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Numerical study of on-board fuel reforming in a catalytic plate reactor for solid-oxide fuel cells

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ARTICLE INFO

Article history:

Received 17 August 2010

Received in revised form

11 November 2010

Accepted 11 November 2010

Available online 20 November 2010

Keywords:

Auxiliary power units

Catalysis

Catalytic plate reactors

Reforming

Partial oxidation

Numerical modeling

ABSTRACT

A pseudo-transient numerical model is used for the simulation of a multi-functional catalytic plate reactor (CPR). The work mainly addresses the problems associated with on-board reforming for solid-oxide fuel cells. Heat management is achieved by indirectly coupling partial oxidation with reforming. Water management is achieved by partially recycling the anode stream from a solid-oxide fuel cell. The model uses detailed heterogeneous chemistry for reforming and oxidation reactions occurring on the catalyst beds.

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1. Introduction

Natural gas is widely used for transportation and worldwide there are 11.2 million vehicles running on natural gas (IANGV, 2010). In this paper we address the on-board reforming of natural gas for SOFC; however, except the reaction kinetics the modeling framework will remain the same for other transport fuels as well. Steam reforming can turn out to be counter productive due to added system complexity associated with on-board water management and the need for an additional reactant stream (Lindermeier et al., 2007). However, once the fuel cell is in operation auto-thermal reforming can be achieved by recycling the anode stream.

Steam reforming of natural gas is an important chemical process for the production of syngas (H_2 and CO) (Schädel et al., 2009; Bharadwaj and Schmidt, 1995), which is an ideal fuel for SOFC. Another important chemical route for the conversion of hydrocarbons to syngas is partial oxidation. Since CH_4 is the major constituent of natural gas we consider CH_4 as the fuel in this study.

There are many studies concerning the reforming and oxidation of CH_4 to syngas (Clarke et al., 1997; Schwiedernoch et al., 2003; Xu and Froment, 1989a; Bharadwaj and Schmidt, 1995). Higher selectivity and better heat and mass transfer rates can be achieved if reforming or oxidation is carried out in micro-structured reactors

(Kolb and Hessel, 2004; Kiwi-Minsker and Renken, 2005). The enhanced heat transfer rate in micro-structured reactors can be exploited in catalytic plate reactors (CPR), where exothermic reactions are combined with endothermic reactions in a single reactor designed like a heat exchanger (Vlachos and Caratzoulas, 2010). In this set-up the exothermic and endothermic reactions are separated by a thin wall as shown in Fig. 1. The system may or may not have porous catalyst coating (Kolb and Hessel, 2004) and different catalysts may be applied on either side of the wall depending on the nature of reactions desired (Vlachos and Caratzoulas, 2010). For example, Ni or Rh can be a good catalyst for reforming on the other hand Pt can be used for oxidation. In CPR the coupling of endothermic and exothermic reactions are accomplished by means of indirect heat transfer.

There are many studies on the modeling of CPR for various applications. Zanfir and Gavriilidis carried out parametric studies related to channel height and catalyst layer thickness by coupling CH_4 reforming and CH_4 oxidation in a CPR. Their model equations resolved the flow channels and the solid wall two dimensionally and the catalyst bed one dimensionally under steady state conditions (Zanfir and Gavriilidis, 2003). Keeping the bed thickness to flow velocity ratio constant, they found that increasing bed thickness resulted in lower conversions. Using the same mathematical model they also analyzed the influence of counter current and co-current arrangement of CPR for steam reforming of methane. They used the global reforming kinetics of CH_4 on Ni reported by Xu and Froment (1989b) in the reforming channel, whereas a first-order kinetics is used to represent the catalytic combustion of CH_4 in the oxidation channel (Zafir and

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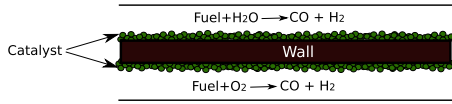


Fig. 1. Schematic representation of a catalytic plate reactor. Fuel combustion takes place in the bottom channel and reforming takes place in the top channel.

Gavriilidis, 2004). Based on this study they concluded that co-flow configuration can avoid hot spots which are observed in the counter flow configuration. Nevertheless, they found the conversion to be higher for counter flow configuration.

In another study Baldea and Daoutidis (2007) carried out transient analysis of CPR. Similar to Zanfir and Gavriilidis they also used a global kinetic formulation for CH_4 reforming reported by Xu and Froment (1989b). For oxidation they have considered water–gas shift reaction and the kinetic expressions are again taken from Xu and Froment (1989b). They found that certain conditions such as increasing the flow velocity resulted in reactor extinction. There are several other reported studies that utilize CPR. For instance, Kim et al. proposed a mathematical model for integrated reaction and heat exchanger model for micro-reactor based water–gas shift process. They carried out a parametric study on the CO conversion (Kim et al., 2005). Deshmukh and Vlachos (2005) modeled propane combustion and ammonia oxidation in a CPR using CFD. Stefanidis et al. (2009) and Stefanidis and Vlachos (2008) studied the coupling of CH_4 steam reforming with propane combustion in a CPR using a pseudo-two dimensional model.

The present work numerically studies the feasibility of on-board fuel processing by combining partial oxidation and steam reforming in a catalytic plate reactor. The work differs from the above-reported ones in various aspects. All the above-reported work implements global reaction kinetics for reforming and oxidation, whereas in this work we implement detailed kinetic models for the surface reactions. The porous media transport is implemented using Dusty Gas Model (DGM). The DGM accurately captures molecular diffusion acting in series with Knudsen diffusion which is typical in micro-structured pore networks and the total species flux evaluated also accounts for pressure driven diffusion. The pure component properties are evaluated from standard kinetic theory expressions. The study is carried out from the perspective of using CPR as an on-board reformer for SOFC-APU by recycling the anode stream. From the application perspective the present study is quite distinct from the above-mentioned reports.

2. Numerical model

The numerical model presented here treats coupled interactions of mass and heat transport with detailed kinetic schemes for chemical reactions occurring on the catalyst surface.

