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# Computational fluid dynamics modeling of the millisecond methane steam reforming in microchannel reactors for hydrogen production

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Methane steam reforming coupled with methane catalytic combustion in microchannel reactors for the production of hydrogen was investigated by means of computational fluid dynamics. Special emphasis is placed on developing general guidelines for the design of integrated micro-chemical systems for the rapid production of hydrogen. Important design issues, specifically heat and mass transfer, catalyst, dimension, and flow arrangement, were explored. The relative importance of different transport phenomena was quantitatively evaluated, and some strategies for intensifying the reforming process were proposed. The results highlighted the importance of process intensification in achieving the rapid production of hydrogen. High heat and mass transfer rates derived from miniaturization of the chemical system are insufficient for process intensification. Improvement of the reforming catalyst is also essential. The efficiency of heat exchange can be improved greatly if the reactor dimension is properly designed. Thermal management is required to improve the reliability of the integrated system. Co-current heat exchange improves the thermal uniformity in the system. The catalyst loading is a key factor determining reactor performance, and must be carefully designed. Finally, engineering maps were constructed to achieve the desired power output, and favorable operating conditions for the rapid production of hydrogen were identified.

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## 1. Introduction

In recent years, there has been an increasing interest in the development of the production of synthesis gas from hydrocarbons due to its importance in many practical applications such as fuel cells and internal combustion engines.<sup>1–4</sup> In industry, synthesis gas is produced by steam reforming of natural gas.<sup>5,6</sup> This reaction is endothermic in nature and requires additional heat to proceed. Therefore, this process is carried out in multi-tubular fixed-bed reactors inserted into a large furnace to supply the heat required. The industrial process operates at temperatures of around 900 °C under high pressures over a Ni/ $\alpha$ -Al<sub>2</sub>O<sub>3</sub> catalyst with contact times exceeding one second.<sup>5,6</sup>

Many efforts have been made recently to understand the catalytic partial oxidation of methane to produce synthesis gas at short contact times.<sup>7,8</sup> In contrast, the rate of the steam reforming reaction of methane is relatively slow.<sup>5,6</sup> Interestingly, microfabrication techniques offer new opportunities for the development of methane steam reforming at short contact times.<sup>9–12</sup> The rapid, efficient conversion of methane to synthesis gas shows great promise for fuel cell applications.<sup>13,14</sup>

Similarly, the rapid, efficient conversion of small amounts of methane to a synthesis gas mixture would improve spark ignition engine and catalytic converter performance while lowering emissions.<sup>15,16</sup> Therefore, further research efforts are needed in the intensification of the steam reforming process.

Valuable intermediates and products can be made under short contact-time conditions.<sup>17,18</sup> Attempts to implement the commercial steam reforming technology in small dimensions have been made. In particular, the steam reforming of methane over highly active catalysts at short contact times is an efficient method for the production of hydrogen.<sup>9–12</sup> Currently, a fast reforming process is a pressing and important topic of study in light of potential hydrogen utilization for fuel-cell-powered automobiles.<sup>16</sup> Consequently, there is considerable interest in the “on-board” conversion of methane to hydrogen at short contact times. Millisecond methane steam reforming has been reported in the literature, including in microchannel or microstructured reactors,<sup>9–12,19–22</sup> and on highly active catalysts such as rhodium.<sup>23</sup> The reaction proceeded at millisecond contact times is feasible due to its fast kinetics and transport,<sup>10,11</sup> and thus hydrogen production can be intensified by hundred to thousand times. Furthermore, the methane steam reforming process has been designed to operate at sub-millisecond contact times,<sup>11</sup> even at less than 100 microseconds at the expense of low equilibrium conversion.<sup>9</sup> Interestingly, recent literature has suggested that methane steam

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reforming over nickel at millisecond contact times is in principle feasible at increased catalyst loadings.<sup>24</sup>

The very high transport rates achievable in microchannel reactors allow for operating highly endothermic and exothermic processes isothermally, which is particularly important in achieving process intensification.<sup>9,19,21,25,26</sup> Process intensification is a strategy aimed at transforming conventional chemical processes into more economical, productive and green processes, in the form of enhanced heat and mass transfer.<sup>21,22,26,27</sup> This technology accelerates processes by enabling reactions to occur at rates up to 1000 times faster than those in conventional systems, which has the potential to reduce reaction times significantly.<sup>9-12,19,20,26,27</sup> Furthermore, these processes can be operated under near isothermal conditions to prevent hot spots and thermal runaway.<sup>25</sup> The net result is that microchannel process technology can significantly improve the efficiency of the production of hydrogen from methane steam reforming.<sup>25-28</sup>

Microchannel reactors have many advantages for chemical production and process development, especially the application in fuel-cell-powered automobiles, because they operate at much shorter residence times, require much simpler equipment, and can be further “scaled-out” or “numbered-up” to achieve any desired large-scale plant capacity.<sup>16</sup> In this way, the capital and operating cost may be reduced, and a higher efficiency may be achieved.<sup>29,30</sup> Recent studies have demonstrated that microchannel reactors exhibit high heat fluxes and high heat exchange efficiencies.<sup>31,32</sup> These chemical processing advantages can be derived from increased heat and mass transfer in small dimensions, eventually leading to improved yield for a steam reforming process.

Much research has focused on the design strategy of millisecond reforming process. The concept of millisecond reforming process may appear to be simple, but the design strategy presents significant challenges that have not yet been fully addressed. To address this issue, great efforts have been made in recent years to illustrate how process intensification can be achieved through microchannel reactors.<sup>33-36</sup> Reaction engineering analysis is necessary to develop new process-intensifying methods to meet the need of the rapid production of hydrogen from steam reforming. To successfully design microfabricated chemical systems and to further intensify the involved chemical process, a detailed understanding of the underlying mechanism of the transport phenomena is also necessary.

The development of design strategies for the rapid production of hydrogen in micro-devices needed for novel fuel-cell-based power sources is the motivation for this study. In order to address the challenges in realizing this vision, the primary focus of this study is on short contact time reaction systems, and specifically on methane steam reforming thermally coupled with catalytic combustion in a microchannel reactor. A microchannel reactor that enabled efficient heat exchange between exothermic and endothermic reactions in an intensified manner was modeled, as an example of short-contact-time reactor modeling. Numerical simulations with detailed transport and chemical kinetic models were conducted to

understand the transport characteristics involved in the system. Some important design issues, such as heat and mass transfer, catalyst, dimension, and flow arrangement, were explored. The primary objective of this study is to develop the general guidelines for the design of integrated micro-chemical systems for the rapid production of hydrogen. Emphasis is placed on how to achieve the rapid production of hydrogen from methane steam reforming by means of process intensification.

## 2. Model development

Mathematical modeling plays an important role in the development of a chemical reactor.<sup>37-40</sup> Computational fluid dynamics is a modeling technique with powerful visualization capabilities to deal with various geometry characteristics and boundary conditions. Due to the dramatic increase of computing power, computational fluid dynamics has become an efficient tool for reformer modeling,<sup>41,42</sup> and has become an increasingly important platform for reformer design.<sup>43,44</sup> Additionally, computational fluid dynamics modeling is necessary to provide in-depth understanding of the operating characteristics of steam reforming reactors,<sup>41-44</sup> and to evaluate the benefits and disadvantages associated with new reformer designs.<sup>12</sup> Various novel designs for steam reforming reactors have been developed by using computational fluid dynamics modeling.<sup>12</sup> Computational fluid dynamics modeling serves not only as an efficient design tool,<sup>41-44</sup> but also as a means to interpret experimental data.<sup>45,46</sup> Furthermore, a computational fluid dynamics modeling engineering approach to new reformer designs can avoid costly, iterative experimental design processes.<sup>47</sup>

### 2.1. Description of the chemical system

The chemical system investigated is the methane steam reforming coupled with catalytic combustion occurring in a microchannel reactor. A parallel plate with supported washcoat coatings on opposite sides separates the catalytic combustion channel from the reforming channel. A schematic diagram of the reactor modeled is given in Fig. 1. The endothermic reaction and the exothermic reaction take place on opposite sides of a wall. The separating wall serves as an efficient and compact heat exchanger. Under certain operating conditions, all the combustion (or reforming) channels behave essentially alike. The use of thin washcoats can maximize the potential for very high heat fluxes.<sup>48-50</sup> Recent studies have demonstrated that the thermal coupling between the endothermic and exothermic reactions is important in the design of small-scale chemical systems,<sup>10,11,49,50</sup> and the integration of multiple microchannel reactors presents significant challenges in thermal management.<sup>51,52</sup> The primary focus of this research is on exploring reactor design for optimal thermal coupling and hydrogen production.

