Towards a particle based approach for multiscale modeling of heterogeneous catalytic reactors

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Towards a particle based approach for multiscale modeling of heterogeneous catalytic reactors

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Highlights

- Simulation of convection-reaction-diffusion for heterogeneous catalytic reactors.
- Numerical computation of bulk mutual diffusivity using SRD.
- Incorporation of complex reaction systems while maintaining the interfacial physics.
- Dimensionless analysis to convey physical information between reactive systems.
- Different modeling examples show from linear systems to nonlinear reaction systems.

Abstract

Particle based approaches are one of the recent modeling techniques to overcome the computational limitation in multiscale modeling of complex processes, for example a heterogeneous catalytic reactor. We propose an efficient model for a chemical reactor where hydrodynamics of the solvent is determined by Stochastic Rotation Dynamics and a reaction occurs over a catalytic surface where the reaction kinetics follows the mean-field assumption. We highlight the modeling techniques required to simulate such a system and then validate the model for its separate and combined components of convection, diffusion and reaction(s). A dimensionless analysis helps compare processes occurring at different scales. We determine the Reynolds number, Re, and the Damkohler numbers, Da and DaL, in terms of key quantities. The approach is then used to analyze a reaction (a) following the Langmuir-Hinshelwood kinetics, (b) generating product particles with different self-diffusivity values as compared to the reactant particles. The model developed can further incorporate reactions occurring inside complex geometries (pore diffusion) and also be used to study complex reaction systems for which the mean-field assumption is no longer valid.

1. Introduction

The ability to model physical systems occurring at multiple time scales, with the current computational power at hand, has been extensively used to resolve or discover new phenomena mainly in the field of fluid and solid mechanics (Chen and Doolen, 1998; Steinhauser, 2007), polymers (Sheng et al., 2004), proteins (Kmiecik et al., 2016), and catalysis (Saliccioli et al., 2011).

Fluid dynamics and catalysis problems occur at comparatively different time scales which makes it difficult to incorporate the complete information about one process while modeling the other and vice versa. In the field of Computational Fluid Dynamics, incorporating detailed reaction kinetics is helpful in improving the...
measurements of concentration distribution of reactants or products across the reactor system (Eigenberger, 2008; Kuroki et al., 2009; Wehinger et al., 2015). However, solution convergence at highly nonlinear conditions requires a very high grid resolution which slows down the whole scheme. Sengar et al. (2017) have previously shown that even a simple $A \rightarrow B$ reaction can show nonlinear dependence on reactant concentration under certain possible conditions.

Recent mesoscale modeling techniques, involving a particle based treatment of fluid, like Lattice Boltzmann method (LBM), Stochastic Rotation Dynamics (SRD), and Dissipative Particle Dynamics (DPD) offer an alternative approach in modeling these multiphysics problems. Treatment of particles in ensembles at different spatiotemporal scales reduces the total number of degrees of freedom inside the system, making it computationally more feasible. These modeling techniques, therefore, allow a faster implementation of detailed reaction kinetics coupled with fluid dynamics.

The field of particle based simulations has previously been used to study multiscalar coupling involving flow and adsorption in porous media using a reactive Lattice Based scheme (Vanson et al., 2015), and diffusion and adsorption in MetalOrganic Frameworks (Villemet et al., 2014), transport properties during convection-diffusion-reaction using DPD (Li et al., 2015). For the present work, we will employ SRD to simulate heterogeneous nonlinear reactions.

Stochastic Rotation Dynamics (SRD) (Malevanets and Kapral, 1999), uses a discrete-time dynamics of coarse-grained particles (Tüzel et al., 2008) and heterogenous (Tucci and Kapral, 2004; Huang et al., 2007; Padding and Louis, 2006), homogenous (Rohlf et al., 2009; Wehinger et al., 2015). However, solution convergence at unsteady state. To simulate the reaction step, the complete reaction cycle is broken down into individual reaction steps like adsorption, desorption and surface reaction (Sengar et al., 2017). This makes it convenient to observe the effects of any intermediary reaction process on the bulk fluid mixture by making the other processes fast enough.

As the number of participating components in the system increases, the classical definition of self diffusion coefficient, according to Fickian Diffusion, is not enough (Krishna and Wesselingh, 1997; Taylor and Krishna, 1993). Further, for reactions involving gases, a difference in number of moles of products and reactants will result in an internal pressure gradient which can still be accounted for in the mutual diffusivity at unsteady state or a pseudo steady state. We incorporate all these effects in the Maxwell-Stefan diffusivity.

In Section 2.1, we briefly discuss the particle-particle interactions in SRD followed by a description of the numerical techniques, Section 2.2, to calculate the Maxwell-Stefan diffusivity for a binary mixture of SRD particles with different mass leading to different self-diffusivity. A set of dimensionless parameters are defined in Section 3.1 to optimize the processes occurring at different time and length scales, treated in subsequent Sections 3.2, 3.3, 3.4, 3.5. We first perform a cold flow analysis on flow around a sphere, Section 3.2, and pressure drop across a packed bed, Section 3.3, to validate the viscous and convective coupling in an SRD fluid in the presence of spherical obstacles. This is followed by the introduction of a chemical reaction on the surface of the spheres and introduction of three dimensionless numbers, which are the Damkohler numbers $Da$ and $Da_0$, and the Peclet number $Pe$. The model is validated for reaction-diffusion coupling over a catalytic sphere placed in a pool of reactant in Section 3.4. In Section 3.5, we finally combine the three processes convection, diffusion and reaction, and analyse the net conversion as a function of the above.

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Table E.5

List of symbols in order of appearance in the manuscript.