Channel flow: The flow of gas mixture through the top and bottom channels is modeled as plug flow according to (Zhu et al., 2005)

$$\frac{d(\rho u Y_k)}{dz} = \frac{4}{D_h} J_k W_k, \quad k = 1, \dots, K_g. \quad (1)$$

Here J_k is the molar flux of species k at the interface between the flow channel and the porous catalyst bed. Since the channel dimensions we consider here are quite small (~ 1 mm), the plug flow assumption is adequate to represent the species transport in the channels. Summing Eq. (1) over all chemical species K_g leads to the total continuity equation as

$$\frac{d(\rho u)}{dz} = \sum_{k=1}^{K_g} \frac{4}{D_h} J_k W_k. \quad (2)$$

The local density in the flow channel is calculated from the ideal gas equation as

$$\rho = \frac{p \bar{W}}{RT}, \quad (3)$$

with \bar{W} defined as

$$\bar{W} = \sum_{k=1}^{K_g} X_k W_k. \quad (4)$$

The energy balance equation for the flow channels is written as

$$\rho u c_p \frac{dT}{dz} = \frac{4}{D_h} h (T_s - T) + \frac{4}{D_h} \sum_{k=1}^{K_g} J_k W_k h_k. \quad (5)$$

The heat transfer coefficient h in the flow channels is defined as

$$h = \frac{\text{Nu} \lambda_f}{D_h} \quad (6)$$

The Nusselt number Nu is defined according to (Hayes and Kolaczkowski, 1997)

$$\text{Nu} = 3.095 + 8.933 \left(\frac{1000}{\text{Gz}} \right)^{-0.5386} \exp \left(- \frac{6.7275}{\text{Gz}} \right), \quad (7)$$

where Gz is the Graetz number defined as

$$\text{Gz} = \frac{D_h}{z} \text{Re Pr}. \quad (8)$$

Porous media flow: The flow through porous bed is modeled as one dimensional, but transverse to the direction of flow through the channels as

$$\frac{dJ_k}{dy} = A_s \dot{s}_k, \quad k = 1, \dots, K_g. \quad (9)$$

The species molar flux J_k in the porous bed is evaluated using DGM equation as (Zhu et al., 2005)

$$J_k = - \sum_{l=1}^{K_g} \mathcal{D}_{kl}^{\text{DGM}} \nabla [X_l] - \left(\sum_{l=1}^{K_g} \mathcal{D}_{kl}^{\text{DGM}} \frac{[X_l]}{\mathcal{D}_{l,\text{Kn}}^e} \right) \frac{B_g}{\mu} \nabla p. \quad (10)$$

The DGM diffusion coefficients $\mathcal{D}_{kl}^{\text{DGM}}$ can be represented as a matrix inverse

$$\mathcal{D}_{kl}^{\text{DGM}} = \mathcal{H}^{-1}, \quad (11)$$

where the elements of the \mathcal{H} matrix are

$$h_{kl} = \left[\frac{1}{\mathcal{D}_{k,\text{Kn}}^e} + \sum_{j \neq k} \frac{X_j}{\mathcal{D}_{kj}^e} \right] \delta_{kl} + (\delta_{kl} - 1) \frac{X_k}{\mathcal{D}_{kl}^e}. \quad (12)$$

The Knudsen diffusion coefficient $\mathcal{D}_{l,\text{Kn}}^e$ in the above equation is defined as

$$\mathcal{D}_{l,\text{Kn}}^e = \frac{\varepsilon d_p}{\tau} \sqrt{\frac{8RT}{\pi W_k}}. \quad (13)$$

Solid phase heat balance: Since the time scale of heat transport in the solid media is larger compared to mass transport and chemical kinetics transient equations are used for the heat balance. Moreover, the thickness of parallel plate reactor is much smaller compared to its length, therefore, any variation in temperature along the thickness is neglected and hence modeled as one dimensional according to

$$\rho_s c_{ps} \frac{\partial T_s}{\partial t} = \frac{\partial}{\partial z} \left(\lambda \frac{\partial T_s}{\partial z} \right) - \frac{4}{D_h} h (T_s - T). \quad (14)$$

The first and second terms on the right-hand side of Eq. (14), respectively, represent the axial conduction in the parallel plate and the heat lost to the gas-phase.

The equation systems are solved using a pseudo-transient approach. A time integration is carried out for the solid phase heat balance. For each time step, Eqs. (1)–(3), (5) and (9) form a system

Table 1
Inlet gas compositions and geometric parameters used for calculations.

Gas-phase	Reform side	Combustion side	Geometry	Reforming side	Combustion side
Inlet conditions			Flow channel		
Composition % (vol)	–	30.0% CH ₄	Channel length (cm)	5	5
	–	14.7% O ₂	Channel height (mm)	1	1
	–	55.3% N ₂	Catalytic bed		
Velocity (m s ⁻¹)	–	0.4	thickness (μm)	10–100	50
Temperature (K)	923.15	923.15	Porosity	40%	40%
Pressure (atm)	1	1	Tortuosity	4.0	4.0
			Pore diameter (μm)	1	1
			Particle diameter (μm)	1	1
			Specific area × 10 ⁻⁵ (m ⁻¹)	0.5–15	3.5
			Thermal diffusivity × 10 ⁶ (m ² s ⁻¹)	3.4	3.4
Solid wall parameters					
Thickness (mm)	1				
Thermal diffusivity × 10 ⁶ (m ² s ⁻¹)	45				

The compositions and velocity at the reforming channel inlet is decided by the recycle ratio used. For parameter study the inlet compositions at the reforming channel are fixed as 13.6% CH₄, 40.8% H₂O, 20.4% CO₂ and 25.2% N₂ entering at a velocity of 0.2 m/s and 923.15 K.

of differential algebraic equations, which are solved after discretization using the DAE solver LIMEX (Deuflhardt et al., 1987).