There are many active metals used to promote the methane steam reforming process. In the present work, rhodium is considered due to its very high catalytic performance for the steam reforming of methane at very short contact times,<sup>53,54</sup> but



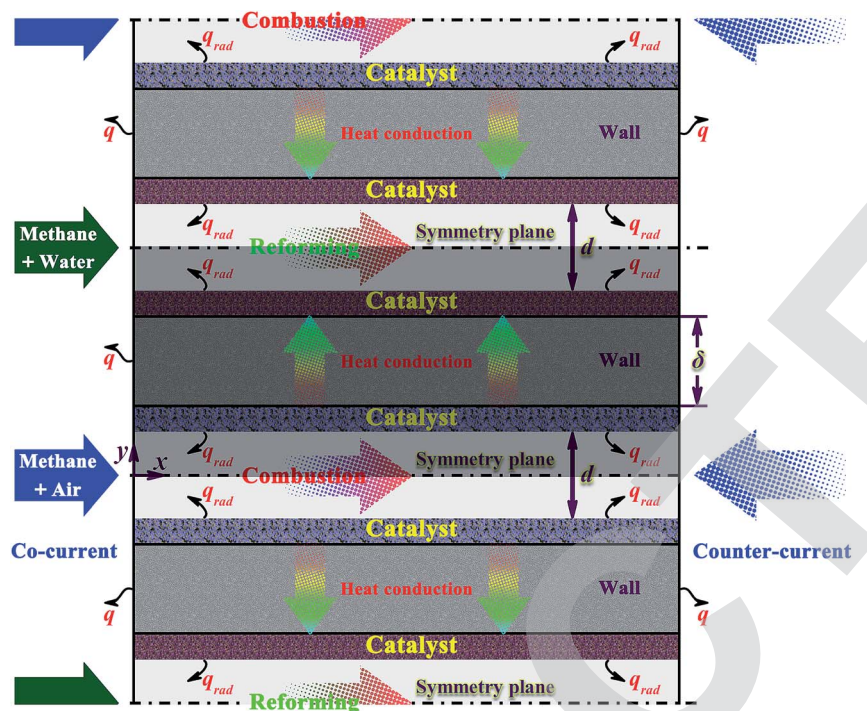


Fig. 1 Schematic diagram of the microchannel reactor with alternating combustion and reforming channels modeled in this study. Due to the inherent symmetry of the reactor, only half of each channel, the separating plate, and the two catalyst washcoats are modeled to reduce the computational cost. The shaded region represents the computational domain of the microchannel reactor modeled in this study.

nickel is the most abundantly used metal. Furthermore, nickel-based catalysts can withstand very high temperature and exhibits good mechanical strength. Therefore, the issue that needs to be addressed is whether nickel can achieve the rapid production of hydrogen from methane steam reforming, as will be discussed in detail later.

In order to evaluate the effect of various operating conditions and design parameters, a reference point is established. The “base case” given in Table 1 is defined as a typical set of operating conditions and design parameters. The term “base case” refers to a context in which the conversion is almost complete on each side of the reactor, the formation of hot-spots or cold-spots is nearly impossible, and there is a good thermal balance between the endothermic and exothermic reactions. The molar steam-to-carbon ratio at the inlet is set as 3.0.<sup>55</sup> If desired, the reactor width is set as 10.0 mm to obtain flow rates. A full-scale reactor design may include a much larger number of parallel plate channels, usually ranging from a few hundreds to many thousands or even more. Due to the inherent symmetry of the reactor, only half of each channel, the connecting plate, and the two catalyst washcoats are modeled to reduce the computational cost. The shaded region in Fig. 1 represents the computational domain of the reactor modeled in this paper.

On the other hand, high pressures are obviously desirable to methane reformers, and this will be the subject of future work. If the reactor is operated at high pressures, particular emphasis should be placed on the thermodynamic limit of the small scale reforming system. In addition, the thickness of the wall separating the reforming and combustion side should be several

millimeters as required to maintain a high-pressure differential. Furthermore, the small scale reformer should also be designed to withstand several tens of atmospheres of pressure differential at high temperatures.

## 2.2. Mathematical model

Due to the small dimension of the system, transport processes are significantly affected by diffusion. However, the length scale is still sufficiently large in comparison with the mean free path. Therefore, classical, continuum descriptions are appropriate. To reduce the complexity in the modeling of the system, it is assumed that ideal gas law is applicable and the system is in a steady state. Since the Reynolds numbers are less than 280, reactive flows in each of the channels are laminar, thus making possible to accurately simulate reaction and transport processes. The radiative heat transfer between the surfaces of the plates is modeled on each side of the reactor, as thermal radiation may be significant.<sup>52</sup> Simple estimates indicate that for the optical path lengths and species partial pressures of interest in this study, gas-phase radiative heat transfer is negligible in comparison with convection and surface-to-surface radiative heat transfer. The commercial software ANSYS Fluent<sup>56</sup> is adopted. Detailed reaction mechanisms and multicomponent transport are included in the model. Despite the small dimensions involved, gas-phase combustion may be significant, and thus both gas-phase and surface reaction mechanisms are included in the model. External subroutines are linked to the ANSYS Fluent software in order to model detailed gas-phase and surface reaction mechanisms. This is



Table 1 Parameters of the model used for the base case

	Combustion side	Reforming side	Unit
<b>Geometry</b>			
Channel length	50.0		mm
Channel height	0.2		mm
<b>Solid wall</b>			
Thermal conductivity	80.0 (300 K)		W (m K) <sup>-1</sup>
Thickness	0.2		mm
External heat loss coefficient	20		W (m <sup>2</sup> K) <sup>-1</sup>
<b>Gas phase</b>			
Inlet methane-air equivalence ratio	0.8	—	—
Inlet molar steam-to-carbon ratio	—	3.0	—
Inlet pressure	0.1	0.1	MPa
Inlet temperature	300	400	K
Inlet velocity	6.0	2.0	m s <sup>-1</sup>
<b>Catalyst</b>			
Porosity	0.5	0.6	—
Tortuosity factor	3	3	—
Catalyst/geometric surface area	20	20	—
Mean pore diameter	20	20	nm
Thickness	0.08	0.08	mm
Density of catalyst surface sites	2.72 × 10 <sup>-9</sup> (platinum)	2.72 × 10 <sup>-9</sup> (rhodium) 2.66 × 10 <sup>-9</sup> (nickel)	mol cm <sup>-2</sup>

implemented by using CHEMKIN-CFD (developed by ANSYS Inc.; further information is also available in the literature<sup>56</sup>), which is a plug-in chemistry solver that can be linked to the ANSYS Fluent software,<sup>56</sup> to solve the chemistry part of the solution. The equations for conservation of total mass, momentum, energy, and chemical species are solved in the gas phase.

Continuity equation:

$$\frac{\partial(\rho u)}{\partial x} + \frac{\partial(\rho v)}{\partial y} = 0. \quad (1)$$

Momentum equations:

$$\frac{\partial(\rho uu)}{\partial x} + \frac{\partial(\rho vu)}{\partial y} + \frac{\partial p}{\partial x} - \frac{\partial}{\partial x} \left[ 2\mu \frac{\partial u}{\partial x} - \frac{2}{3}\mu \left( \frac{\partial u}{\partial x} + \frac{\partial v}{\partial y} \right) \right] - \frac{\partial}{\partial y} \left[ \mu \left( \frac{\partial u}{\partial y} + \frac{\partial v}{\partial x} \right) \right] = 0, \quad (2)$$

$$\frac{\partial(\rho uv)}{\partial x} + \frac{\partial(\rho vv)}{\partial y} + \frac{\partial p}{\partial y} - \frac{\partial}{\partial x} \left[ \mu \left( \frac{\partial v}{\partial x} + \frac{\partial u}{\partial y} \right) \right] - \frac{\partial}{\partial y} \left[ 2\mu \frac{\partial v}{\partial y} - \frac{2}{3}\mu \left( \frac{\partial u}{\partial x} + \frac{\partial v}{\partial y} \right) \right] = 0. \quad (3)$$

Energy equation:

$$\frac{\partial(\rho uh)}{\partial x} + \frac{\partial(\rho vh)}{\partial y} + \frac{\partial}{\partial x} \left( \rho \sum_{k=1}^{K_g} Y_k h_k V_{k,x} - \lambda_g \frac{\partial T}{\partial x} \right) + \frac{\partial}{\partial y} \left( \rho \sum_{k=1}^{K_g} Y_k h_k V_{k,y} - \lambda_g \frac{\partial T}{\partial y} \right) = 0. \quad (4)$$

Chemical species balance in the gas phase:

$$\frac{\partial(\rho u Y_k)}{\partial x} + \frac{\partial(\rho v Y_k)}{\partial y} + \frac{\partial}{\partial x} (\rho Y_k V_{k,x}) + \frac{\partial}{\partial y} (\rho Y_k V_{k,y}) - \dot{\omega}_k W_k = 0, \quad k = 1, \dots, K_g. \quad (5)$$

in the above equations,  $\rho$  is the density of the mixture;  $u$  and  $v$  are the streamwise and transverse components of the flow velocity of a fluid, respectively;  $x$  and  $y$  are the streamwise and transverse coordinates in the system, respectively;  $p$ ,  $\mu$ , and  $\lambda_g$  are the pressure, dynamic viscosity, and thermal conductivity of the mixture, respectively;  $h$  is the specific enthalpy of the mixture;  $T$  is the absolute temperature;  $K_g$  is the total number of gaseous species;  $Y_k$ ,  $h_k$ ,  $\dot{\omega}_k$ , and  $W_k$  are the mass fraction, specific enthalpy, rate of appearance, and relative molecular mass of the  $k$ -th gaseous species, respectively;  $V_{k,x}$  and  $V_{k,y}$  are the streamwise and transverse components of the diffusion velocity of the  $k$ -th gaseous species, respectively.

The diffusion velocity vector of the  $k$ -th gaseous species can be expressed as follows:<sup>57</sup>

$$\vec{V}_k = -D_{k,m} \nabla \left[ \ln \left( \frac{Y_k \bar{W}}{W_k} \right) \right] + \left[ \frac{D_k^T W}{\rho Y_k \bar{W}} \right] \nabla (\ln T), \quad (6)$$



wherein the term  $D_{k,m}$  is the diffusivity for the  $k$ -th gaseous species in the mixture,  $\bar{W}$  is the relative molecular mass of the mixture,  $D_k^T$  is the thermal diffusivity of the  $k$ -th gaseous species, and the nabla symbol  $\nabla$  denotes the differential operator del, *i.e.*, nabla operator.

The ideal gas equation of state is

$$p = \frac{\rho RT}{\bar{W}}, \quad (7)$$

where  $R$  denotes the ideal gas constant. The caloric equation of state can be written as

$$h_k = h_k^o(T_o) + \int_{T_o}^T c_{p,k} dT, \quad (8)$$

where  $h_k^o$  is the specific enthalpy of the  $k$ -th gaseous species at the specified temperature  $T_o$ , and  $c_{p,k}$  is the specific heat capacity of the  $k$ -th gaseous species at constant pressure.

