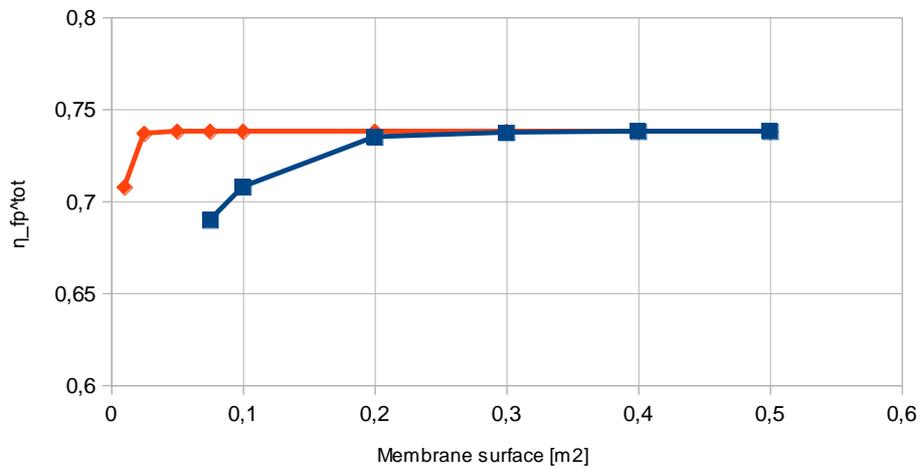


Figure 3. Total fuel processor efficiency of the membrane-based fuel processor plant at the reforming pressure 35 bar as a function of the membrane surface and the hydrogen permeance (blue line – hydrogen permeance = $8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, red line - hydrogen permeance = $8.5 \times 10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$).

Table 1. Required membrane surface for different hydrogen permeance values

Membrane surface [m ²]			Fuel processor efficiency [%]	Hydrogen Recovery [%]
$K_{H_2} = 8.5 \times 10^{-4} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$	$K_{H_2} = 8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$	$K_{H_2} = 8.5 \times 10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$		
3	0,3	0,03	80.11	97.6
2	0,2	0,02	79.87	97.6
1	0,1	0,01	76.89	95.3
0,75	0,075	0,0075	74.32	89.9

2.2.2 Temperature analysis

While high temperatures lead to higher conversions due to the endothermic nature of methane steam reforming reaction, they can be very detrimental for the stability of hydrogen permeable membranes. In order to determine an optimal trade-off between the reformer performance and the stability of the membranes, the fuel processor efficiency in the temperature range 590 – 850 °C is estimated and the results are shown in Fig. 4. In this analysis the variation in permeability with temperature is not considered.

Although the fuel processor efficiency is almost constant for temperatures higher than 600 °C, the hydrogen flow rate differs significantly being the highest at 770 °C, when it equals 3.092 kg/h. By decreasing the reforming temperature to 600 °C, the hydrogen flow rate drops to 2.251 kg/h, a value that does not meet the required hydrogen flow to feed 50 kW electrical power from the FC stack.

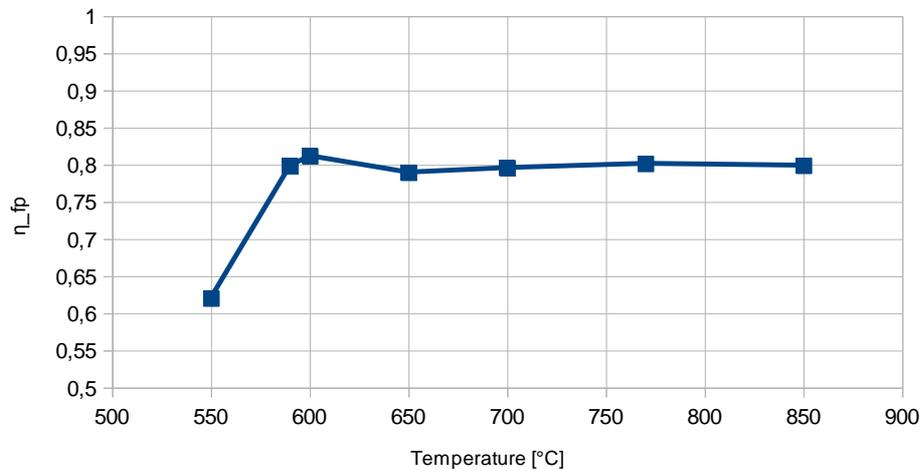


Figure 4. Fuel processor efficiency of the membrane reformer plant as a function of the reforming temperature.