Boundary conditions: The inlet boundary conditions at time $t=0$ are given in Table 1. At the reactor exit the following Neumann boundary conditions are used:

$$\left. \frac{\partial T}{\partial z} \right|_{z=L} = 0, \quad \left. \frac{du}{dz} \right|_{z=L} = 0, \quad \left. \frac{dY_k}{dz} \right|_{z=L} = 0. \quad (15)$$

At the interface between the wall and the porous bed zero flux boundary condition is imposed for species transport, i.e.,

$$J_k = 0, \quad k = 1, \dots, K_g. \quad (16)$$

Elementary kinetics: A modified Arrhenius expression is used for the calculation of reaction rate constant based on mean field approximation according to

$$k_{fi} = A_i \left(\frac{T}{T_0} \right)^{\beta_i} \exp \left(-\frac{E_{ai}}{RT} \right) \prod_{k=K_g+1}^{K_g+K_s} \theta_k^{\mu_{ki}} \exp \left(-\frac{\varepsilon_{ki} \theta_k}{RT} \right) \quad (17)$$

Here, k_{fi} is the rate constant for the i th reaction, μ_{ki} and ε_{ki} are the parameters modeling the surface coverage dependency of rate constant for i th reaction and θ_k is the surface coverage of the k th chemical species. The site coverages of surface species θ_k can be obtained by solving the following equation:

$$\frac{d\theta_k}{dt} = \frac{\dot{s}_k}{\Gamma}, \quad k = K_g + 1, \dots, K_g + K_s, \quad (18)$$

where Γ is the available site density. At steady state the above equation becomes

$$\dot{s}_k = 0, \quad k = K_g + 1, \dots, K_g + K_s, \quad (19)$$

for surface adsorbed species indicating that the surface composition is invariant in time (although varying spatially). More details about surface chemistry implementation are published elsewhere (Janardhanan et al., 2007).

3. Results and discussions

On-board reforming generally leads to added system complexity since H₂O and/or CO₂ needs to be handled separately. Additionally, since reforming reaction is endothermic it also demands for energy input. Therefore, for practical applications auto-thermal reforming is carried out by recycling the anode stream. In this work we address the problems of heat and H₂O/CO₂ management by using a CPR which couples exothermic oxidation of CH₄ with steam/dry reforming.

Although we are considering CH₄ (major constituent of natural gas) in this study, the modeling framework remains the same for any transportation fuel. Another advantage of using micro-scale reactors for this application is that tight coupling of cell stack and the reformer can be achieved.

A schematic representation of the CPR is shown in Fig. 1. In such a set-up the heat demand for steam reforming is met by the heat liberated from the oxidation reactions. The gas-phase fuel compositions at channel inlets and the geometrical parameters for the reactor are listed in Table 1. The elementary step surface reaction mechanism considered for CH₄ reforming on Ni consists of 42 reactions among six gas-phase species and 12 surface adsorbed species (Janardhanan and Deutschmann, 2006). The surface reaction mechanism used for the partial oxidation of CH₄ on Rh consists of 38 reactions among six gas-phase species and 11 surface adsorbed species (Schwiedernoch et al., 2003). Gas-phase reactions are neglected due to the low temperatures considered here. The rationale for choosing Rh for partial oxidation is that in comparison to Ni, Rh based catalysts display both high activity and stability during the partial oxidation of CH₄ to synthesis gas (de Smet, 2000).

For recycling study we consider 30% CH₄ and 70% air entering the oxidation channel at 923.15 K and 40 cm s⁻¹. Assuming that the products from the oxidation and reforming channels are fully converted to H₂O and CO₂ in a SOFC (i.e., 100% fuel utilization) the molar flow rate of CH₄ stream at the reformer channel inlet is set according to

$$\dot{n}_{\text{CH}_4} = \frac{\dot{n}_{\text{H}_2\text{O}}}{3}. \quad (20)$$

Here R is the recycle ratio and $\dot{n}_{\text{H}_2\text{O}}$ is the molar flow rate of H₂O in the anode stream from SOFC. In SOFC there is mass addition into the anode chamber which is equal to the mass loss in the cathode chamber. However, we neglect any mass added into the anode chamber in calculating the molar flow rate of H₂O at the anode exit.

The exit compositions from the oxidation channel are displayed in Fig. 2. The major products from the oxidation channel are H₂, CO and H₂O. CH₄ conversion reaches steady state after 80 s and H₂O composition decreases as time progresses. Pure air is considered to enter the reforming channel at time $t=0$ and the inlet compositions begin to change as products start forming at the oxidation channel exit. Fig. 3 displays the inlet compositions at the reforming channel inlet as a function of time. 20% of the anode stream is recycled to the reforming channel with additional CH₄ stream according to Eq. (20). Since the anode exhaust consists only H₂O, CO₂, and unreacted N₂, the reforming channel inlet consists of CH₄ in

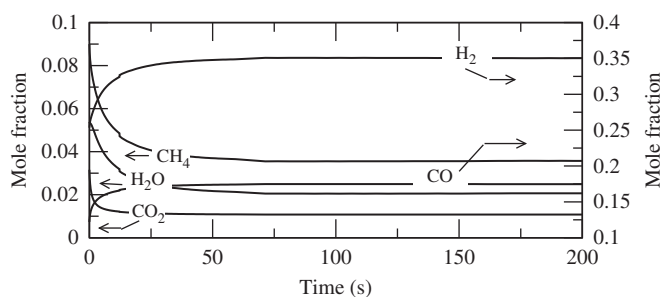


Fig. 2. Species mole fractions at the exit of the oxidation channel. The inlet conditions for this case are given in Table 1. $A_{s,o}=35 \times 10^4 \text{ m}^{-1}$, $L=50 \mu\text{m}$, $u=0.4 \text{ m s}^{-1}$ and $T=923.15 \text{ K}$.