The equation of surface species coverage can be written as

$$\sigma_m \frac{\dot{s}_m}{\Gamma} = 0 \quad \text{with } m = K_g + 1, \dots, K_g + K_s, \quad (9)$$

in which  $\sigma_m$  is the site occupancy of the  $m$ -th surface species,  $\dot{s}_m$  is the rate of appearance of the  $m$ -th surface species,  $\Gamma$  is the density of catalyst surface sites,  $m$  is the total number of gaseous and surface species, and  $K_s$  is the total number of surface species. The adsorption rate constant is given by

$$k_{ad,k} = \left( \frac{2}{2 - s\theta_{free}} \right) \frac{s}{\Gamma^m} \sqrt{\frac{RT}{2\pi W_k}}. \quad (10)$$

where  $s$  is the sticking coefficient and  $\theta_{free}$  is the surface coverage of free sites.

The energy balance for the separating wall can be written as

$$\frac{\partial}{\partial x} \left( \lambda_s \frac{\partial T}{\partial x} \right) + \frac{\partial}{\partial y} \left( \lambda_s \frac{\partial T}{\partial y} \right) = 0, \quad (11)$$

wherein the term  $\lambda_s$  is the thermal conductivity of the solid wall. A high wall thermal conductivity is used, as given in Table 1. High wall thermal conductivities make possible efficient heat exchange within a micro-chemical system.<sup>10–12</sup>

Each of the gas–washcoat interfaces is defined by species and energy boundary conditions. The boundary condition for the gaseous species at each of the specified gas–washcoat interfaces is given by

$$(\rho Y_k V_{k,y})_{interface} + \eta F_{cat/geo} W_k (\dot{s}_k)_{interface} = 0, \quad k = 1, \dots, K_g, \quad (12)$$

in which  $\eta$  is the effectiveness factor; the ratio of catalytic surface area,  $A_{catalytic}$ , to geometric surface area,  $A_{geometric}$ , of the catalyst washcoat, *i.e.*, catalyst/geometric surface area, is defined as follows:<sup>58</sup>

$$F_{cat/geo} = \frac{A_{catalytic}}{A_{geometric}}. \quad (13)$$

The diffusional limitations inside the catalyst washcoat is accounted for:<sup>59</sup>

$$\eta = \frac{\dot{s}_i^{eff}}{\dot{s}_i} = \frac{\tanh(\Phi)}{\Phi}, \quad (14)$$

with Thiele modulus  $\Phi$  given by

$$\Phi = \delta_{catalyst} \left( \frac{\dot{s}_i \gamma}{D_{i,eff} C_{i,interface}} \right)^{0.5}. \quad (15)$$

In the above equation,  $\dot{s}_i^{eff}$  is the effective rate of appearance of the  $i$ -th species on the surface of the catalyst,  $\delta$  is the thickness of the washcoat, and  $C_{i,interface}$  is the concentration of the  $i$ -th species at the gas–washcoat interface. The catalytic surface area per unit volume of the washcoat,  $\gamma$ , is defined as

$$\gamma = \frac{F_{cat/geo}}{\delta_{catalyst}}. \quad (16)$$

The effective diffusivity of the  $i$ -th species inside the catalyst washcoat,  $D_{i,eff}$ , is computed by adjusting for catalyst porosity and tortuosity

$$\frac{1}{D_{i,eff}} = \frac{\tau_p}{\varepsilon_p} \left( \frac{1}{D_{i,molecular}} + \frac{1}{D_{i,Knudsen}} \right), \quad (17)$$

in which  $\tau_p$  is the tortuosity factor for diffusion inside the washcoat,  $\varepsilon_p$  is the porosity of the washcoat, and  $D_{i,molecular}$  is the diffusivity for molecular diffusion of the  $i$ -th species in the mixture. The diffusivity for Knudsen diffusion of the  $i$ -th species is defined as

$$D_{i,Knudsen} = \frac{d_{pore}}{3} \sqrt{\frac{8RT}{\pi W_i}}. \quad (18)$$

where  $d_{pore}$  represents the mean pore diameter.

At each of the specified gas–washcoat interfaces, heat generation by reactions must be included in the model, and thus the energy boundary condition can be written as

$$\dot{q}_{rad} - \lambda_g \left( \frac{\partial T}{\partial y} \right)_{interface^-} + \lambda_s \left( \frac{\partial T}{\partial y} \right)_{interface^+} + \sum_{k=1}^{K_g} (\dot{s}_k h_k W_k)_{interface} = 0. \quad (19)$$

in which  $\dot{q}_{rad}$  is the radiative heat flux. The washcoat is considered isothermal in the transverse direction due to its small thickness. The total heat loss to the ambient is modeled as

$$q = h_o(T_{w,o} - T_{amb}) + q_{rad,\infty}. \quad (20)$$

where  $q$  is the heat flux,  $h_o$  is the external heat loss coefficient,  $T_{w,o}$  is temperature at the external surface of the solid wall, and  $T_{amb}$  is the ambient temperature. The net radiation flux to the ambient,  $q_{rad,\infty}$ , is modeled as

$$q_{rad,\infty} = \varepsilon_{s-\infty} F_{s-\infty} \sigma (T_{w,o}^4 - T_{amb}^4). \quad (21)$$

in which  $\varepsilon_{s-\infty}$  is the effective emissivity of the external surface of the solid wall,  $F_{s-\infty}$  is the view factor from the external surface of the solid wall to the ambient environment, and is assumed to be unity,<sup>60</sup> and  $\sigma$  is the Stefan–Boltzmann constant.



### 2.3. Chemical kinetic model

The chemical reactions used in the reactor modeling are catalytic combustion, steam reforming, water gas-shift, and reverse methanation. The catalytic combustion of methane is somewhat complicated. The possible occurrence of gas-phase combustion is discussed below. The contribution of gas-phase reactions is usually assumed to be negligible in small scale combustion systems. However, gas-phase reactions may be significant,<sup>61</sup> especially at high temperatures;<sup>62,63</sup> gas-phase reactions can be sustained in a confined space with a dimension as small as several sub-millimeters.<sup>61</sup> In the present investigation, the reaction temperature is high enough so that gas-phase combustion is possible. Therefore, gas-phase reactions are not negligible on the combustion side, thus increasing the complexity in the modeling of the combustion kinetics. On the reforming side, gas-phase reactions are negligible, as will be discussed later. Note that the catalysts defined in this paper are the same as those used in the reaction mechanisms taken from the literature works described below, but the characteristics of the catalyst structure such as the tortuosity factor, porosity, and catalyst/geometric surface area are user-defined. Furthermore, only isotropic catalyst structures are considered in the present work, as described in detail in the previous section.

For the combustion process, the gas-phase reaction mechanism used is the Leeds methane oxidation mechanism.<sup>64,65</sup> The mechanism contains 105-step elementary reactions involving 25 species. All of the 25 species are included in the chemical kinetic model used. These species include all the reactants used, all the products formed, and the reaction intermediates formed, *i.e.*, the temporary products and/or reactants in the reaction steps of the mechanism used. An order of species has been accurately defined in this mechanism, based on hydrocarbon molecules, H/C/O molecules, hydrocarbon radicals, H/C/O radicals, and buffer gases, with each class ordered in terms of increasing complexity.<sup>64,65</sup> The surface reaction mechanism used here is that proposed by Deutschmann *et al.*<sup>66</sup> The mechanism contains 24-step elementary reactions involving 11 surface species and 9 gaseous species. Further information about this mechanism can be found on the DETCHEM website.<sup>67</sup>

For the reforming process, gas-phase reactions have been found to contribute only at relatively high temperatures (in excess of 950 °C).<sup>68</sup> Therefore, gas-phase reactions are negligible to reduce the computational complexity. The surface reactions involved in the methane steam reforming over rhodium are modeled using the mechanism proposed by Karakaya *et al.*<sup>69</sup> The mechanism can be easily implemented into a chemistry solver (*e.g.*, CHEMKIN<sup>70</sup> and Surface-CHEMKIN<sup>71</sup>). The Karakaya's mechanism contains 48-step elementary reactions involving 12 surface species and 6 gaseous species. The surface reactions involved in the methane steam reforming over nickel are modeled using the mechanism proposed by Delgado *et al.*<sup>72</sup> The Delgado's mechanism contains 52-step elementary reactions involving 14 surface species and 6 gaseous species. Further information about the Karakaya's and Delgado's mechanisms can be found on the DETCHEM website.<sup>67</sup> On the

other hand, the gas-phase reactions involved in the methane steam reforming over either rhodium or nickel are negligible, as discussed earlier.

Special attention should be paid to the catalyst loading used on each side of the reactor. The catalyst loading is expressed through the area factor,  $F_{cat/geo}$ , as defined in eqn (13). This parameter can be used to account for the influence of catalyst loading, because there exists a linear relationship between area factor and catalyst loading.<sup>58</sup> The area factor is also used to describe the dependence of the overall reaction rate on catalyst loading.<sup>58</sup> The area factor is set as 20 on each side of the reactor. In this context, complete conversion is possible on each side of the reactor under the conditions given Table 1, as will be discussed in detail later.

It is worth making a brief discussion about the issue related to the definition of the catalyst loading used in the model. It is too complicated to model the actual structure of a catalyst washcoat.<sup>58</sup> It is therefore necessary to simplify the washcoat domain of the system for the implementation of the model developed in this paper based on commercial computational fluid dynamics software, such as ANSYS Fluent. Both the area factor and the density of catalyst surface sites should be defined in order to account for the effect of the catalyst loading used. With the surface area factor, ANSYS Fluent can compute the total surface area on which the reaction takes place in each cell by multiplying this value by the volume of the cell. In this context, the evolution of chemical species and temperature in the two catalyst washcoats could be reasonably and accurately modeled with the model developed in this paper.