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$a_0$</td>
<td>SRD unit cell [m]</td>
</tr>
<tr>
<td>$k_B$</td>
<td>Boltzmann constant [kg m$^{-2}$ s$^{-1}$ K$^{-1}$]</td>
</tr>
<tr>
<td>$m_i$</td>
<td>Mass of single SRD particle of species $i$ [kg]</td>
</tr>
<tr>
<td>$\Delta t$</td>
<td>Hydrodynamic time step [s]</td>
</tr>
<tr>
<td>$x$</td>
<td>SRD rotational angle</td>
</tr>
<tr>
<td>$\gamma_i$</td>
<td>Number of species $i$ particles per unit SRD cell [m$^{-3}$]</td>
</tr>
<tr>
<td>$D_i$</td>
<td>Self-diffusivity of species $i$ [m$^{2}$ s$^{-1}$]</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Friction force per particle of species $i$ [m$^{-1}$]</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Chemical potential of particle of species $i$ [kg m$^2$ s$^{-2}$]</td>
</tr>
<tr>
<td>$\varepsilon_i$</td>
<td>Number fraction of species $i$</td>
</tr>
<tr>
<td>$T$</td>
<td>Temperature [K]</td>
</tr>
<tr>
<td>$u_i$</td>
<td>Average velocity of particles of species $i$ [m s$^{-1}$]</td>
</tr>
<tr>
<td>$u_i$</td>
<td>External force on single particle of species $i$ [kg m s$^{-2}$]</td>
</tr>
<tr>
<td>$a_i$</td>
<td>Acceleration because of external force on particle of species $i$ [m s$^{-3}$]</td>
</tr>
<tr>
<td>$v_i$</td>
<td>Velocity of individual particle of species $i$ [m s$^{-1}$]</td>
</tr>
<tr>
<td>$N_i$</td>
<td>Number of spheres in simulation setup</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Vacancy fraction over the surface</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Fraction of species $i$ over sphere surface</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Probability of adsorption of species $i$ on sphere surface</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Probability of desorption of species $i$ from sphere surface</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Probability of reaction of species $i$ on sphere surface</td>
</tr>
<tr>
<td>$N_{cat}$</td>
<td>Number of catalytic sites on a single sphere</td>
</tr>
<tr>
<td>$k_{cat}$</td>
<td>Rate constant for surface process $X$ occurring for particle $i$ per area of catalyst surface</td>
</tr>
<tr>
<td>$k_{ads}$</td>
<td>Adsorption-desorption Equilibrium constant for species $i$ [m$^3$]</td>
</tr>
<tr>
<td>$D_{ads}$</td>
<td>Damkohler number</td>
</tr>
<tr>
<td>$D_{des}$</td>
<td>Longitudinal dispersion coefficient [m$^2$ s$^{-1}$]</td>
</tr>
<tr>
<td>$D_{r}$</td>
<td>Tortuosity of packed bed column</td>
</tr>
<tr>
<td>$D_{l}$</td>
<td>Average velocity of solvent in packed bed column [m s$^{-1}$]</td>
</tr>
<tr>
<td>$t_{sim}$</td>
<td>Initial run time of simulation setup to achieve mechanical equilibrium [s]</td>
</tr>
<tr>
<td>$r_i$</td>
<td>Volumetric reaction rate constant [m$^3$ s$^{-1}$]</td>
</tr>
<tr>
<td>$A_{catal}$</td>
<td>Area of single catalytic sphere [m$^2$]</td>
</tr>
<tr>
<td>$D_{l}$</td>
<td>Damkohler number along the longitudinal direction of simulation setup</td>
</tr>
<tr>
<td>$Pe$</td>
<td>Viscous length in the simulation setup where reaction occurs [m]</td>
</tr>
<tr>
<td>$r_i$</td>
<td>Schmidt number</td>
</tr>
<tr>
<td>$\rho_i$</td>
<td>Local number density of species $i$ over catalytic sphere [m$^{-3}$]</td>
</tr>
<tr>
<td>$Re$</td>
<td>Number of strips on a single reaction sphere</td>
</tr>
<tr>
<td>$\varepsilon_i$</td>
<td>Area of a single strip [m$^2$]</td>
</tr>
<tr>
<td>$\eta_i$</td>
<td>Diffusive flux of species $i$ [m$^2$ s$^{-1}$]</td>
</tr>
<tr>
<td>$\phi_i$</td>
<td>Total number density of all particles per unit cell [m$^{-3}$]</td>
</tr>
<tr>
<td>$\alpha_i$</td>
<td>Diffusion matrix [m$^2$ s$^{-1}$]</td>
</tr>
<tr>
<td>$\beta_i$</td>
<td>Mass fraction of species $i$</td>
</tr>
</tbody>
</table>
2. Model development

The previous study (Sengar et al., 2017) highlights the use of SRD with reaction kinetics to model simple heterogeneous chemical reactions in a micro reactor. Reactants and products were considered with equal mass and equal diffusivity. The further model development in this work is divided into two parts. The first part talks about the inclusion of characteristically distinct reaction species and obtaining the mutual diffusivity for such a system while the second part explores the geometry of fluid-solid interface as compared to the planar geometry in Sengar et al. (2017).

2.1. Stochastic rotation dynamics

Each participating species of the initial reaction mixture is represented by a system of point-like particles. The system starts with homogeneously distributed \( N \) particles with number fraction and mass of particle of each species being \( \chi_i \) and \( m_i \) respectively. The 3D space is discretized into a cubic lattice of size \( a_0 \) and time is discretized into time intervals of \( \Delta t \), which governs the displacement of particles before successive collisions.

Streaming of particles occurs according to the Euler scheme, \( \mathbf{r}_{i}^{t+1} = \mathbf{r}_{i}^{t} + \mathbf{v}_{i}^{t} \Delta t, \) where \( \mathbf{r}_{i} \) and \( \mathbf{v}_{i} \) are the position and velocity vectors of particle \( i \) at time \( t \). The second step is the velocity update which occurs according to the collision rule \( \mathbf{v}_{i}^{t+1} = \mathbf{v} + \mathbf{F} + \mathbf{G} \Delta t, \) where \( \mathbf{v} \) is the center-of-mass velocity of all particles sharing the same cell and \( \mathbf{G} \) is a stochastic rotation matrix that rotates the velocities by a fixed angle \( \gamma \) while conserving energy and momentum.