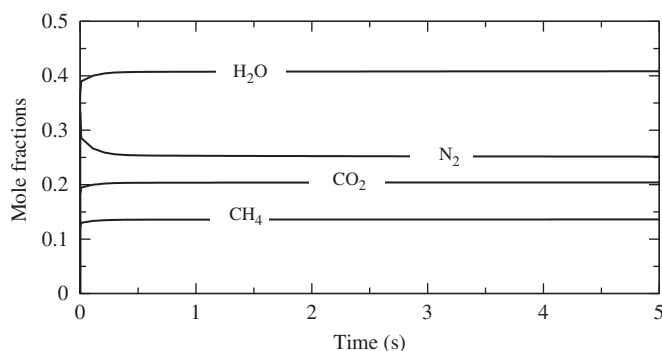


Fig. 3. Species mole fractions at the inlet of reforming channel. 20% of the anode stream is recycled. $A_{s,o}=35 \times 10^4 \text{ m}^{-1}$, $A_{s,r}=50 \times 10^4 \text{ m}^{-1}$, $L=50 \mu\text{m}$ and $T=923.15 \text{ K}$.

addition to the species in the anode exhaust. Eq. (20) maintains a C/O ratio of 3 at the reforming channel inlet.

Assuming that CH_4 is stored on-board at 300 K, and the fuel cell exhaust stream temperature is 1050 K, 20% of the anode recycle is enough to bring the temperature of the mixture at the reforming channel inlet to 792 K. This is calculated according to the following enthalpy balance:

$$\dot{m}_{\text{ref}} h_{\text{ref}}(T) = \dot{m}_{\text{anode}} h_{\text{anode}}(1050 \text{ K}) + \dot{m}_{\text{CH}_4} h_{\text{CH}_4}(300 \text{ K}). \quad (21)$$

In the above equation, the enthalpies are calculated using JANAF polynomials and is solved using Newton iteration. Sensible heat available with the rest of 80% anode stream and the cathode exhaust (O_2 depleted air) can be heat exchanged to bring the temperature of the fuels to 923 K at the reactor inlet. With 20% anode recycle the reforming channel inlet velocity at steady state is $\sim 22 \text{ cm/s}$.

The exit compositions from the reforming channel are shown in Fig. 4. Similar to the oxidation channel exit, steady state is achieved in $\sim 80 \text{ s}$. For the case considered here the exit temperature of the stream is shown in Fig. 5. An optimal condition would be a C/O ratio in the oxidation channel which meets the energy requirement of the reforming reactions in the reforming channel. However, due to the limitations posed by the oxidation mechanism (the elementary kinetics model that we use for partial oxidation of CH_4 on Rh is valid only for C/O ratios above 1.7) we do not strive to seek the optimum conditions. Moreover, the optimum conditions can vary from case to case since they are also depended on other parameters such as inlet flow rates, inlet temperatures, specific area, etc.

The species profiles within the oxidation channel at steady state are shown in Fig. 6. O_2 in the oxidation channel is fully converted within few millimeters from the reactor inlet. Once O_2 is fully converted the H_2O which is formed in the oxidation phase further

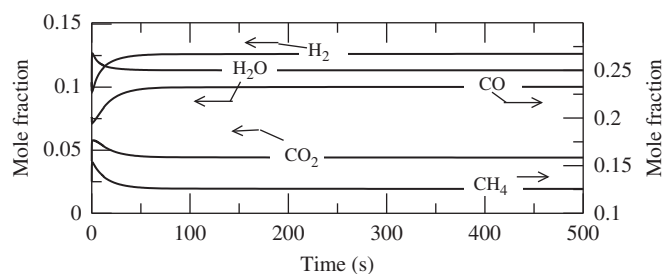


Fig. 4. Species mole fractions at the exit of reforming channel. 20% of the anode stream is recycled. $A_{s,o}=35 \times 10^4 \text{ m}^{-1}$, $A_{s,r}=50 \times 10^4 \text{ m}^{-1}$, $L=50 \mu\text{m}$ and $T=923.15 \text{ K}$.

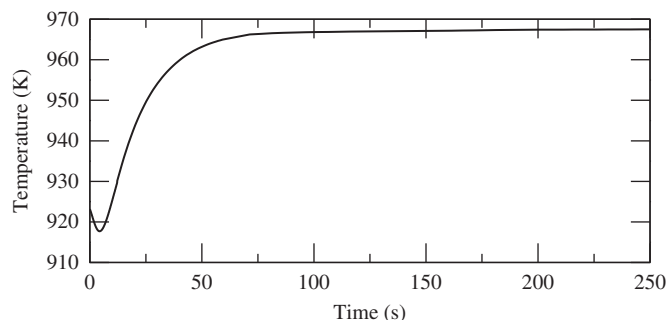


Fig. 5. Temperature of the reforming and oxidation channel at the reactor exit as a function of time. Since they overlap only one is shown.

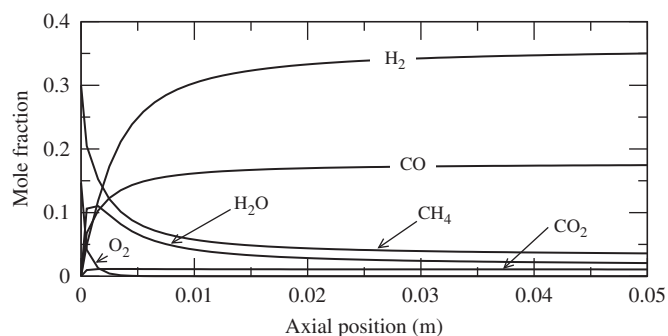


Fig. 6. Species profiles in the oxidation channel as a function of reactor position at steady state. $A_s=35 \times 10^4 \text{ m}^{-1}$, $L=50 \mu\text{m}$ and $T=923.15 \text{ K}$.