Thermochemical information is obtained from the provided kinetics schemes. Transport properties are obtained from the Sandia CHEMKIN transport database.<sup>57</sup> The gas-phase and surface reaction rates are handled with the CHEMKIN<sup>70</sup> and Surface-CHEMKIN,<sup>71</sup> respectively.

### 2.4. Computation scheme

The conservation equations must to be solved with appropriate boundary conditions. Inlet temperatures, pressures, velocities, and gas compositions are specified as the feed conditions. At each of the gas-washcoat interfaces, the no-slip condition is applied. Variations in physical properties of the mixture with composition and temperature are included in the model. To compare the modeling results with an independent model, the CEA (Chemical Equilibrium with Applications) software is used to generate the equilibrium data. This software is based on minimization of Gibbs free energy.<sup>70</sup>

The physical domain of the system must be transformed into a computational domain through a mesh. An orthogonal staggered mesh is used, and the node spacing is kept relatively small within the washcoat and near the entrance to the reactor. A typical mesh, totaling approximately 40 000 nodes, is used for the base case given in Table 1. In addition, a mesh consisting of 80 000 nodes in total is used in the case of largest dimension. For the base case, typical fluid node spacing is 250  $\mu\text{m}$  in the axial direction and 2.5  $\mu\text{m}$  in the transverse direction; typical wall node spacing is 250  $\mu\text{m}$  in the axial direction and 5.0  $\mu\text{m}$  in



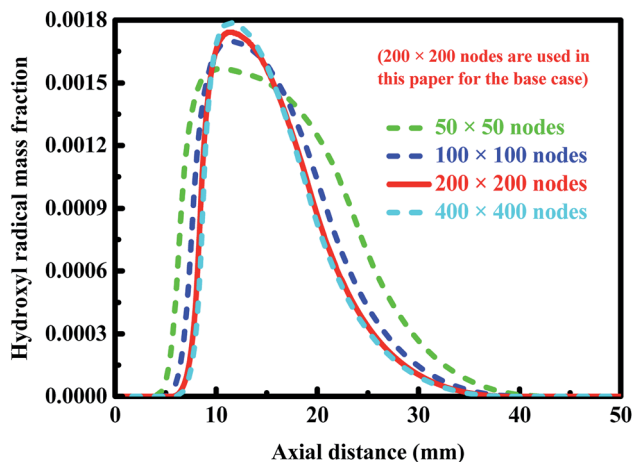


Fig. 2 Profiles of the hydroxyl radical concentration along the longitudinal axis of the combustor for some of the meshes used for the base case. The operating conditions used in the numerical simulations conducted here are given in Table 1.

the transverse direction; typical washcoat node spacing is 250  $\mu\text{m}$  in the axial direction and 1.0  $\mu\text{m}$  in the transverse direction. A mesh independence test is performed.

Fig. 2 shows the profiles of the hydroxyl radical concentration along the longitudinal axis of the combustor for some of the meshes used for the base case. Refer to Table 1 for the operating conditions used in the numerical simulations performed here. As the mesh density increases, a trend toward convergence can be found for the solution. The coarsest mesh, totaling 2500 nodes, cannot accurately describe the concentration of the hydroxyl radical appeared in the reaction region and its peak position. Therefore, the coarsest mesh fails to accurately describe the chemical reaction occurring in the combustion channel. In contrast, the solution obtained from a mesh consisting of tens of thousands of nodes used for the base case can be considered to be reasonably accurate. A larger mesh density, up to 160 000 nodes in total, offers no significant advantage.

The solution is deemed converged as the residuals of the conservation equations are less than  $10^{-6}$ . Due to the inherent stiffness of the detailed reaction mechanisms used, the convergence of the solution is often difficult. A summary of the operating conditions and design parameters used in the computational fluid dynamics simulations performed in this study is given in Table 2.

### 3. Results and discussion

#### 3.1. Analysis of reaction and transport time scales

The methane steam reforming reaction can proceed over rhodium at millisecond contact times due to enhanced transport characteristics and the high catalytic activity of the catalyst used.<sup>10,11</sup> Miniaturization of chemical processes results in enhanced transport rates and concomitant size reduction. On the other hand, the order of magnitude of intrinsic reaction and transport time scales are estimated to achieve intensification of

the reforming process. Note that the time scale defined here refers to an order of magnitude of time.

Fig. 3 shows the intrinsic reaction and transport time scales in the reforming channel and in the combustion channel. Note that a base-10 log scale is used for the vertical axis. The intrinsic reaction time scale refers to the time required for the concentration of the fuel to reduce to half its initial value

$$\tau_{\text{reaction}} = \frac{0.5C_{\text{fuel,in}}}{\left[ S_{\text{fuel}}(S/V') + \dot{\omega}_{\text{fuel}} \right]}, \quad (22)$$

where  $C_{\text{fuel,in}}$  is the concentration of the fuel at the inlet,  $S$  is the surface area of a channel, and  $V'$  is the volume of a channel. On the reforming side, gas-phase reactions are negligible, as discussed earlier. The intrinsic reaction time scales are estimated at the temperatures specified in Fig. 3. At temperature 1500 K, at which the surface reaction mechanisms<sup>69,72</sup> used in this study are thermodynamically inconsistent, the intrinsic reforming time scale is estimated by applying a polynomial extrapolation method. The time-scale of the mass transfer in the transverse direction refers to the time required for the transport of reactive species to gas-washcoat interfaces over half of the channel height by molecular diffusion

$$\tau_{y,\text{diffusion of species}} = \frac{d^2}{4D_{\text{fuel}}}, \quad (23)$$

where  $d$  is channel height, and  $D_{\text{fuel}}$  is the diffusivity for the fuel in the mixture.

Similarly, the time-scale of the heat transfer in the transverse direction can be expressed as

$$\tau_{y,\text{diffusion of heat}} = \frac{d^2}{4D^T} \quad \text{with } D^T = \frac{\lambda_g}{\rho c_p}, \quad (24)$$

in which  $D^T$  is the thermal diffusivity.

The mass and heat transfer in the axial direction is determined by the mean residence time

$$\tau_x \approx \frac{l}{\bar{u}}. \quad (25)$$

where  $l$  is the length of the reactor and  $\bar{u}$  is the mean flow velocity of a fluid.

Fig. 3(a) shows the intrinsic reaction and transport time scales in the reforming channel. When the temperature increases, the intrinsic reforming reaction time scale decreases from about 800 to 0.2 milliseconds in the case of rhodium, and from about 700 to 3 milliseconds in the case of nickel. The difference in intrinsic reforming reaction time scale between the two catalysts is minor at low temperatures, but significant at higher temperatures. Numerical simulations are carried out under the various feed-composition conditions, the results indicate that the ratio of the intrinsic reforming reaction time scale in the case of rhodium to that in the case of nickel can be up to 20 within the temperature range examined here. At a temperature of 1200 K and 50% conversion, representative operating conditions, the rate of the reforming reaction in the case of rhodium is about one order of magnitude faster than in the case of nickel.



Table 2 Operating conditions and design parameters used in the numerical simulations performed in this study

Conditions and parameters	Combustion side	Reforming side	Unit
<b>Reactor configuration</b>			
Channel length	50.0	50.0	mm
Channel height	0.2–0.8	0.2–0.8	mm
Wall thickness	0.2 mm	0.2	mm
Wall thermal conductivity	80.0 (300 K)		W (m K) <sup>-1</sup>
External heat loss coefficient	20		W (m <sup>2</sup> K) <sup>-1</sup>
Flow arrangement	Co-current flow and counter-current flow		—
<b>Catalyst washcoat</b>			
Catalyst	Pt/Al <sub>2</sub> O <sub>3</sub>	Rh/Al <sub>2</sub> O <sub>3</sub> and Ni/Al <sub>2</sub> O <sub>3</sub>	
Ratio of catalyst loadings	0.5–2.0	0.5–2.0	—
Washcoat thickness	0.08	0.08	mm
Catalyst/geometric surface area	20	20	—
Washcoat mean pore diameter	20	20	nm
Density of catalyst surface sites	2.72 × 10 <sup>-9</sup> (platinum)	2.72 × 10 <sup>-9</sup> (rhodium) 2.66 × 10 <sup>-9</sup> (nickel)	mol cm <sup>-2</sup>
Washcoat porosity	0.5	0.6	—
Washcoat tortuosity factor	3	3	—
<b>Operating conditions</b>			
Reaction temperature (time-scale analysis)	900, 1200, 1500	900, 1200, 1500	K
Inlet composition	0.8 (equivalence ratio)	3.0 (molar steam-to-carbon ratio)	—
Inlet temperature	300	400	K
Inlet pressure	0.1	0.1	MPa
Inlet velocity	0.3–6.0	0.2–6.0	m s <sup>-1</sup>
Wall temperature threshold	1500	1500	K
<b>Kinetic mechanisms</b>			
Gas-phase combustion	Hughes <i>et al.</i> <sup>64</sup> and Turányi <i>et al.</i> <sup>65</sup>	—	—
Catalytic combustion	Deuschmann <i>et al.</i> <sup>66</sup>	—	—
Steam reforming	—	Karakaya <i>et al.</i> <sup>69</sup> (rhodium)	—
Steam reforming	—	Delgado <i>et al.</i> <sup>72</sup> (nickel)	—

On the other hand, the transfer time scale in the reforming channel is on the order of tens of microseconds in the transverse direction, and on the order of a few milliseconds in the axial direction, as shown in Fig. 3(a). The intrinsic reforming reaction time scale is about one order of magnitude shorter than the residence time at temperature 1500 K in the case of rhodium, but become longer than or comparable to the residence time at temperatures 900 and 1200 K in the case of both rhodium and nickel. For the latter, possible or even considerable incomplete conversion occurs. Overall, the intensification of the reforming process is achieved in terms of transport, but the process can benefit from the improvement of the reforming catalyst.