To ensure Galilean invariance (Ihle and Kroll, 2003), a random grid-shift procedure (Ihle and Kroll, 2001) is employed. The origin of the lattice is displaced by a stochastic 3D vector (which is same for all lattices per time step) whose magnitude of each Cartesian component is defined by a random number between 0 and \( a_0 \). To account for the under filling of cells near boundaries caused by the grid-shift procedure, introduction of ghost particles for single solvent species (Lamura et al., 2001) was suggested which was expanded to preserve momentum when multiple components might be present in such cells (Sengar et al., 2017). To maintain the system at fixed temperature even in the presence of external forces, a Galilean invariant thermostat (Padding and Louis, 2006), that rescales particle velocities in the center-of-mass velocity frame of their cell, is used.

Transport coefficients like viscosity and diffusion coefficient for single species solvents have extensively been studied in Ihle and Kroll (2003), Tüzel et al. (2003, 2006), Kikuchi et al. (2003), Gompper et al. (2009), Noguchi and Gompper (2008), Pooley and Yeomans (2005). For a single SRD particle species, the diffusion coefficient is written in a parametric form as

\[
D = \frac{k_B T \Delta t}{2m} \left( \frac{3}{2} \gamma - 1 + \gamma \left( 1 - \cos \alpha \right) \right)
\]

When a solution consists of more than one component, the mutual interaction between particles is described by a mutual diffusivity. For systems representing ideal gases with low interaction between particles of different components, a Fickian definition of mutual diffusivity is used. However, for real fluids when cross-interaction terms dominate, a Maxwell-Stefan approach is necessary. Maxwell-Stefan diffusivity for multicomponent mixtures has been studied previously using Molecular Dynamics, see e.g. van de Ven-Lucassen (1998). In this work, we will use a similar approach to analyse binary mixtures.

The force acting on a single particle of species \( i \) in an SRD mixture with \( n \) components, at a constant temperature \( T \) can be written as the gradient of chemical potential, \( \mu_i \) (van de Ven-Lucassen, 1998; Taylor and Krishna, 1993)

\[
d_i = -\nabla \mu_i = \sum_{j=1}^{n} \frac{k_B T}{D_{ij}} (\mathbf{u}_i - \mathbf{u}_j)
\]

\( \mathbf{u}_i \) and \( \chi_i \) are the average velocity and mole fraction of species \( i \) respectively. The solution obeys the ideal gas law \( PV = N_kT \), where \( N \) is the total number of particle of all species. In presence of an external field \( \mathbf{F} \), the net force on a single particle is \( \mathbf{F}_i - \nabla \mu_i \). The above interpretation of \( D_{ij} \) can be obtained from a non-equilibrium or an equilibrium approach.

2.2. Diffusivity in multi-component mixtures

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2.2.1. NEMD approach

The Non-Equilibrium Molecular Dynamics approach (Berendsen, 1991) works by applying equal and opposite external forces on different components in a solution mixture such that the net external force is zero and therefore the net momentum across all particles remains conserved at each discrete time step. For a binary mixture, the approach can be simplified and force and momentum conservation equation can respectively be written as

\[
N m_1 \chi_1 \mathbf{a}_1 + N m_2 \chi_2 \mathbf{a}_2 = 0
\]

\( \mathbf{a}_i \) is the force acting on particles of species \( i \). For small external forces, the velocity response is

\[
\frac{d\mathbf{u}_i}{dt} = \frac{k_B T}{m_i D_{ij}} (\mathbf{u}_j - \mathbf{u}_i)
\]

The factor \( k_B T/D_{ij} \) represents the mutual mobility between species \( 1 \) and \( 2 \) caused by internal friction. The coupled differential
equations can be solved exactly to get the terminal velocity of particles under this external force. For \( t \to \infty \), the terminal velocity of species 1 is

\[
u_1(\infty) = \frac{a_1 m_1 k_B T}{m_1 + m_2} \frac{D_{12}}{m_2} \tag{5}
\]

For sufficiently small external forces, we therefore expect a linear dependence of terminal velocity of a species on the force applied, which can be used to calculate the value of the mutual diffusion coefficient \( D_{12} \).

2.2.2. Green-Kubo relations

Another approach is by employing linear response theory in statistical mechanics. It has been shown (van de Ven-Lucassen, 1998) that the velocity response of a solvent in presence of another species can be used to estimate the Green-Kubo relation for the Maxwell-Stefan diffusion coefficient as

\[
D_{12} = \frac{k_B T}{3N_1} \int_0^\infty \left( \sum_{i=1}^{N_1} \int_0^t \left[ \sum_{j=1}^{N_1} \nu_1^i(t) - \sum_{j=1}^{N_2} \nu_2^j(t) \right] dt \right) \tag{6}
\]

If cross-species velocity terms are neglected, we get the pure mutual diffusion coefficient \( D_{12}^p \).

\[
D_{12}^p = \frac{k_B T}{3} \int_0^\infty \langle \mathbf{v}_1(0) \mathbf{v}_1(t) \rangle dt + \frac{k_B T}{3} \int_0^\infty \langle \mathbf{v}_2(0) \mathbf{v}_2(t) \rangle dt \tag{7}
\]

\( \mathbf{v}_i \) refers to velocities averaged over the entire species.

Simulations were carried out to calculate the diffusion coefficient from the above two approaches which serves as a validation for the SRD description of particles.

2.2.3. Measurement of mutual diffusion coefficient expressions

For the NEMD approach, the system was initialized with an equimolar binary mixture of particles with mass \( m_1 \) and \( m_2 \). The self-diffusivity coefficients of these particles are in ratio \( D_1/D_2 = m_2/m_1 \). In Fig. 1, the terminal velocity of particles of type 1 is calculated versus an applied force \( a_1, m_1 = 1 \) and \( m_2 \) takes values 0.2, 0.5, 1.0, 2.0. When \( m_2 = 1.0 \), the \( D_{12} \) value obtained from the slope is equal to the self-diffusivity coefficient of both particles 1 and 2.