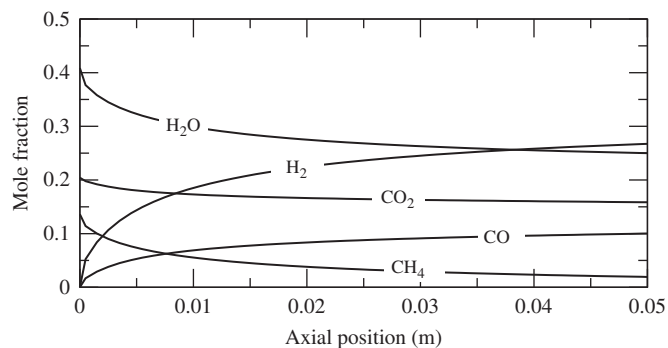


Fig. 7. Species profiles in the reforming channel as a function of reactor position at steady state. $A_s=50 \times 10^4 \text{ m}^{-1}$, $L=50 \mu\text{m}$ and $T=923.15 \text{ K}$.

converts CH_4 downstream the reactor. CH_4 mole fractions at the exit of the reactor is about 3.5%. Similarly the species profiles for the reforming channel at steady state are shown in Fig. 7. Unlike the

oxidation channel the species profiles does not reach steady state within the reforming channel. However, the bed thickness can be increased in-order to achieve complete conversion of CH_4 within the reactor.

For parameter study we fix the inlet compositions at the reforming channel as 13.6% CH_4 , 40.8% H_2O , 20.4% CO_2 and 25.2% N_2 entering at a velocity of 0.2 m s^{-1} and 923.15 K. The oxidation channel inlet composition is given in Table 1. The inlet velocity and bed thickness at the oxidation side are maintained, respectively, at 40 cm s^{-1} and $50 \mu\text{m}$.

The conversion in the reforming channel and the channel exit temperature as a function of specific catalyst area is shown in Fig. 8. As expected the conversions increase with increasing specific area. Specific area of the Rh bed is kept constant at $35 \times 10^4 \text{ m}^{-1}$ and that of Ni bed is varied from 5×10^4 to $150 \times 10^4 \text{ m}^{-1}$. The conversion reaches the maximum possible value asymptotically. Significant increase in conversion is achieved at lower values. Increasing specific area beyond $100 \times 10^4 \text{ m}^{-1}$ does not affect the conversion significantly. Increasing the specific area of the reforming bed also brings down the exit temperature. The temperature drops because of the higher extent of reforming achieved due to increased specific area.

The conversion as a function of bed thickness on the reforming side is displayed in Fig. 9. The thickness of Rh bed is kept constant at $50 \mu\text{m}$ and the Ni bed thickness is varied from 10 to $100 \mu\text{m}$. Similar to specific area, CH_4 conversion also increases with increasing bed thickness. From the trend it is obvious that the conversion will level off asymptotically with bed thickness.

The selectivity for H_2 in the reforming channel as well as in the oxidation channel is shown in Fig. 10. As expected the selectivity increases with increasing specific area on the reforming side,

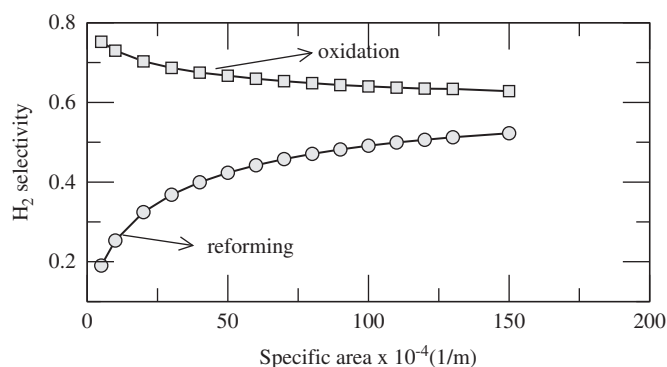


Fig. 10. H_2 selectivity a function of specific area. The inlet velocity for the reforming channel is 0.2 m s^{-1} and the bed thickness is $50 \mu\text{m}$.

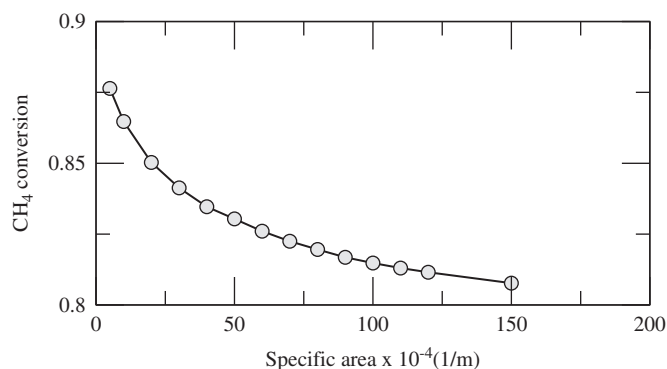


Fig. 11. Methane conversion in the oxidation channel as a function of specific catalyst area on the reforming side. The specific area on the oxidation bed is kept constant at $35 \times 10^4 \text{ m}^{-1}$. The inlet velocity for the reforming channel is 0.2 m s^{-1} and both the bed thicknesses are $50 \mu\text{m}$.

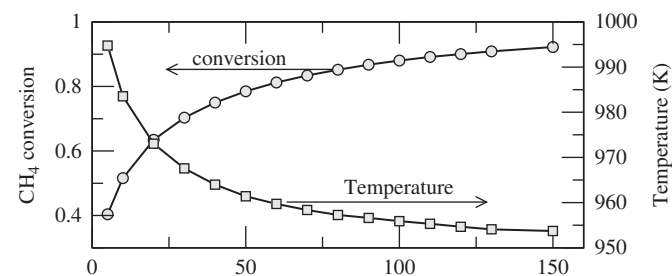


Fig. 8. Methane conversion in the reforming channel and the products exit temperature as a function of specific catalyst area. The specific area on the oxidation bed is kept constant at $35 \times 10^4 \text{ m}^{-1}$. The inlet velocity for the reforming channel is 0.2 m s^{-1} and the bed thickness is $50 \mu\text{m}$.

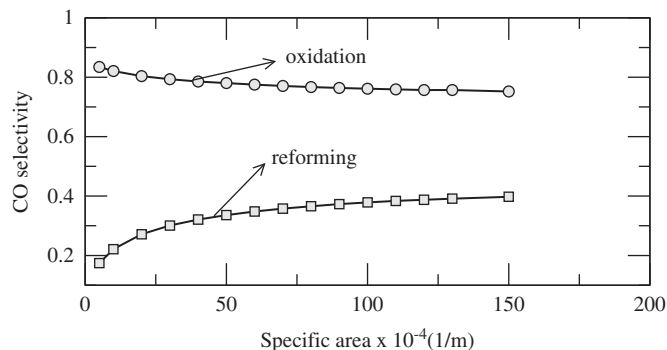


Fig. 12. CO selectivity a function of specific area. The inlet velocity in the reforming channel is 0.2 m s^{-1} and both the bed thicknesses are $50 \mu\text{m}$.