Fig. 3(b) shows the intrinsic reaction and transport time scales in the combustion channel. As the temperature increases, the intrinsic combustion time scale decreases from about 0.005 to 0.5 milliseconds. As expected, the intrinsic combustion time scale is much shorter than the intrinsic reforming time scale, especially at low temperatures. The difference in time scale between intrinsic combustion and transverse transport is not significant at temperature 1200 K, but is up to one order of magnitude at temperatures 900 and 1500 K, as shown in Fig. 3(b). Therefore, the intensification of

the combustion process is necessary in terms of transport. The improvement of the combustion catalyst is needed at low temperatures, as shown in Fig. 3(b). The residence time is much longer than all the other time scales, and thus complete conversion is always possible on the combustion side.

The rapid production of hydrogen from methane steam reforming in a microchannel reactor poses a technical challenge to the development of novel engineered catalysts.<sup>9</sup> Novel engineered steam reforming catalysts play a strong role in increasing the productivity per unit volume. The enhanced heat and mass transfer of novel engineered catalysts may lead to improved efficiency and performance of methane steam reforming.<sup>73</sup> To fully take the heat and mass transfer advantages of microchannel reaction technology so that rapid production of hydrogen and a greater volumetric productivity can be achieved, it is necessary to develop highly active and stable methane steam reforming catalysts. Commercial methane steam reforming catalysts are currently based on nickel supported on refractory materials doped with a variety of promoters.<sup>74</sup> Nickel is preferable to the expensive rhodium from the perspective of cost. Significant improvements have been realized in increasing the performance of the catalysts used for methane steam reforming.<sup>74</sup>





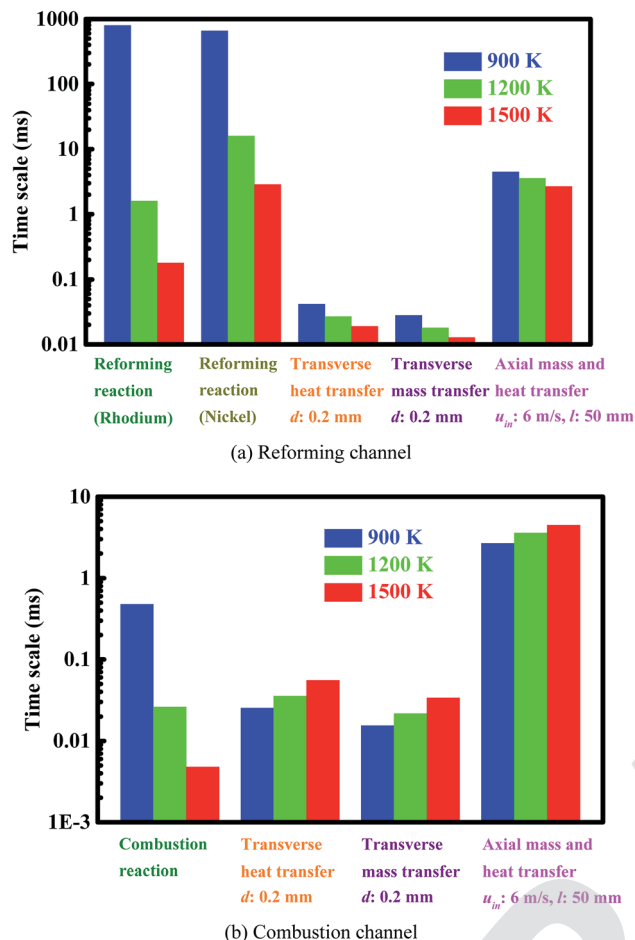


Fig. 3 Intrinsic reaction and transport time scales (Panel (a)) in the reforming channel and (Panel (b)) in the combustion channel. Note that a base-10 log scale is used for the vertical axis.

Methane steam reforming over nickel at millisecond contact times is in principle feasible, provided that the catalyst used is properly designed with the aid of a tailored microchannel reactor design.<sup>24</sup> Further reductions in contact time may be reasonably achieved by increasing the catalyst thickness in a manner that minimizes heat and mass transport limitations through careful design. The development of lower-cost, high-activity nickel-based catalysts is highly desirable, yet challenging.<sup>74</sup> Methane steam reforming catalysts containing more than one active species are being investigated, and the performance of some of these novel catalyst formulations has been evaluated by Turchetti *et al.*<sup>75</sup> In addition, kinetics of methane steam reforming over different catalyst-support combinations have also been studied, providing a tool for the design of steam methane reformers.<sup>75</sup> Recently, the effect of the promoter in nickel-based catalysts has been investigated for steam methane reforming, and it has been found that the promoter plays an important role in determining the performance of these catalysts.<sup>75,76</sup> The sintering of supports and catalysts is generally unavoidable and is a very complicated process which is accelerated at higher temperatures. On the other hand, membrane reactors show great promise for the rapid production of

hydrogen from methane steam reforming.<sup>75,76</sup> Methane steam reforming over nickel may be allowed to be carried out at millisecond contact times through intensifying the process by adopting suitable strategies, such as a high-performance catalyst and a tailored membrane reactor design.

### 3.2. Coupling of exothermic and endothermic reactions

The effect of thermal coupling on the reactor performance is investigated, since the management of heat is very important for reactor design.<sup>52,77</sup> Fig. 4(a) shows the axial temperature and conversion profiles on both sides of the reactor. The wall temperature first increases and then decreases slowly along the longitudinal axis of the reactor. The outlet conversion is complete on the combustion side, whereas an outlet conversion of 98.8% is obtained on the reforming side. Fig. 4(b) shows the axial heat flux profiles on both sides of the reactor. The sub-peak appeared on the generated heat flux profile indicates the onset of gas-phase combustion, providing another route for heat generation. Catalytic combustion can provide the amount of heat required to raise the temperature of both the streams

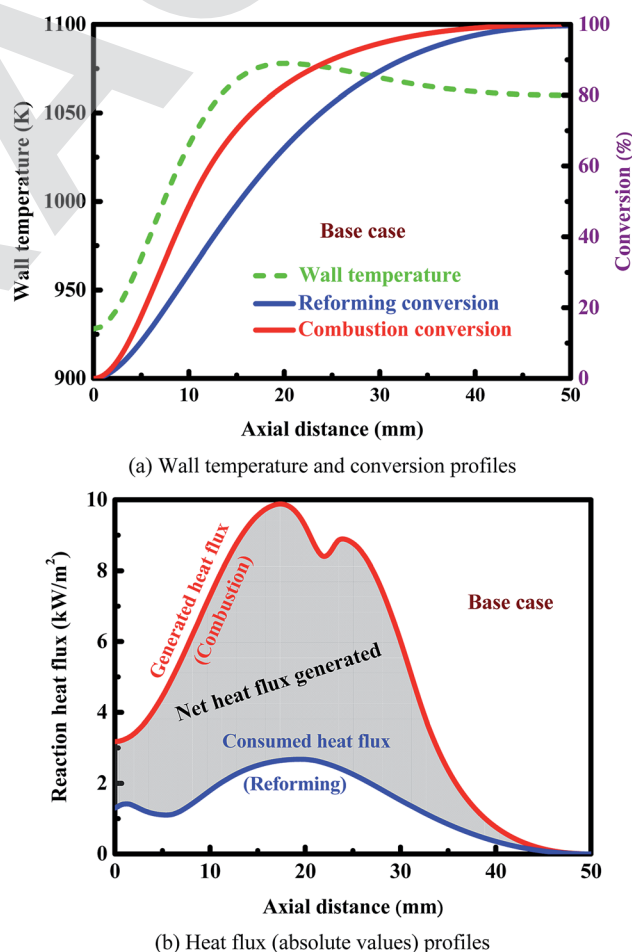


Fig. 4 (Panel (a)) Wall temperature and conversions in both channels, and (Panel (b)) reaction heat fluxes along the longitudinal axis of the reactor. The shaded region in Panel (b) represents the net amount of heat generated within the system.



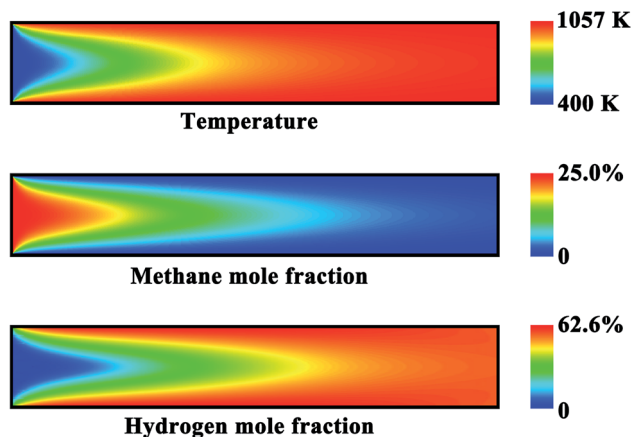


Fig. 5 Contour plots of the temperature and methane and hydrogen mole fractions in the reforming channel for the base case.

and to drive the reforming reaction. The net amount of heat generated is represented by the shaded region in Fig. 4(b). The axial position of the peak appeared on the generated heat flux profile is almost the same as that appeared on the consumed heat flux profile. The rate of the combustion reaction is much faster than that of the reforming reaction, and thus the amount of the heat released from combustion is much greater than that of the heat consumption of the reforming reaction. As a result, both of the two fluids are heated in the first half of the reactor, as illustrated in Fig. 4(a). After the middle of the reactor, there is a very small difference in temperature between the two fluids, and thus no significant heat transfer will be achieved along the remaining length of the reactor.

Fig. 5 shows the contour plots of the temperature and methane and hydrogen mole fractions in the reforming channel for the base case. While transverse temperature differences are negligible in the solid phase, there are steep species and temperature gradients in the gas phase near the reaction region, as shown in Fig. 5. In this context, a two-dimensional model is necessary to understand the transport characteristics involved in the system, because diffusion of both energy and species is not negligible in the axial direction. Overall, a good thermal balance is achieved between the two fluids under the conditions examined for the base case.