It is also worth noting that at $T \leq 600$ °C there is enough chemical energy in the retentate, present in the form of unconverted reactants and non-recovered hydrogen, to sustain the thermal power requirements of the reformer without the need for additional delivery of natural gas to the catalytic burner, as can be seen in Table 2. For this reason, the reformer should not be operated at temperature below 600 °C since some chemical energy present in the retentate might be lost resulting in the fuel processor efficiency drop.

Moreover, although hydrogen yield is almost constant for temperatures higher than 650 °C, an efficiency drop can be observed at 850 °C and it could be even more pronounced at higher temperatures since the thermal power required to sustain the process increases, and so does the natural gas sent to the burner [2]. It can be concluded that the optimal temperature to perform the steam reforming process in membrane-based reformer should be 770 °C. This value is found considering only the thermo chemical parameters. However, it is noted that current Pd membrane reactor in operation (Tokyo Gas) apply 550-575 °C as the temperature for long run operation.

Table 2. Operating parameters at different temperatures

Temperature	NG Reformer mass flow [kg/h]	NG Burn mass flow [kg/h]	H2 mass flow [kg/h]	NG compress or Cons [kW]	Air Flow [kg/hr]	Other Reformer Aux Cons. [kW]
850	6,9	3,053	3,084	1,784	610,641	6,42
770	6,9	3,047	3,092	1,784	609,417	6,411
700	6,9	2,97	3,046	1,784	593,919	6,259
650	6,9	2,12	2,762	1,784	424	4,643
600	6,9	0,25	2,251	1,784	50	1,116
590	6,9	-	2,136	-	-	-
550	6,9	-	1,66	-	-	-

2.2.3 Steam to Carbon analysis

Steam to carbon ratio (S/C) is essential for the methane conversion and it is generally chosen to be higher than the stoichiometric ($S/C > 3$) [6]. Apart from shifting the equilibrium to the products side, higher S/C ratios minimize coke formation and deposition on the reforming catalysts which might lead to catalyst deactivation [7]. On the other hand, overestimation of the ratio could result in the decreased fuel processor efficiency due to increased heat requirements for the sustainability of the process [5]. The effect of steam to carbon ratio on the fuel processor efficiency of both the conventional and the membrane-based fuel processor is shown in Fig. 5.

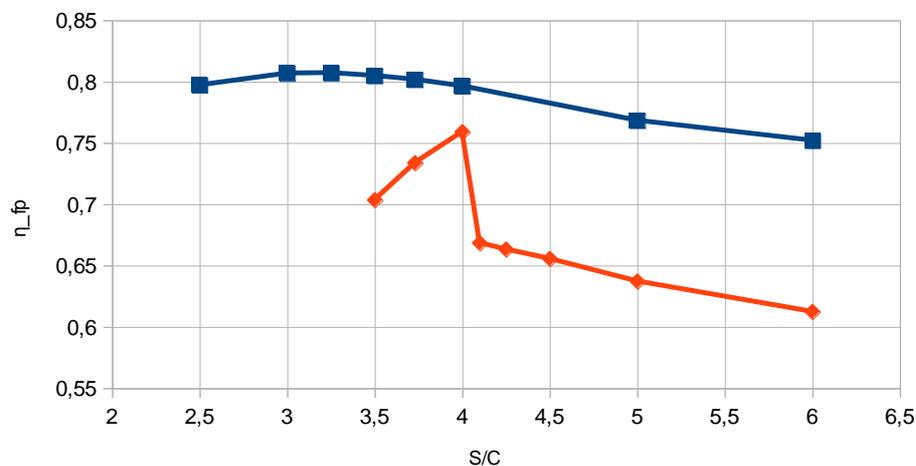


Figure 5. The effect of steam to carbon ratio, S/C, (red line – the baseline FP plant, blue line – membrane-based FP).