To calculate the mutual diffusion coefficient using the Green-Kubo relation Eq. (6), 2 species of mass \( m_1 = 0.5 \) and \( m_2 = 1.0 \) are chosen. The number fraction of species 1 is varied from 0 to 1 and the velocity correlation functions for both particles are calculated to estimate the mutual diffusivity \( D_{12} \). This is shown in Fig. 2.

The linear curve \( D_{12}^p \) is the pure mutual diffusion coefficient when no cross-correlation terms are included. The higher the difference between \( D_{12} \) and \( D_{12}^p \), the more the different species interact and act like non-ideal mixtures. The difference is maximum when \( \chi_1 = 0.5 \).

3. Convection, diffusion and reaction across spheres

For the next part of the work, we define a sphere, the surface of which may be reactive or not. We will then study the flow properties of a fluid around this sphere as a function of its Reynolds number \( Re \). This is followed by a methodology to incorporate reactions over the surface and the conditions under which a linear approximation for reaction can be done. The linear treatment of reaction is then used to verify (a) the change in flux of the reactant species over the sphere with time, and (b) conversion in a packed-bed reactor with dispersion.

3.1. Dimensionless analysis

In a reactor, the three main processes namely convection, reaction and diffusion never occur at the same time scale. It is convenient in such multiscale systems to measure the variables of a process \( X \) occurring at a time scale with respect to the variables of another process \( Y \) at a different time scale. We, thus, define characteristic dimensions, at a characteristic time scale, to analyze any variable of any process at any time scale.

The three dimensions chosen are the radius of a single sphere \( R \), self-diffusion coefficient of initial solvent species \( D_A \), and the
viscosity of the solvent \( \eta_a \). Since we treat a single SRD particle as part of an ideal fluid, the ideal-gas law is obeyed \( PV = Nk_B T \) with \( N \) being the number of SRD particles. Physical quantities and their dimensions are shown in Table 2. For any physical quantity \( X \), we will represent the corresponding dimensionless quantity as \( X^e \).

3.2. Flow across a sphere

Lamura et al. (2001), Allahyarov and Gompper (2002) have previously shown the use of SRD to model flow past a circular and square 2D cylinder and 3D spheres. Here we will first verify our previously shown the use of SRD to model flow past a circular and square 2D cylinder and 3D spheres. Here we will first verify our implementation, by comparing the predicted flow field around a 3D sphere with their results.

The system consists of a sphere of radius \( R \) present in a box of dimensions \( L_x, L_y, L_z \) with the sphere center at \( L_x/2, L_y/2, L_z/4 \). The boundaries of the simulation box are periodic and the surface of the sphere has no-slip boundary conditions. To ensure that the periodic boundaries do not affect the hydrodynamics of the system, we choose \( L_x > L_y, L_y \geq 10R \) (\( z \) being the direction of flow). An SRD fluid \( A \) is present inside the box at \( t = 0 \). For all \( t > 0 \), a constant pressure difference between \( z = 0 \) and \( z = L_z \) is applied leading to a flow along the \( z \) direction. When mechanical equilibrium is established, the velocity profile of the fluid along the \( z \)-direction can be studied. For small values of \( Re = \rho_A u_{max}(2R)/\eta_A \) (in dimensionless units \( Re = 2\rho_A u_{max}^2/\eta_A \) ), the velocity vectors are streamlined over the surface of sphere. As \( Re \) increases, a recirculation region is formed behind the sphere (Churchill, 1988) followed by vortex shedding (Taneda, 1956) for \( Re > 200 \).

To avoid slip of the velocity field at the sphere surface caused by low number density of particles in cells near the surface, a bulk fill rule for such cells has been implemented. For details, we refer to Appendix A.

The size of the wake region is defined as the length of the region with a negative time-averaged velocity behind the sphere. In Fig. 3, we compare the results obtained from our SRD model with those of Allahyarov and Gompper (2002) and the experimental data (Taneda, 1956). Between \( 20 < Re < 100 \), the graph shows a monotonic increase in the wake length which is in good agreement with established results.

3.3. Pressure drop in a packed column

The wake formation behind a sphere depicts a phenomenon where coupled convection and viscosity effects cause disturbance in the flow of fluid behind a sphere. The fluid faces resistance in its path leading to pressure difference at difference faces of the sphere along the direction of fluid flow. In a packed-bed arrange-ment of spheres, this pressure difference adds up and for laminar flows at relative low \( Re \), the pressure drop per unit length of the reactor can be correlated by the Kozeny-Carman relation

\[
\Delta P/L_z = 150 \frac{\eta_A u_o}{(2R)^2} (1 - \epsilon)^2/\epsilon^3
\]

where \( u_o \) is the superficial velocity of fluid in the reactor and \( \epsilon \) is the gas porosity. The above equation can be non-dimensionalized according to Table 2 as follows:

\[
f = \frac{\Delta P/L_z R^3}{\eta_A D_A} = 75 \frac{u_o R}{2D_A} \frac{(1 - \epsilon)^2}{\epsilon^3} - 75Re \frac{(1 - \epsilon)^2}{4\rho_A^2} \frac{1}{\epsilon^4}
\]

where \( \rho_A^2 \) is the dimensionless mass density and \( Re \) can also be written as

\[Re = 2\rho_A^2 u_o^2\]

To model this pressure drop, we consider an open-boundary 3D box \( (L_x, L_y, L_z) \) with \( N \) spheres randomly placed inside. A periodic boundary system ensures no wall-effects are included. Spheres lying close to the boundaries are wrapped around to contribute to the packed column, see Fig. 5. The SRD fluid is placed around the spheres at \( t = 0 \). The SRD particles are then given an acceleration \( g \) along the \(+z\)-axis. The mean free path of SRD particles should remain comparable or smaller than the smallest distance between spheres inside the packed column. This will ensure that the continuum approximation (Knudsen number \( \ll 1 \) ) will still be valid in these simulations.