### 3.3. Effect of catalyst loading

Numerical simulations are performed at various catalyst loadings, by means of adjusting  $F_{\text{cat}/\text{geo}}$ . This parameter can be thought of as a measure of catalyst loading.<sup>58</sup> The ratio of catalyst loadings, which is used to investigate the influence of catalyst loading, is defined as

$$K = \frac{F_{\text{cat}/\text{geo}}}{(F_{\text{cat}/\text{geo}})_{\text{base case}}} \quad (26)$$

Fig. 6 shows the wall temperature profiles along the longitudinal axis of the reactor and the outlet conversions on both sides of the reactor under different catalyst-loading conditions.

Fig. 6(a) and (b) show the results obtained on the combustion side. The rate of the combustion reaction decreases with decreasing the loading of the combustion catalyst. Therefore, a smaller amount of combustion catalyst results in lower wall temperatures (Fig. 6(a)) and a lower outlet conversion on each side of the reactor (Fig. 6(b)). The smallest amount of combustion catalyst (*i.e.*,  $K_{\text{combustion}} = 0.5$ ) causes a cold spot to be formed within the wall at the axial position about 8 mm (the blue line in Fig. 6(a)). In contrast, the largest amount of combustion catalyst (*i.e.*,  $K_{\text{combustion}} = 2.0$ ) causes a hot spot to be formed within the wall at the axial position about 12.6 mm (the red line in Fig. 6(a)). This is a situation that needs immediate attention, since hot spots may lead to rapid deactivation of the catalysts used on each side of the reactor due to high temperatures.

Fig. 6(c) and (d) show the results obtained on the reforming side. A larger amount of reforming catalyst does not improve the performance of the reactor. The wall temperature decreases with increasing the loading of the reforming catalyst (Fig. 6(c)). In certain cases (*e.g.*,  $K_{\text{reforming}} = 2.0$ ), a cold spot is formed due to the cooling effect caused by a large amount of reforming catalyst (the red line in Fig. 6(c)). Overall, a larger amount of reforming catalyst results in lower wall temperatures (Fig. 6(c)) and a lower outlet conversion on each side of the reactor (Fig. 6(d)). In contrast, a smaller amount of reforming catalyst results in higher wall temperatures (Fig. 6(c)) and a higher outlet conversion on the combustion side (Fig. 6(d)), but a lower outlet conversion on the reforming side due to the lack of enough amount of reforming catalyst (Fig. 6(d)).

Fig. 7 shows the axial heat flux profiles obtained for two representative cases. In the case of  $K_{\text{reforming}} = 2.0$  (Fig. 7(a)), the heat flux generated is smaller than that consumed near the entrance to the reactor, thus decreasing the wall temperature. At a certain axial position (about 8 mm), the heat flux generated becomes larger than that consumed, thus increasing the wall temperature. As a result, a cold spot is formed, as shown in Fig. 7(a) as well as Fig. 6(c) (the red line). In the case of  $K_{\text{reforming}} = 0.5$  (Fig. 7(b)), the opposite occurs. The heat flux generated is much larger than that consumed near the entrance to the reactor, thus increasing the wall temperature. In the first half of the reactor, a large amount of heat released by catalytic combustion is transferred to the reforming side, and thus a peak appears on each of the two heat flux profiles. As a result, a hot spot is formed at axial position about 15.6 mm, as shown in Fig. 7(b) as well as Fig. 6(c) (the blue line). In this case, the catalyst loadings must be carefully designed to avoid deactivation of the catalyst on each side of the reactor.

### 3.4. Effect of reactor dimension

The reactor dimension should be properly designed, as large channels exacerbate diffusional resistance. Numerical simulation are performed at different channel heights to understand the role of reactor dimension. Additionally, several dimensionless numbers are used to characterize the relative strengths of the different transport phenomena. Some dimensionless numbers are shown in Fig. 8(a) for different reformer



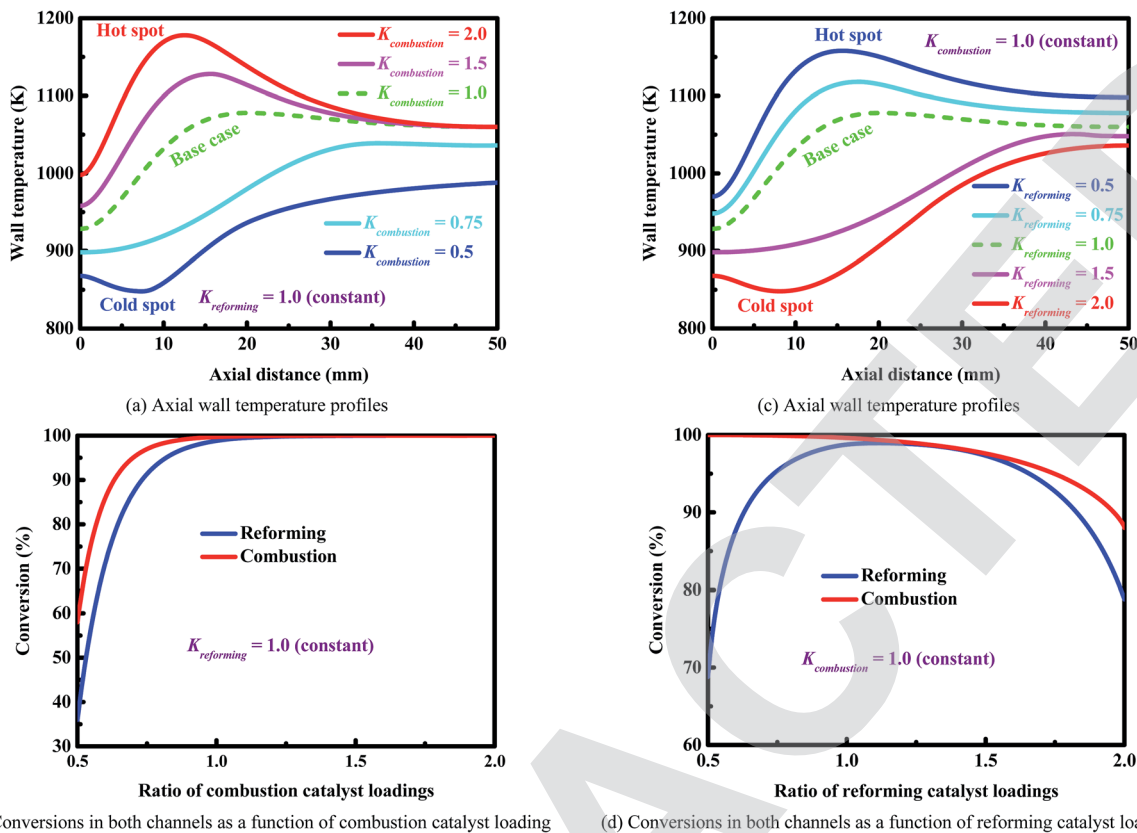


Fig. 6 Wall temperature profiles and conversions (Panels (a) and (b)) on the combustion side and (Panels (c) and (d)) on the reforming side.

dimensions at constant inlet velocities. The transverse Damköhler number is defined as the ratio of the time scale of transverse mass transfer to intrinsic reaction

$$Da_y = \frac{\tau_{y,\text{diffusion of species}}}{\tau_{\text{reaction}}} = \frac{d^2 s_{\text{fuel}}(S/V)}{2D_{\text{fuel}} C_{\text{fuel,in}}}. \quad (27)$$

The Péclet number, which characterizes the strength of convection relative to molecular diffusion, is defined as the ratio of the time scale of mass transfer in the transverse direction to residence time

$$Pe = \frac{\tau_{y,\text{diffusion of species}}}{\tau_x} = \frac{d^2 \bar{u}}{4D_{\text{fuel}} l}. \quad (28)$$

The Fourier number is defined as the ratio of residence time to the time scale of heat transfer in the transverse direction

$$Fo = \frac{\tau_x}{\tau_{y,\text{diffusion of heat}}} = \frac{4\lambda_g l}{d^2 \bar{u} \rho c_p}. \quad (29)$$

The three dimensionless numbers defined above give the relative strengths of the different phenomena of heat and mass transport. The thermodynamic properties of the mixture depend on the local temperature and composition, and they are computed at a point located at 10.0 mm downstream of the entrance for the sake of simplicity. This location is sufficiently

downstream to allow preheating of the mixture, but not too far downstream where the reforming reaction has proceeded to completion, as illustrated in Fig. 5. The transverse Damköhler number is not sensitive to the reformer dimension, as shown in Fig. 8(a), since the reforming reaction is under kinetic control, as illustrated in Fig. 3(a). Fig. 8(a) also quantifies the relative importance of the transport time scales in the transverse direction and in the axial direction (red and blue lines). Under the conditions examined here, all of the Fourier numbers are greater than 10, whereas all of the Péclet numbers are strictly less than 0.1. Therefore, the time scales for both heat and mass transport in the transverse direction are about one order of magnitude shorter than those in the axial direction. On the other hand, the transverse transfer time scales are quite shorter than the intrinsic reforming reaction time scale (refer to Fig. 3(a) for more details). Overall, the above results, based on an analysis of the time scales, indicate that the intensification of the steam reforming process has been well implemented in terms of interphase heat and mass transfer in the reforming channel.