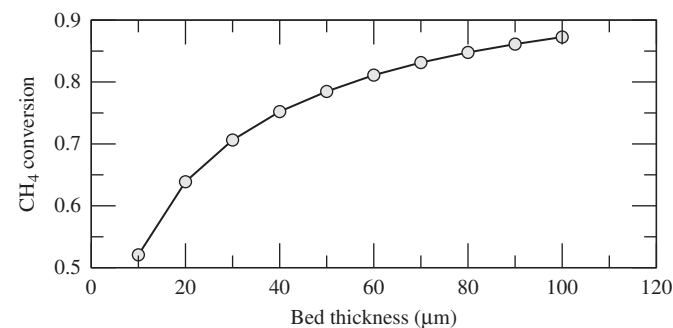


Fig. 9. Methane conversion in the reforming channel as a function of bed thickness. The bed thickness of oxidation channel is kept constant at $50 \mu\text{m}$. Specific catalyst area for oxidation side is 35×10^4 and for reforming side is $50 \times 10^4 \text{ m}^{-1}$. The inlet velocity for reforming channel is 0.2 m s^{-1} .

whereas it decreases slightly on the oxidation side. Fig. 11 shows the CH_4 conversion in the oxidation channel as a function of specific area on the reforming side. As the specific area increases, more CH_4 is converted on the reforming side. However, the temperature drops due to the higher extent of reforming and as a result CH_4 conversion drops on the oxidation side. Since the conversion drops the H_2 selectivity also decreases correspondingly. The CO selectivity for both the sides is shown in Fig. 12. The selectivity to CO is higher compared to H_2 for the oxidation channel. However, for the reforming channel even though the CO selectivity is comparable to

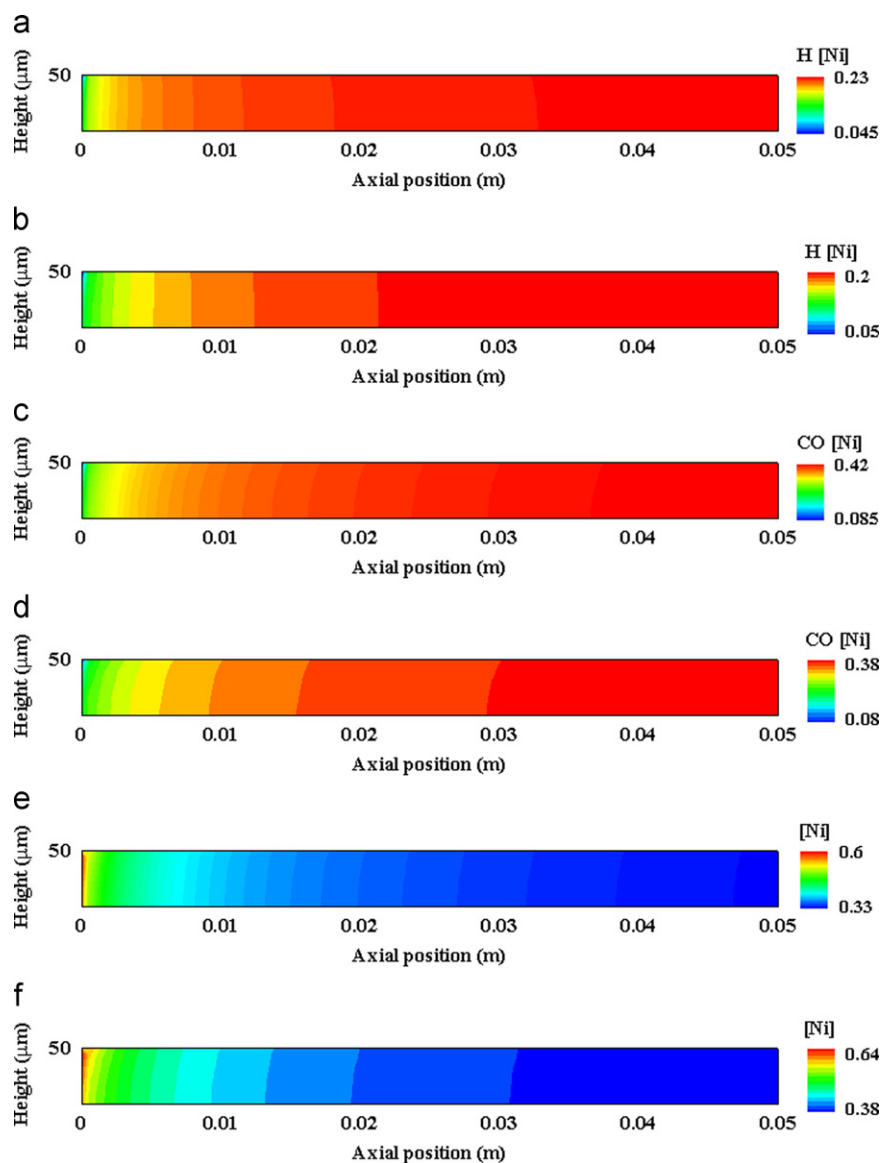


Fig. 13. Surface coverages of major surface adsorbed species and the free Ni surfaces on the reforming side at 0.1 ms and at steady state. (a) Surface coverage of H ad-atoms on the reforming bed at 0.1 ms. (b) Surface coverage of H ad-atoms on the reforming bed at steady state. (c) Surface coverage of CO on the reforming bed at 0.1 ms. (d) Surface coverage of CO on the reforming bed at steady state. (e) Free Ni surfaces on the reforming bed at 0.1 ms. (f) Free Ni surfaces on the reforming bed at steady state.

that of H_2 selectivity, at higher specific area, the selectivity is more towards H_2 .