At mechanical equilibrium, the hydraulic pressure drop per unit length will be equal to the net force density on all SRD particles inside the packed column is \( \Delta P/L = \rho_A g \).

We can thus study the response of an independent acceleration \( g \) on the dependent variable \( u_o \), according to \( f = \rho_A^2 g \), where \( f \) is given by Eq. (9).

Fig. 4 plots the dimensionless pressure drop \( f \) as a function of \( Re \) and compares the results from the simulations for different bed porosities \( \epsilon \). Dashed lines represent the dimensionless Kozeny-Carman relation and the inverted triangles are the points obtained from SRD results. The match is quite good which shows that the fluid hydrodynamics at low porosities can be efficiently modeled by our current approach.

![Fig. 3. The length of wake behind the sphere as a function of \( Re \).](image)

![Fig. 4. Dimensionless pressure drop as a function of particle Reynolds number, \( Re \), for \( \epsilon = 0.4, 0.5, 0.6 \). For each curve, \( g^* \) is varied to obtain a \( u_o^* \) which determines \( Re \). Simulation parameters in terms of SRD units, Table 1: \( \gamma_A = 200 \text{mN/m}, \Delta_t = 0 \text{ms}, \text{Ma} = 1.0 \sqrt{kT/\rho_A}, N_s = 150, L_z = 3000a, L_x = 300a, L_y = 500a \).](image)
3.4. Irreversible reaction on a sphere

The SRD model developed was validated (Sengar et al., 2017) for a reaction-diffusion case with reaction over a thin catalytic strip. We incorporated Langmuir-Hinshelwood kinetics for a $A=B$ reaction and estimated the product concentration profile based on Damkohler numbers when any single participating process in the reaction cycle (adsorption, desorption or site reaction) was rate determining. It was established that a linear assumption of reaction is only valid when adsorption of reactant species over the catalytic surface will be the rate determining step.

The following validation cases demonstrate application of an SRD based fluid in a system with convection, diffusion and reaction occurring at different time scales.

We first consider an example of a sphere placed in an infinite bath of reacting solvent $A$ such that a linear reaction occurs at the surface of a sphere, Lu et al. (2018). The competing processes are the reaction of $A$ at the surface which leads to a local concentration gradient causing an incoming diffusion flux of $A$. Diffusion and reaction compete in this unsteady state process.

To simulate this process, a sphere of radius $R$ is placed in the center of the simulation box. In this example, the surface of the sphere is one single reactive strip since absence of convection ensures the system is isotropic, Appendix A.

The surface of the sphere, with area $A_s$, consists of $N_{cat}$ reactive sites that are either vacant (with surface fraction $\theta_s$) or contain adsorbed species $i$, with surface fraction $\theta_i$. Fluid particles of species $i$ hitting the catalytic strip, within a time step $\Delta t$, can adsorb with a probability $p_{A,i}^r$ which relates to the adsorption rate constant $k_{ads,i}^r$.

see Table 3. A successful adsorption reduces the fractional vacancy $\theta$ of the catalyst by $1/N_{cat}$ while increasing $\theta$ by $1/N_{cat}$.

Adsorbed particles stay at their positions, representing a surface-diffusion limited process. During the time step $\Delta t$, particles of species $i$ may desorb from the surface or further react with probabilities $p_{A,i}^d$ or $p_{A,i}^r$ that relate to the rate constants of desorption and reaction, see Table 3. The global thermostat, Section 2.1, ensures the reaction is treated isothermally.

For all $t > 0$, $A$ is adsorbed on the surface of the sphere with a probability $p_{A,ads}^r$. As described above, adsorbed $A$, with a fractional occupancy $\theta$, over the catalytic strip, can further desorb (with probability $p_{A,des}^d$) or react (with probability $p_{A,react}^r$) to form adsorbed species $B$, with surface fraction $\theta_B$. Adsorbed $B$ can further desorb (with probability $p_{B,des}^d$) or react back to form adsorbed $A$ (with probability $p_{B,react}^r$). Desorption resulting in bulk particles of $B$ can re-adsorb with a probability $p_{A,ads}^r$. For a linear irreversible reaction assumption, $p_{A,ads}^r = p_{B,ads}^r = p_{A,des}^d = 0$ and $p_{A,react}^r = 1.0$ (Sengar et al., 2017). Such a probability parameter set also ensures that the catalytic surface itself will not have its own inherent time scale since it will be equilibrated almost instantly. Refer to Appendix A for a detailed description of coupling reactions with the bulk.

The reaction at the surface of the sphere gives rise to a concentration gradient leading to a diffusive flux of the reactant towards the catalyst, further fueling the reaction. The surface reactive boundary condition (Collins and Kimball, 1949) is

$$D_{A,0} \frac{\partial n_A(r,t)}{\partial r} \bigg|_{r=R} = k_{ads,i}^r \gamma_A(t)$$

(11)

$D_{A,0}$ is the time varying mutual diffusivity of $A$ and $B$ near the catalytic surface since the process is at an unsteady state. If the catalytic strip achieves a steady state almost instantaneously, no concentration depletion region is formed near the sphere and the total number density of particles over the sphere always remains a constant. In this simple example, since particles $A$ and $B$ are only distinguishable by a tag, $D_{A,0} = D_{A,0}$.