Fig. 8(b) shows the effect of channel height on the maximum wall temperature and conversion at constant inlet velocities. The dimension of the reactor plays an important role in determining the performance of the system in terms of both conversion and temperature. As the dimension of the reformer increases, three effects result in a decrease in the conversion on the reforming side: (a) the resistance to mass transfer increases



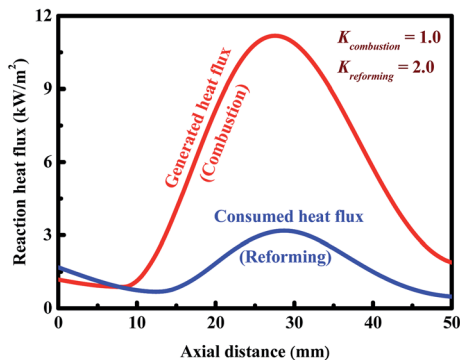
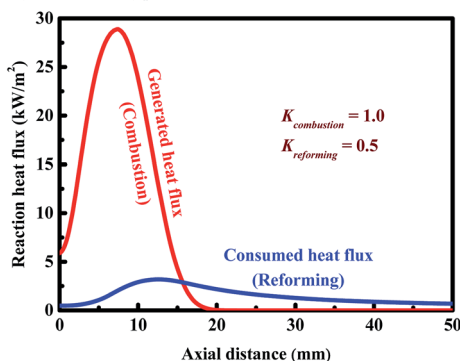
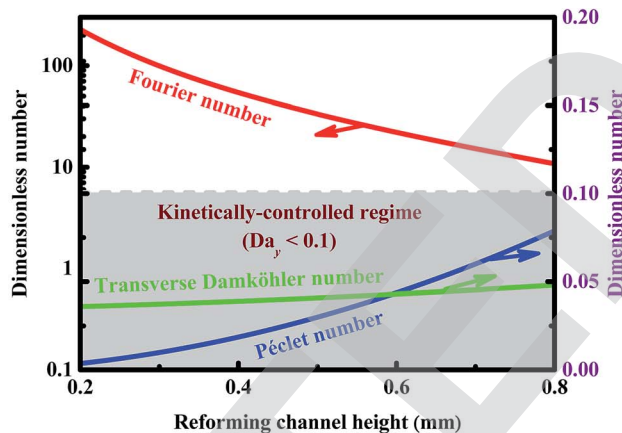
(a) Heat flux (absolute values) profiles in the case of  $K_{\text{combustion}} = 1.0$  and  $K_{\text{reforming}} = 2.0$ (b) Heat flux (absolute values) profiles in the case of  $K_{\text{combustion}} = 1.0$  and  $K_{\text{reforming}} = 0.5$ 

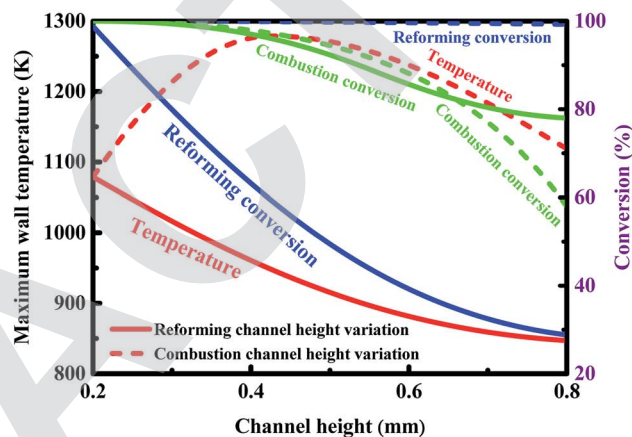
Fig. 7 Reaction heat fluxes along the longitudinal axis of the reactor. Panel (a): a cold spot is formed in the reactor. Panel (b): a hot spot is formed in the reactor.

in the transverse direction, (b) the total amount of the reforming catalyst used is relatively insufficient, and (c) there is a larger quantity of the reactants required to be heated on the reforming side. In addition, the maximum wall temperature decreases with increasing the dimension of the reformer. On the other hand, at all of the combustor dimensions examined here, complete conversion on the reforming side can always be achieved (Fig. 8(b)). The reactor performance is determined by both the energy input and the external resistance to mass transfer. As dimension of the combustor increases, the conversion on the combustion side decreases due primarily to the enhanced external resistance to mass transfer. In contrast, the temperature increases first and decreases later (Fig. 8(b)). In the case of smaller combustors, conversion is almost complete on each side of the reactor; in addition, the temperature increases with increasing the dimension of the combustor due primarily to the increased quantity of the reactants. In the case of larger combustors, the conversion is incomplete on the combustion side, and the temperature decreases with increasing the dimension of the combustor due primarily to the enhanced external resistance to mass transfer.

Overall, the reactor dimension can significantly affect the performance of the system in terms of both conversion and temperature for a given inlet velocity. The reactor dimension needs to be properly designed to improve the efficiency of heat exchange.



(a) Transverse Damköhler, Fourier, and Péclet numbers



(b) Maximum wall temperature and conversion

Fig. 8 Effect of channel height on (Panel (a)) the Péclet, Fourier, and transverse Damköhler numbers as well as (Panel (b)) the maximum wall temperature and conversion. The solid and dashed lines reflect the variation of reformer dimension and combustor dimension, respectively. The inlet flow velocities of the reforming stream and the combustible stream are kept constant at 2.0 and 6.0  $\text{m s}^{-1}$ , respectively. In order to obtain the above dimensionless numbers, the thermodynamic properties of the mixture are computed at a point located at 10.0 mm downstream of the entrance for the sake of simplicity.

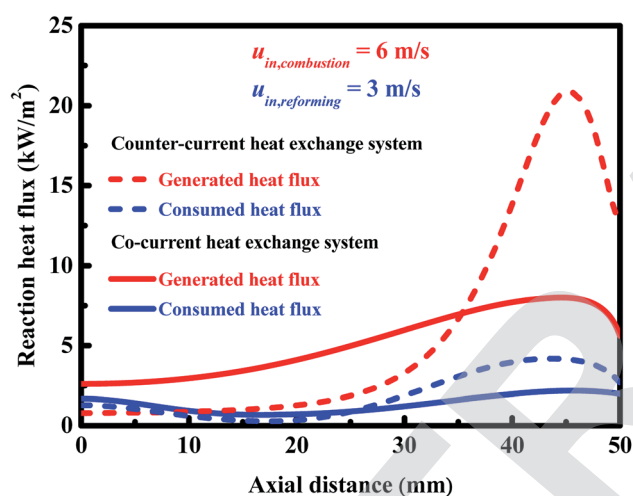
### 3.5. Effect of flow arrangement

The arrangement of the endothermic and exothermic flows can greatly affect the internal heat exchange and thus the reactor performance.<sup>78</sup> Counter-current heat exchange systems, where the hot fluid and the cold fluid flow in opposite directions to each other, offer many advantages over co-current (*i.e.*, parallel) heat exchange systems, where the two fluids flow in the same direction.<sup>79</sup> However, an efficient operation requires not only a good thermal coupling, but also high reliability. The effect of flow arrangement on the reactor performance is investigated. Comparisons are made in terms of reactor performance between the two heat exchange systems. Since a high production rate of hydrogen is highly desirable in practice,<sup>10–12</sup> high reforming stream flow rates are considered here.

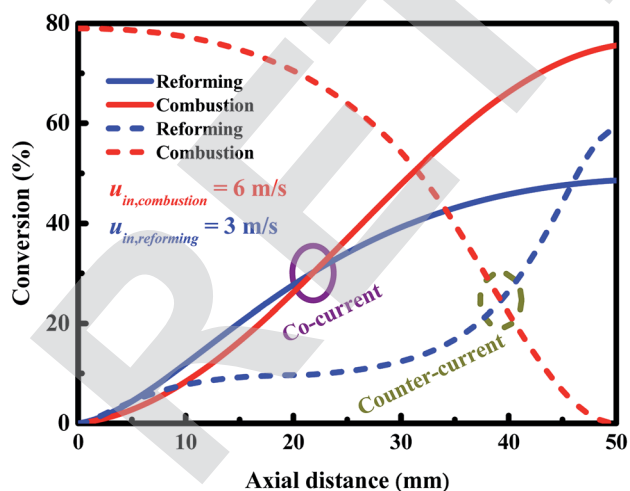
Both arrangements of the endothermic and exothermic flows in the alternate channels have been illustrated in Fig. 1. A



comparison is made based on the axial heat flux and conversion profiles (Fig. 9(a) and (b)). Fig. 9(a) shows the axial heat flux profiles obtained for both heat exchange systems. The axial heat flux profiles are more uniform for the co-current heat exchange system, but more concentrated towards the reformer outlet for the counter-current heat exchange system. The ratio of the amount of heat generated to heat consumed is 0.6 at the reformer entrance for the counter-current heat exchange system and 1.6 for the co-current heat exchange system. Accordingly, the cooling effect, as described early, becomes more pronounced and a cold spot is formed for the counter-current heat exchange system. The ratio of the amount of heat generated to heat consumed is 4.8 at the reformer outlet for the counter-current heat exchange system and 2.8 for the co-current heat exchange system, eventually leading to localized over-heating for each of the two systems.



(a) Heat flux (absolute values) profiles



(b) Conversion profiles

Fig. 9 (Panel (a)) Reaction heat fluxes and (Panel (b)) conversions along the longitudinal axis of the reactor. The solid and dashed lines refer to the co-current heat exchange system and the counter-current heat exchange system, respectively. The reactor is better balanced thermally by implementing co-current heat exchange.

Fig. 9(b) shows the axial conversion profiles for the two flow arrangements. The counter-current heat exchange system has a higher conversion at the combustor outlet than the co-current heat exchange system. For the counter-current heat exchange system, most of the heat generated by catalytic combustion is released near the combustor entrance (*i.e.*, near the reformer outlet; refer to Fig. 1 for more details), and subsequently transferred to the reforming side. Therefore, special attention should be given to the thermal imbalance within the counter-current heat exchange system, and thermal management is always necessary to improve the reliability of the system. Furthermore, the thermal imbalance greatly increases the occurrence probability of gas-phase combustion, the counter-current heat exchange system must be carefully designed to avoid thermal runaway failures. On the other hand, the counter-current heat exchange system has a higher conversion at the reformer outlet than the co-current heat exchange system.