The coverages major surface adsorbed species at 0.1 ms and at steady state on the reforming bed and the oxidation bed are shown, respectively, in Figs. 13 and 14. The surface coverages of H atoms on Ni surface at 0.1 ms and steady state are shown, respectively, in Fig. 13(a) and (b). The initial surface coverages of H atoms are higher compared to that at steady state. Fig. 13(c) and (d) shows that CO is the major surface adsorbed species initially as well as at steady state. Free Ni surfaces at 0.1 ms as well as at steady state are shown, respectively, in Fig. 13(e) and (f).

In the case of oxidation the surface coverage of H atoms is insignificant at 0.1 ms (Fig. 14(a)) compared to that at steady state (Fig. 14(b)). CO is the major surface adsorbed species at initial time scales as well as at steady state which are displayed, respectively, in Fig. 14(c) and (d). Significant C deposition is observed downstream the reactor (Fig. 14(e) and (f)), which stems from unreacted CH_4 in the absence of a reactant. Near the reactor inlet, there is an enough oxygen which converts CH_4 into products and therefore the C deposition is minimal near the inlet. Fig. 14(g) and (h), respectively,

shows the free Rh surfaces at 0.1 ms and at steady state. Due to C coverage downstream the reactor Rh surfaces are not freely available for adsorption.

Finally a comparison with thermodynamic equilibrium predictions can ensure that the calculations performed do not violate the underlying physics. For comparison with equilibrium predictions, we carry out isothermal calculations. The inlet conditions are same as those used for parameter study. The bed thickness on both sides is fixed at $50 \mu\text{m}$ and the specific area is fixed at $35 \times 10^4 \text{ m}^{-1}$ for the oxidation side and $50 \times 10^4 \text{ m}^{-1}$ for the reforming side. Figs. 15 and 16 shows the CH_4 conversion, respectively, on the oxidation side and the reforming side in comparison with equilibrium predictions as a function of temperature. In both cases the kinetic predictions are within the limit set by equilibrium.

4. Conclusions

We have presented a quasi-two dimensional pseudo-transient numerical model for a catalytic plate reactor. The viability

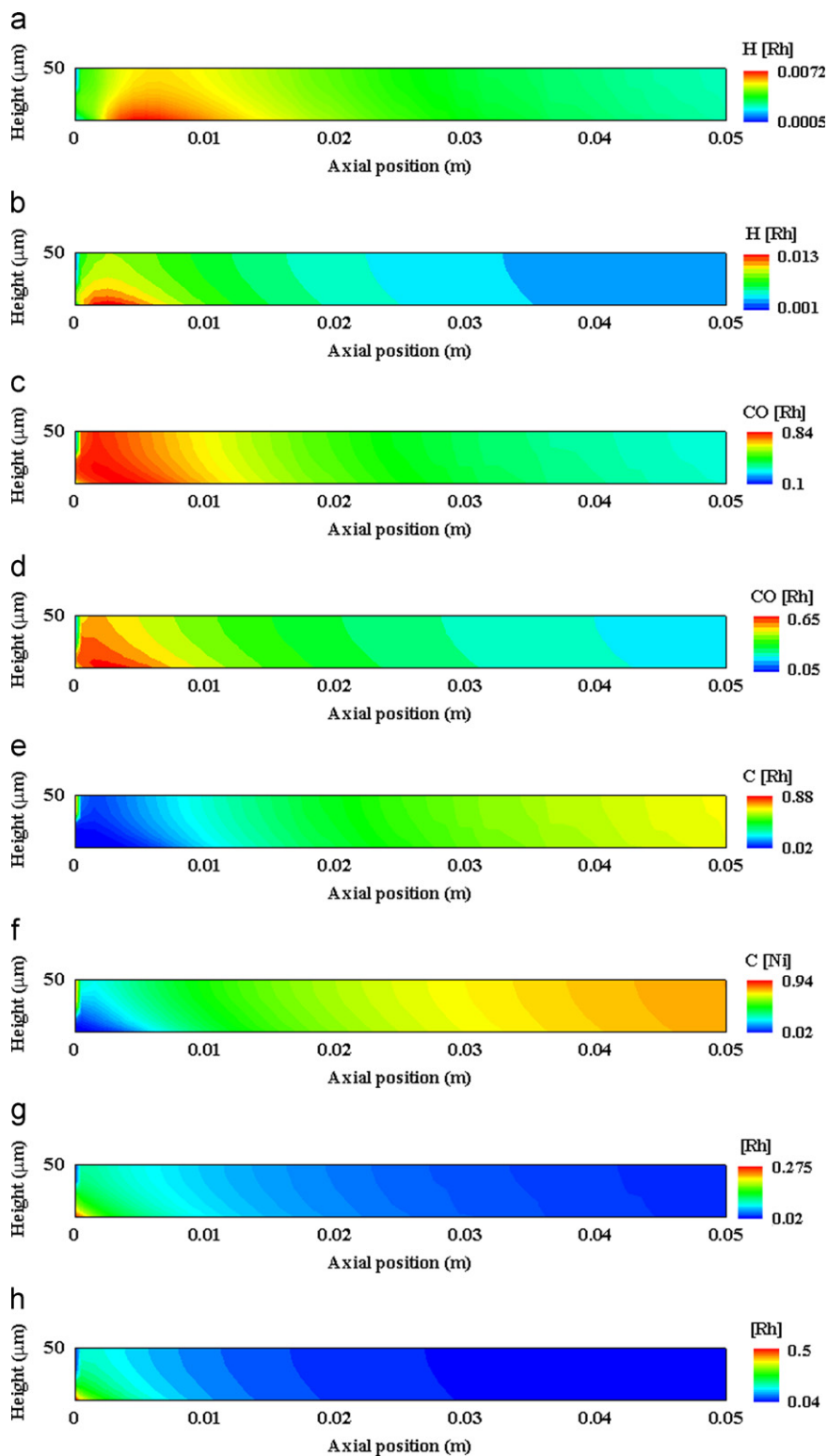


Fig. 14. Surface coverages of major surface adsorbed species and the free Rh surfaces on the oxidation side at 0.1 ms and at steady state. (a) Surface coverage of H ad-atoms on the oxidation bed at 0.1 ms. (b) Surface coverage of H ad-atoms on the oxidation bed at steady state. (c) Surface coverage of CO on the oxidation bed at 0.1 ms. (d) Surface coverage of CO on the oxidation bed at steady state. (e) Surface coverage of C on the oxidation bed at 0.1 ms. (f) Surface coverage of C on the oxidation bed at steady state. (g) Free Rh surfaces on the oxidation bed at 0.1 ms. (h) Free Rh surfaces on the oxidation bed at steady state.