Lu et al. (2018) have presented an analytical solution for the number density $\gamma_A(r,t)$ of reactant species $A$ in the region surrounding the sphere.

$$\gamma_A(r,t) = 1 - \frac{D_A}{r^2(1+D_A)} \left[ \text{erf} \left( \frac{x}{2\sqrt{\tau}} \right) - e^{-x^2} \text{erfc} \left( \frac{2\tau + x}{2\sqrt{\tau}} \right) \right]$$

$$\tau = \left(1 + D_A \right)^{-1}, x = (1 + D_A) \left( r^2 - 1 \right)$$

(12a)

(12b)

where $\gamma_A(1,0)$ is the constant number density of $A$ in the region outside the sphere, $r > R$, at $t = 0$. The Damkohler number $Da$ is equal to dimensional-less rate constant $k_{ads,i}^r$.
For $x = 0$, $r^s = 1$, close to the surface of the sphere, Eq. (12) becomes

$$\frac{\gamma_A(1, r^s)}{\gamma_A(1, 0)} = 1 - \frac{Da}{1 + Da} [1 - e^{-erf(\sqrt{r})}]$$

(13)

At steady state, $\lim \tau \rightarrow \infty$,

$$\frac{\gamma_A(1, r^s)}{\gamma_A(1, 0)} = \frac{1}{1 + Da}$$

(14)

For a reaction limited case, $Da \ll 1$, the local concentration over the surface of the sphere stays almost constant with time and for a diffusion limited case, $Da \gg 1$, the diffusive flux of reactants is not able to compensate for the fast reaction and the concentration decreases quickly near the catalytic surface, see Fig. 6. For fast reactions, the dimensionless half-life of the reactant concentration near the catalytic surface occurs approximately over the time scale $t^s = 1/(1 + Da)^2 = 1/Da^2$ where $t^s = 1$ represents time scale for diffusion over a length scale of sphere radius. The logarithm of the dimensionless half-life time is plotted against the logarithm of Damkohler number, Fig. 7.

Fig. 6 shows the concentration of $A$ as a function of distance $r^o$ away from the sphere after a diffusive time scale, $r^o = 1$ for different Damkohler numbers.

It becomes a computational challenge to tackle a simple reaction-diffusion case when $Da \ll 1$ or $Da \gg 1$. As has been observed in this example, SRD is quite efficient in resolving the two processes even when they occur at complete different time scales. Next step is to introduce a third time scale into our problem, the convective time scale.

3.5. Dispersion with reaction in a packed bed

Up to this point, we have validated the model for at most two competing processes (reaction - diffusion, convection - diffusion) at the same time. Now, we introduce a linear reaction, as described in Section 3.4, over the surfaces of all spheres in the packed bed, as defined in Section 3.3, and study conversion across the reactor length as a function of the interplay between convection, diffusion and reaction.

The system, see Fig. 5, has periodic boundaries at $x = 0$, $x = L_x$, $y = 0$, $y = L_y$, $z = 0$, $z = L_z$. The direction of flow is the $z$-direction. To neglect transverse dispersion effects, we will always ensure that $L_x/L_z > 10$, $L_y/L_z > 10$ (Klinkenberg et al., 1953; Delgado, 2006). For gas-like fluids, the longitudinal dispersion $D_l$ depends on the bulk diffusivity of the fluid components, the average fluid velocity and the sphere diameter (Gunn, 1987; Alves et al., 2006; Delgado, 2006):

$$\frac{D_l}{Da} = \frac{1}{\tau^s} + \frac{1}{12} \frac{u(2R)}{Da}$$

(15)

where $\tau^s$ is the tortuosity of the packed bed column and $u$ is the average fluid velocity in the longitudinal direction. For a packed bed of spheres, we can set $\tau = \sqrt{2}$ (Lanfrey et al., 2010). Non-dimensionalizing Eq. (15) with the help of Table 3, we find

$$D_l^* = \frac{1}{\tau^s} + \frac{u^3}{6}$$

(16)

We will consider a three-component system, containing a majority inert solvent species $S$, a reactive species $A$ which can adsorb over the surface of the catalytic spheres and undergo a reaction, to produce product species $B$. For a linear reaction, the reaction rate constant will be $k_{ads}^R$. $A$, $B$ and $S$ only differ by a tag therefore we can make the simplifying assumption of constant diffusivity/displacement across the axial direction as long as $Re$ is not too high ($< 200$).

The reactor is full of solvent particles $S$ at the start of system $t = 0$. We maintain a constant pressure difference for all $t > 0$ and let the system attain mechanical equilibrium, similar to Section 3.3. This initial run time of the setup is time $t_{setup}$.

For all $t > t_{setup}$, the following two processes occur. Firstly, as solvent $S$ starts to move out of the system at $z = L_z$, it is made to
Consider a cross-sectional volume element of length $\Delta z$ along $z$ direction and $L_x, L_y$ along $x$ and $y$ direction respectively. For sufficiently high number of spheres in the model reactor, the total number of catalytic spheres in this volume element will be equal to $N_{A} = N_{0}\Delta z/L_z$. If the adsorption flux of $A$ on a single sphere inside this volume element is $k_{ads}^{A} \gamma_{A}(t) A_{cat}$, the adsorption flux of $A$ in the volume element will be $k_{ads}^{A} \gamma_{A}(t) A_{cat} N_{0} \Delta z/L_z$. The volumetric adsorption flux $k_{ads}^{A} \gamma_{A}(t)$ will be

$$k_{ads}^{A} \gamma_{A}(t) = \frac{k_{ads}^{A} A_{cat} N_{0} \Delta z}{L_x L_y L_z} \gamma_{A}(t)$$ (18)

$L_x L_y L_z$ is the volume of the model reactor, $A_{cat} = 4\pi R^2$ and $N_0$, the total number of catalytic spheres, can be related to the total reactor volume as

$$\frac{4}{3} N_0 \pi R^3 = (1-\epsilon) L_x L_y L_z$$ (19)

Using Eqs. (18) and (19), we can write the volumetric reaction rate constant as

$$k = \frac{3}{4} \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (20)

and the dimensionless volumetric reaction rate constant is

$$k^* = \frac{3}{4} \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (21)

An analytical solution for Eq. (17) has been presented in Levenspiel (1962). The reactant concentration $\Psi(z) = \gamma_{A}(z)/\gamma_{A0}$ across the length of the reactor $\lambda = z/L_z$ is

$$\Psi = \frac{(1+q) \exp(\frac{1}{4} Pe_1 (1-\lambda)) - (1-q) \exp(-\frac{1}{4} Pe_1 (1-\lambda))}{(1+q)^2 \exp(\frac{1}{4} Pe_1) - (1-q)^2 \exp(-\frac{1}{4} Pe_1)}$$ (22)

where $q = \sqrt{1 + 4 Da_{cat}/Pe_1}$, $Pe_1 = \omega D_0^L/\nu^2$ is the dimensionless Peclet number relating the convection and dispersion time scales along the longitudinal length in the reactor. $Da_{cat} = k^* L_x^2/\nu^2$ is the Damköhler number relating the reaction and convection time scales along the longitudinal direction over a single sphere radius.