Comparisons are made between the two heat exchange systems in terms of multiple performance criteria, and the results are summarized in Table 3. A good thermal balance is achieved within the co-current heat exchange system, which can result in hot spot elimination and thus is a potential advantage. Better utilization of the overall heat generated by catalytic combustion is possible for the counter-current heat exchange system, making it possible to improve the conversion on each side of the reactor. The counter-current heat exchange system will cause a larger temperature gradient over the length of the reactor, and in the transverse direction. Therefore, thermal management is always necessary to avoid thermal runaway failures and reactor extinguishing. For the counter-current heat exchange system, localized insufficient heat exchange may cause hot spots or cold spots to form. The occurrence of hot spots imposes more severe constraints on the wall materials and catalysts used in the manufacture of microchannel reactors; cold spots may cause reactor extinguishing.

### 3.6. Engineering maps and operating windows

The microchannel reactor should be properly designed to meet the requirements of the rapid production of hydrogen. In light of this, engineering maps are presented here, and operating windows are also identified. Fig. 10(a) gives an operation diagram representing the power output at different residence times of the combustible stream, assuming full utilization of the hydrogen produced from methane steam reforming. The shaded region in Fig. 10(a) depicts the operation window delimited by two types of the limits indicated. Fig. 10(a) shows that a balance between the two flow rates is required to ensure reliable operation of the device. The reactor can be operated effectively within a specified power range which varies based on the flow rate of combustible stream, ranging from the minimum power output to the maximum power output, as described below.

The minimum power output is determined by the stability limits of the materials and catalysts used.<sup>10–12</sup> Since the reactor generates heat during operation, thermal management is required to ensure that the device is operating below the



Table 3 Comparison between the co-current heat exchange system and the counter-current heat exchange system

	Co-current	Counter-current
Inlet flow velocity of the reforming stream	3.0 m s <sup>-1</sup>	
Inlet flow velocity of the combustible stream	6.0 m s <sup>-1</sup>	
Outlet conversion on the reforming side	48.6	58.7
Outlet conversion on the combustion side	75.6	79.0
Maximum wall temperature	1047 K	1160 K
Minimum wall temperature	870 K	848 K
Transverse temperature difference in the gas phase		
Maximum	570 K	687 K
Minimum	-66 K	-180 K
Ratio between overall generated and consumed heat fluxes	3.78	3.26
Ratio between local generated and consumed heat fluxes		
At the reformer outlet	2.8	4.8
At the reformer entrance	1.6	0.6

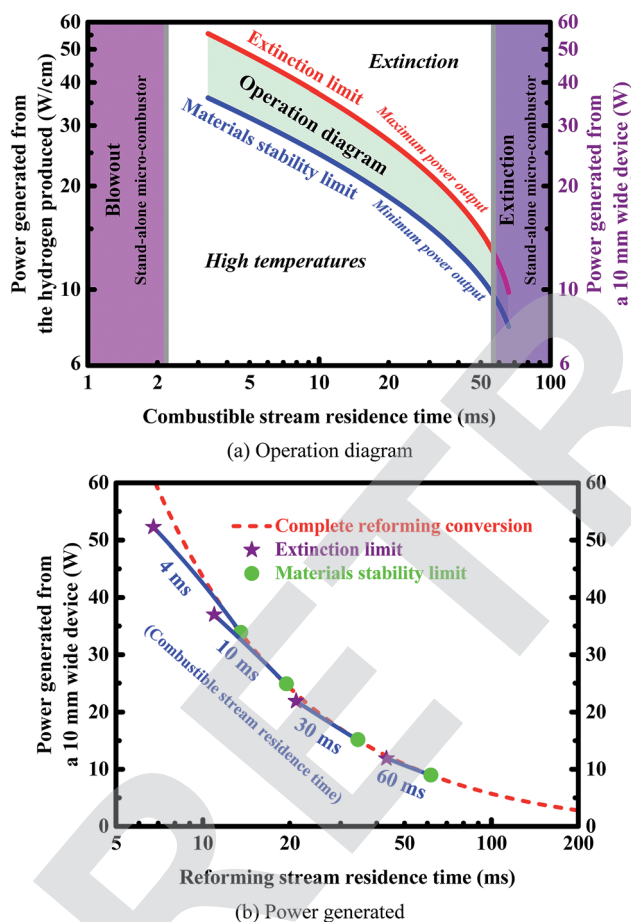


Fig. 10 Panel (a) shows an operation diagram representing the power output at different residence times of the combustible stream, assuming full utilization of the hydrogen produced from methane steam reforming. The gray vertical lines represent the lower and upper bounds determined by the critical flow velocities in a stand-alone micro-combustor. Panel (b) shows the power generated at different residence times of the reforming stream, assuming full utilization of the hydrogen produced from methane steam reforming in a reactor with a width of 10.0 mm. The "circle" and "pentagram" symbols indicate the stability limits of combustion and materials, respectively.

maximum operating temperature allowed. The heat generated by the catalytic combustion leads to a significant increase in catalyst temperature. Therefore, the stability of catalysts at high temperatures is of considerable interest. It is possible to design reactors in which efficient heat transfer is used to minimize temperature rise but particular attention must be paid in all cases to the temperature stability of the catalysts and materials used. In the context of high heat fluxes, the wall temperatures of an integrated system should not exceed a certain threshold in order to avoid deactivation of the catalyst on each side of the reactor. The threshold temperature (*i.e.*, peak operating temperature or maximum normal operating temperature) of the reactor is set as 1500 K,<sup>10–12</sup> beyond which the device may no longer function.

On the other hand, the maximum power output is determined by the maximum amount of hydrogen that can be produced before extinction occurs when the flow rate of reforming stream is sufficiently high. Outside this range of power output, the reactor will no longer function when the flow rate of combustible stream is kept constant. However, higher power output can be achieved by connecting multiple reactor units in parallel.

The reactor can also be operated effectively within a specified range of combustible stream residence-time, which is determined by the stability limits of combustion, *i.e.*, both extinction and blowout.<sup>80</sup> Extinction occurs due to a lack of heat generated, whereas blowout occurs due to a lack of residence time. The critical flow velocities in a stand-alone micro-combustion system are determined on the basis of a methane–air mixture in which the inlet composition is given in Table 1 (*i.e.*, inlet equivalence ratio 0.8). The resulting prohibited window where combustion cannot be self-sustained is illustrated in Fig. 10(a) as the vertical shaded regions to ensure that the residence time of the combustible stream lies within well-defined lower and upper bounds determined by the critical flow velocities in a stand-alone micro-combustor.

The power output as a function of the flow rate of the reforming stream is investigated at various residence-times of the combustible stream by using a two-parameter continuation. Fig. 10(b) shows the power generated at different residence



times of the reforming stream, assuming full utilization of the hydrogen produced from methane steam reforming in a reactor with a width of 10.0 mm. The “circle” and “pentagram” symbols indicate the stability limits of combustion and materials, respectively. Appropriate adjustments should be made to the flow rates on both sides of the reactor in order to ensure high outlet conversions, as shown in Fig. 10(b). The flow rates need to be carefully designed to meet the requirements of the stability of both combustion and materials.

The engineering maps presented in Fig. 10 can serve as a design tool for improving and optimizing the performance of an integrated micro-chemical system. Based on the operation regime presented in Fig. 10(a), an estimation about the range of the residence time of the combustible stream can be made for desired power output. Based on the power output plotted in Fig. 10(b), an estimation about the range of the residence time of the reforming stream can finally be made. Unfortunately, the range of each of the residence times is rather narrow, and thus the flow rates need to be carefully designed, as discussed above. The middle range of the residence time allowed for the combustible stream (Fig. 10(a)) and the lower bound of the residence time allowed for the reforming stream (Fig. 10(b)) will ensure higher fuel utilization and nearly complete conversion on the reforming side. This holds great promise for applications in fuel cells.

## 4. Conclusions

Methane steam reforming coupled with catalytic combustion in microchannel reactors was investigated, with the aim of achieving the rapid production of hydrogen. In order to optimize the performance of the small scale system, the effect of four typical variables of interest was investigated. The main points can be summarized as follows:

- The rapid production of hydrogen from methane steam reforming is feasible, provided that the reforming catalyst, flow rate, and reactor dimension are properly designed.
- Miniaturization of the chemical system is insufficient for process intensification. An efficient reforming catalyst is also necessary. Overall, miniaturization of the chemical system and the improvement of the reforming catalyst must be symbiotic.
- The reactor dimension needs to be properly designed to achieve high transport rates for the intensification of the steam reforming process.
- The catalyst loading is a key factor determining reactor performance, and must be carefully designed to avoid hot spots within the wall or low conversions on both sides of the reactor.
- Counter-current operation is not desirable due to the greater extremes of temperature, despite the fact that slightly better performance can be achieved. In contrast, co-current operation is recommended, as the reactor is better balanced thermally.
- A balance between the flow rates on both sides of the reactor is required to ensure reliable operation of the system. The flow rates need to be carefully designed to meet the requirements of the stability of both materials and combustion.

• From a practical point of view, the costs of the noble metal catalysts used here might be tolerable for microfabricated chemical systems. However, it is highly desirable to develop high-activity catalysts with lower cost.

The results may prove useful to those involved in the design, optimization and modification of the methane steam reforming process for the small-scale production of hydrogen and other related processes. While microchannel reactors offer many advantages for chemical production and process development, further research efforts are required to realize this micro-reaction technology. Special attention should be paid to the problem of catalyst deactivation. In addition, thermal management is always necessary to improve the reliability of these microfabricated chemical systems. This potential solution is currently being investigated. Furthermore, novel design methodologies are clearly needed to realize excellent thermal uniformity, high transport rates, and low pressure drop, and thus further work in design optimization is needed.

## Conflicts of interest

There are no conflicts to declare.

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