For both the configurations, there is a fuel processor efficiency increase as the ratio increases at lower values attributed to enhanced methane conversion and greater hydrogen production. When the increased heat requirements of the reforming process prevail the equilibrium shift effect at higher ratios, an efficiency drop can be observed. Compared to the traditional plant, fuel processor efficiency of the membrane-based reformer is higher at the whole S/C ratio investigated and its maximum efficiency can be obtained at a lower ratio, $S/C = 3.0$, while in the case of the traditional reformer it is obtainable at $S/C = 4.0$. Moreover, the A/C ratio of 200 is effective for the whole range of S/C ratios investigated, whereas in the traditional fuel processor plant case a higher air to carbon ratio is required at ratios lower than 3.5 to burn the reformer tail gasses in the combustor due to lower natural gas feed requirements ($A/C = 450$). Thermal energy of the reformer effluent at $S/C = 3.0$ in the baseline configuration plant was enough to sustain the thermal equilibrium due to increased amount of H_2 and unconverted CH_4 , as shown in Table 3.

Table 3. Composition of the baseline plant steam reformer effluent at S/C =3 and S/C=3.73.

S/C = 3		S/C = 3,73	
Component	Fraction	Component	Fraction
CH ₄	0,135	CH ₄	0,111
O ₂	-	O ₂	-
N ₂	0,012	N ₂	0,012
CO ₂	0,43	CO ₂	0,46
H ₂ O	0,024	H ₂ O	0,025
CO	0,056	CO	0,049
C ₂ H ₆	trace	C ₂ H ₆	trace
C ₃ H ₈	trace	C ₃ H ₈	trace
H ₂	0,357	H ₂	0,377
NH ₃	< 0.001	NH ₃	< 0.001
Q _{ref}	27,24 kW	Q _{ref}	28,5 kW

2.2.4 Air to Carbon analysis

Behavior of the fuel processor unit in the traditional configuration and in the membrane-based reformer configuration at the variation of air to carbon ratio (A/C) is shown in Fig. 6.

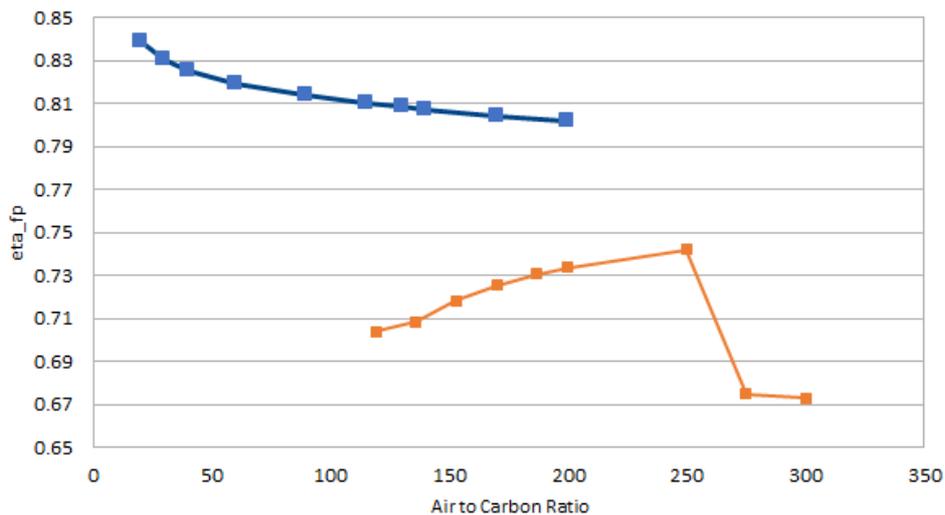


Figure 6. The effect of air to carbon ratio on the membrane based fuel processor (blue curve) and on the baseline FP configuration (red curve).

For the baseline fuel processor varying the A/C ratio in the range (119 – 300) the maximum fuel processor efficiency, $\eta_{fp} = 74.20\%$, is achieved at A/C = 250, while further increase of the ratio results in an efficiency decrease. In contrast, the fuel processor efficiency of the



membrane-based reformer unit increases with the decreasing A/C ratio from 300 to 20, reaching the highest value, $\eta_{fp} = 83.84 \%$, at A/C = 20.