Fig. 9 plots the variation of reactant concentration along the reactor length for $Da_{cat} = 1.0$ and $Da_{cat} = 10.0$, respectively. For the present simulation framework, the reaction occurs between $z = L$ and $z = 0$ as species $A$. For more information about mass and momentum conservation when carrying out this step for different mass particles, see Appendix C. As species $A$ starts to fill the reactor, the reaction over the sphere starts with rate $k_{ads}^{A}$ producing species $B$. As has been discussed in the previous section, for a linear assumption of reaction, $B$ once formed will not desorb back on any catalytic sphere and eventually exit the system from $z = L_s$. For every particle $B$ leaving the system, a particle $A$ re-enters the system from $z = 0$. As the reaction progresses, the system will develop a distribution of $A, B$ along the reactor length. Any species that diffuses back and leaves the system at $z = 0$ will be reintroduced from the boundary at $z = L_s$ as the same species. However, such an implementation will also result in $B$ formed at the start of the reactor to re-enter the reactor from $z = L_s$ after diffusing backwards, affecting the output concentration of reactants or products.

To remove this effect, we define inert inlet and outlet regions of length $L$ near the boundaries in the $z$-direction whose dimension should be at least the minimum of $L_x, L_y$. Species $A$ will only react between $z = 0$ and $z = L - L_s$. This inlet region gives the product species $B$, formed just near the inlet, some space to diffuse back (without leaving the system) and eventually be convected away by the flow in $+z$ direction.

The simplifications involved in resolving the system into three primary time scales are the following. Firstly, the catalyst should equilibrate almost instantaneously, as compared to convective and reactive time scales, to avoid depletion region formation near the catalyst and also to ensure the reaction remains linear (Sengar et al., 2017). Secondly, the time taken for particles to disperse across one box-length in the transverse direction ($x, y$) should be comparable or more than the time taken for particles to convect across the reactor length in the flow direction.

The variation of the number density of reactant species $A$ along the $z$ direction can now be written as

$$\frac{D_x}{\Delta z^2} \frac{d^2 \gamma_{A}}{dz^2} - u \frac{d \gamma_{A}}{dz} + r_{A} = 0$$ (17)

where $u$ is the steady velocity of solvent particles $S$ at $t = t_{start}$ (under our assumptions, the system distribution of velocities of particles across the reactor for $t < t_{start}$ and $t > t_{finish}$ will be the same) and $r_{A}$ is the rate of consumption of $A$ per unit volume per unit time, $r_{A} = -k \gamma_{A}$. Note that the previously mentioned $k_{ads}$ differs from the current $k$.

We can derive its value as follows.

$$k_{ads}^{A} = \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (20)

and the dimensionless volumetric reaction rate constant is

$$k^* = \frac{3}{4} \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (21)

An analytical solution for Eq. (17) has been presented in Levenspiel (1962). The reactant concentration $\Psi(z) = \gamma_{A}(z)/\gamma_{A0}$ across the length of the reactor $\lambda = z/L_z$ is

$$\Psi = \frac{(1+q) \exp(\frac{1}{4} Pe_1 (1-\lambda)) - (1-q) \exp(-\frac{1}{4} Pe_1 (1-\lambda))}{(1+q)^2 \exp(\frac{1}{4} Pe_1) - (1-q)^2 \exp(-\frac{1}{4} Pe_1)}$$ (22)

where $q = \sqrt{1 + 4 Da_{cat}/Pe_1}$, $Pe_1 = \omega D_0^L/\nu^2$ is the dimensionless Peclet number relating the convection and dispersion time scales along the longitudinal length in the reactor. $Da_{cat} = k^* L_x^2/\nu^2$ is the Damköhler number relating the reaction and convection time scales along the longitudinal direction over a single sphere radius.

Consider a cross-sectional volume element of length $\Delta z$ along $z$ direction and $L_x, L_y$ along $x$ and $y$ direction respectively. For sufficiently high number of spheres in the model reactor, the total number of catalytic spheres in this volume element will be equal to $N_{A} = N_{0}\Delta z/L_z$. If the adsorption flux of $A$ on a single sphere inside this volume element is $k_{ads}^{A}(t) A_{cat}$, the adsorption flux of $A$ in the volume element will be $k_{ads}^{A}(t) A_{cat} N_{0} \Delta z/L_z$. The volumetric adsorption flux $k_{ads}^{A}(t)$ will be

$$k_{ads}^{A}(t) = \frac{k_{ads}^{A} A_{cat} N_{0} \Delta z}{L_x L_y L_z} \gamma_{A}(t)$$ (18)

$L_x L_y L_z$ is the volume of the model reactor, $A_{cat} = 4\pi R^2$ and $N_0$, the total number of catalytic spheres, can be related to the total reactor volume as

$$\frac{4}{3} N_0 \pi R^3 = (1-\epsilon) L_x L_y L_z$$ (19)

Using Eqs. (18) and (19), we can write the volumetric reaction rate constant as

$$k = \frac{3}{4} \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (20)

and the dimensionless volumetric reaction rate constant is

$$k^* = \frac{3}{4} \frac{k_{ads}^{A}(1-\epsilon)}{R}$$ (21)
and \( z = L_z - L \), so the effective reactor length should be \( L'_e = L_z - 2L \). However, figures are provided by mapping the modified reactor length, \( L'_e \) over the real reactor length \( L_z \) make it convenient to analyze.