of catalytic plate reactor as on-board reformer is studied. The problems associated with $\text{H}_2\text{O}/\text{CO}_2$ management for on-board reforming can be mitigated by recycling the anode stream with the heat released from the oxidation reactions

proceeding in the parallel channel of CPR. The system studied exhibits an excellent heat management by supplying the necessary heat requirement of the endothermic reforming reactions.

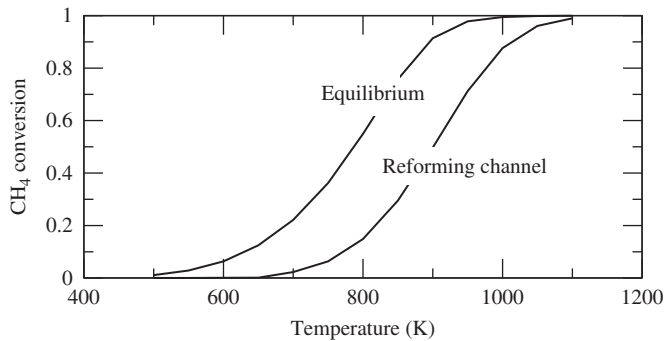


Fig. 16. Comparison of CH₄ conversion on the reforming side with equilibrium predictions.

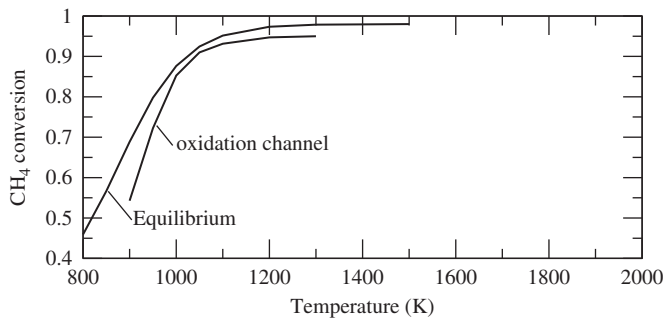


Fig. 15. Comparison of CH₄ conversion on the oxidation side with equilibrium predictions.

Nomenclature

A_i	pre exponential factor
A_s	specific catalyst area (m^{-1})
$A_{s,o}$	specific catalyst area for oxidation bed (m^{-1})
$A_{s,r}$	specific catalyst area for reforming bed (m^{-1})
B_g	permeability (m^2)
c_p	specific heat (J/kg K)
D_h	hydraulic diameter of the flow channel (m)
d_p	particle diameter (m)
D_{kl}^{DGM}	matrix of dusty gas model diffusion coefficients ($\text{m}^2 \text{s}^{-1}$)
$D_{l,Kn}^e$	effective Knudsen diffusion coefficient of species l ($\text{m}^2 \text{s}^{-1}$)
$D_{k,m}$	diffusion coefficient of species k in the mixture ($\text{m}^2 \text{s}^{-1}$)
D_k^e	effective diffusion coefficient of species k in the porous media ($\text{m}^2 \text{s}^{-1}$)
D_{kl}^e	effective binary diffusion coefficient of species k in the porous media ($\text{m}^2 \text{s}^{-1}$)
\bar{D}_k	diffusion coefficient of species k in the porous media ($\text{m}^2 \text{s}^{-1}$)
E_{ai}	activation energy for i th reaction (J mol^{-1})
d_p	pore diameter (m)
Gz	Graetz number
H	specific enthalpy of the mixture (J kg^{-1})
h	heat transfer coefficient ($\text{J m}^{-2} \text{s}^{-1} \text{K}^{-1}$)
h_k	specific enthalpy of species k (J kg^{-1})
J_k	molar flux of species k ($\text{mol m}^{-2} \text{s}^{-1}$)
K_g	number of gas-phase species
K_s	number of surface species
k_{fi}	forward reaction rate constant
L	bed thickness (m)
\dot{m}	mass flow rate ($\text{kg m}^{-2} \text{s}^{-1}$)
\dot{n}	molar flow rate ($\text{mol m}^{-2} \text{s}^{-1}$)
Nu	Nusselt number

p	pressure (Pa)
Pr	Prandtl number
R	recycle ratio
R	universal gas constant ($\text{J mol}^{-1} \text{K}^{-1}$)
Re	Reynolds number
\dot{s}_k	surface molar production rate of species k ($\text{mol m}^{-2} \text{s}^{-1}$)
T	fluid temperature (K)
T_s	solid temperature (K)
u	average velocity (ms^{-1})
W_k	molecular weight of species k (kg mol^{-1})
\bar{W}	average molecular weight of the mixture (kg mol^{-1})
$[X_l]$	concentration of species l
X_k	mole fraction of species k
Y_k	mass fraction of species k
y	Y coordinate (m)
z	axial coordinate (m)
β_i	temperature exponent for modified Arrhenius expression
δ_{kl}	Kronecker delta
ε	porosity
ε_{ki}	coverage dependent activation energy for species k in reaction i
θ_k	surface coverage of species k
λ	thermal conductivity ($\text{J m}^{-1} \text{s}^{-1} \text{K}^{-1}$)
μ	viscosity ($\text{Kg m}^{-1} \text{s}^{-1}$)
ρ	density (kg m^{-3})
τ	tortuosity
$\dot{\omega}_k$	gas-phase molar production rate of species k ($\text{mol m}^{-3} \text{s}^{-1}$)
Γ	site density (mol m^{-2})

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