The results presented above have shown the possibility of improving the fuel processor efficiency by implementing membranes into the steam reformer. It has been also proven that the operating parameters, namely the reforming temperature, pressure, S/C and A/C ratio, of such a system determine its performance and the most favorable conditions have been found. The membrane-based reformer operating at $T_{ref} = 770 \text{ }^\circ\text{C}$, $p_{ref} = 35 \text{ bar}$, S/C = 3.73 and A/C = 200 obtains $\eta_{fp} = 80.17 \%$ and $\eta_{fp}^{tot} = 73.80 \%$, while the highest efficiency of the conventional fuel processor at the operating conditions investigated is $\eta_{fp} = 75.49 \%$ and $\eta_{fp}^{tot} = 74.15 \%$. FP efficiency could be improved by 8.63% by implementing membranes into a reformer. It has been also found that the hydrogen permeance of the membrane integrated into the reformer determines the fuel processor efficiency and the membrane area required to obtain needed hydrogen flow and thus the plant costs. Changing a membrane characterized by $K_{H_2} = 8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, which is the current DOE target for 400 °C, with a membrane $K_{H_2} = 8.5 \times 10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, required membrane area to reach $\eta_{fp} = 80 \%$, at the aforementioned operating condition, could be reduced 10 times.



3 Preliminary configurations of the membrane-based fuel processor

3.1 Types of membrane-based reactors

The simplest way to build a membrane reactor is the tube in tube configuration. Such configuration has been extensively applied in lab-scale experiments, most often with tubular membranes of 5-10 mm diameter and catalysts having a particle size around 1 mm in small laboratory reactors where the catalyst was placed as a packed bed inside or around the membrane tube.

A good match of the hydrogen production rate and the hydrogen flux is required to shift reaction equilibria, which requires a sufficiently large membrane area per catalyst volume. For industrial (enlarged) systems the configuration that enables an increase of a membrane area per reactor volume is the tubes in shell configuration where a bundle of membrane tubes is located inside a shell. A catalyst can be packed either inside the tubes or inside the shell. Tubes with large diameter offer only a low membrane area per catalyst volume, *e.g.*, $400 \text{ m}^2/\text{m}^3$ for a 1 cm outer diameter membrane tube having the catalyst inside. For high reaction rates (high temperature, highly active catalyst) this might not be sufficient. Moreover, the transport of hydrogen from the catalyst surface where it is produced to the membrane surface where it is removed may come at a significant resistance depending on the design of the system. In general, the more intimately the catalyst and membrane are connected the lower the mass transfer resistance for hydrogen is. The additional increase of the membrane surface could be obtained by increasing the number of tubes and/or giving a particular shape to the membranes [8].

3.2 Proposed configurations starting from HIWAR geometry

Heat Wall Integrated Reactor (HIWAR) designed by Helbio used in this project allows for steam reforming reactions and combustion reaction to take place simultaneously in the same reactor. This heat exchanger type of a reactor consists of a bundle of tubes inside which a reforming reaction takes place (Figure 7). The combustion catalyst is deposited on the inner surface of the tubes so the heat demanded by the reforming process is transported via the metallic wall very quickly to the steam reforming reaction zone. This enables the design of a very efficient and compact reactor [9].

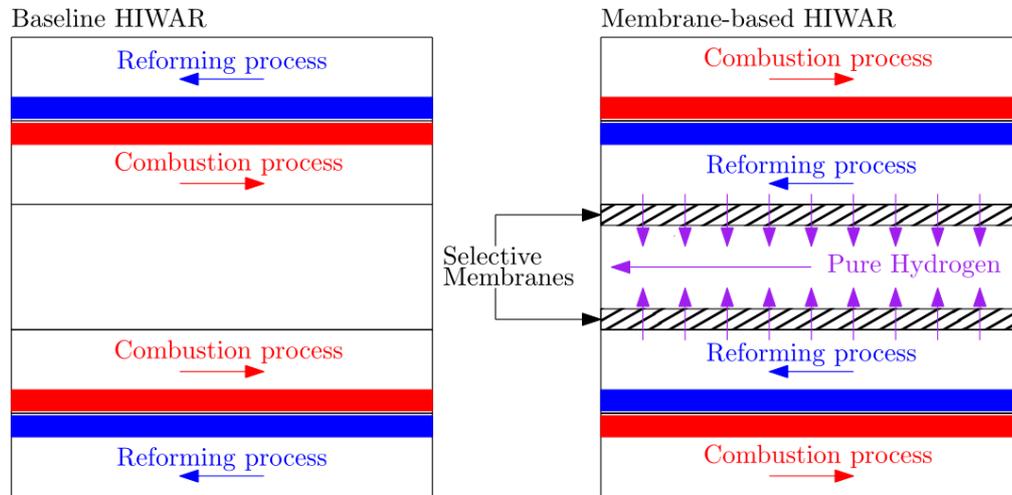


Figure 7. Baseline HIWAR and membrane-based HIWAR scheme.

If the combustion and the reforming zones are switched, it is possible to integrate hydrogen membranes into this kind of a reactor and obtain even higher efficiency without significant changes in its configuration (Figure 7b). In this way, the combustion would take place at the outer surface of a tube providing the heat required by steam reforming reaction at the inner surface of the tube. Produced hydrogen would penetrate through a Pd-based membrane which faces the reforming side to a zone originally marked as a no-flux one. The useful tube length would stay the same, so does the catalyst surface, that is the heat transfer surface. Since there are 7 tubes in the reactor with 26 mm diameter and 1670.5 mm length, total membrane surface that could be achieved is 0.955 m², 0.13645 m² per tube. At $T_{ref} = 770$ °C, $p_{ref} = 35$ bar, and $S/C = 3.73$, the operating conditions under which the membrane-based reformer unit has shown the best performance in this simulation study, taking into account the hydrogen flow rate 3.6761 kg h⁻¹ required by the 50 kW PEM FC stack, the required surface, considering $K_{H_2} = 8.5 \times 10^{-3}$ mol / (m²sPa^{0.5}), has been calculated equal to 0.250 m². Therefore, the selective membranes can be easily integrated in the baseline configuration of the HIWAR. Moreover, if all the available surface of the reactor for membrane integration is used, the permeability could be as low as $K_{H_2} = 2.2 \times 10^{-3}$ mol / (m²sPa^{0.5}) without affecting the performance.



4 Conclusion

The steady state modeling of the fuel processor plant consistent with the Deliverable 4.2 has been performed by the aid of the simulation software Aspen Plus®. The aim of this work is to define optimal operating conditions.

The simulation results have proven the possibility of improving the fuel processor efficiency by implementing membranes into the steam reformer. It has been also shown that the operating parameters, namely the reforming temperature, pressure, S/C and A/C ratio, of such a system determine its performance and the most favorable conditions have been found. The conventional fuel processor unit and the membrane-based reformer unit are compared in terms of fuel processor efficiency. The membrane-based reformer operating at $T_{\text{ref}} = 770 \text{ °C}$, $p_{\text{ref}} = 35 \text{ bar}$, $S/C = 3.73$ and $A/C = 200$ obtains maximum efficiency $\eta_{fp} = 80.17 \%$ and $\eta_{fp}^{\text{tot}} = 73.80 \%$, while the highest efficiency of the conventional fuel processor at the operating conditions investigated is $\eta_{fp} = 75.49 \%$ and $\eta_{fp}^{\text{tot}} = 74.15 \%$. It has been also found that the hydrogen permeance of the membrane integrated into the reformer determines the fuel processor efficiency and the membrane area required to obtain the needed hydrogen flow thus the plant costs. Changing a membrane characterized by $K_{H_2} = 8.5 \times 10^{-3} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, which is the current DOE target, with a membrane $K_{H_2} = 8.5 \times 10^{-2} \text{ mol} / (\text{m}^2\text{sPa}^{0.5})$, required membrane area to reach $\eta_{fp} = 80 \%$, at the aforementioned operating condition, could be reduced 10 times.

An engineering configuration has also been proposed for the steam reformer reactor to integrate the membranes, which does not significantly change the reformer size by making use of internal 'blank tube' for the 'pure' hydrogen exit flow. The overall size of the natural gas reforming system is thus reduced and simplified by removing the pressure swing absorption system.

In conclusion, the optimal working parameters for the system have been summarised, reactor configurations proposed, and the future perspectives for experimental investigations of the optimal reactor configuration and optimal working parameters proposed.



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