All the simulations in Fig. 9 have the following fixed dimensionless parameters (refer to Section 3.1 and Table 2):

\[
\frac{\gamma_{\text{eq}}}{D_a} = 30000; \Delta t^*_a = 0.07; m^*_c = 0.0045; N_s = 1000; L'_e = L_z = 10; L'_o = 100; \epsilon = 0.5.
\]

The base units are \( R = 4.47t_0; \eta/A = 152.77t_0/(\varepsilon A \eta_0); D_A = 2.07/\eta_0 \).

For an SRD fluid, the corresponding Schmidt number, \( S_c \) under the given \( \gamma_{\text{eq}}/D_a \), \( m^*_c \) represents a gas-like fluid behavior (\( H_{g2} \)). All simulations predict Reynolds number \( Re < 100 \). Table 4 highlights the variables parameters used to obtain the different \( D_A \) and \( P_e \).

Table 4

<table>
<thead>
<tr>
<th>Dimensionless numbers</th>
<th>Parameter set</th>
</tr>
</thead>
<tbody>
<tr>
<td>( D_A = 1, P_e = 0.1, Re = 0.087 )</td>
<td>( g^* = 0.05 )</td>
</tr>
<tr>
<td>( D_A = 1, P_e = 0.1, Re = 0.43 )</td>
<td>( g^* = 0.27 )</td>
</tr>
<tr>
<td>( D_A = 1, P_e = 10, Re = 3.2 )</td>
<td>( g^* = 2.0 )</td>
</tr>
<tr>
<td>( D_A = 10, P_e = 1, Re = 0.87 )</td>
<td>( g^* = 0.55 )</td>
</tr>
<tr>
<td>( D_A = 10, P_e = 10, Re = 4.9 )</td>
<td>( g^* = 3.1 )</td>
</tr>
<tr>
<td>( D_A = 10, P_e = 100, Re = 21 )</td>
<td>( g^* = 13.8 )</td>
</tr>
</tbody>
</table>

4. Nonlinear reactor models

In Section 3, we have validated the SRD model for first order reaction system when the mass diffusivity of the reactants equals the mass diffusivity of the products. In this section, we consider a model reactor as described in Section 3.5 and present two situations where (a) reaction order is nonlinear, and (b) product diffusivity is different from reactant diffusivity.

4.1. Binary nonlinear reaction system

In our previous study (Sengar et al., 2017), we have described the need for Langmuir-Hinshelwood type of kinetics to model a simple reaction system like \( A \rightarrow B \).

\[
A \overset{k_{\text{ads}}}{\rightarrow} A + \overset{k_{\text{des}}}{\rightarrow} B \overset{k_B}{\rightarrow} B
\]

The rate equations for this model are:

\[
\frac{dA}{dt} = k_{\text{ads}}A - k_{\text{des}}A - k_A k_{\text{cat}}B
\]

\[
\frac{dB}{dt} = k_B k_{\text{cat}}A - k_B k_{\text{cat}}B - k_B k_{\text{cat}}^2 A
\]

Under a pseudo-steady state assumption, a catalytic surface is locally assumed to be equilibrated. The steady-state fractional occupations of \( A \) and \( B \), respectively, are:

\[
\theta_A_{\text{eq}} = \frac{k_{\text{eq}} A / k_{\text{ads}} A / k_{\text{cat}} A}{1 + k_{\text{eq}} A / k_{\text{ads}} A / k_{\text{cat}} A + k_B k_{\text{cat}} B / k_B k_{\text{cat}} A}
\]

\[
\theta_B_{\text{eq}} = \frac{k_{\text{eq}} B / k_{\text{ads}} B / k_{\text{cat}} B}{1 + k_{\text{eq}} B / k_{\text{ads}} B / k_{\text{cat}} B + k_B k_{\text{cat}} A / k_B k_{\text{cat}} B}
\]

where \( k_{\text{eq}} = k_{\text{ads}} / N_{\text{cat}} k_{\text{des}} \) and \( \gamma_i \) represents the number density of species \( i \) locally over the catalytic strip.

The rate of disappearance of \( A \) at this pseudo-steady state is \( k_{\text{ads}} A / \theta_A_{\text{eq}} - k_{\text{des}} N_{\text{cat}} \theta_A - k_{\text{cat}} B \theta_B_{\text{eq}} \) which is equal to the rate of generation of \( B \), \( -k_{\text{ads}} N_{\text{cat}} \theta_B_{\text{eq}} + k_{\text{cat}} A \theta_A_{\text{eq}} \) which in turn is equal to the rate of surface reaction \( k_B k_{\text{cat}} A / N_{\text{cat}} \).

In Section 3.5, the kinetics equations were written under the condition \( p_{\text{ads}} = 0.0, p_B = 1.0, p_{\text{des}} = 1.0, p_{\text{ads}} = 0.0 \). The rate of disappearance of \( A \) then becomes \( k_{\text{ads}} A / k_{\text{ads}} A / k_{\text{cat}} A \) for any \( p_{\text{ads}} \). This linearity assumption generates a Damkohler number which is a constant across the reactor system. This might not be the case when the probability parameters mentioned above change. We will now use the probability parameters defined to generate Fig. 9 as the base case and compare the scenarios when \( p_{\text{ads}}^2 \) and \( p_{\text{ads}}^3 \) changes.

The setup of the reactor is the same as described in Section 3.5. Previously, the equilibration time of the catalyst was zero. This enabled us to assume a steady state concentration of reactants and products across the reactor already after a time \( t_{\text{eq}} = t_{\text{reactor}} \), where \( t_{\text{reactor}} = L' / u \), the typical residence time of particles inside the reactor.

Now, we also have to account for the equilibration lifetime, \( t_{\text{eq}} \), of each catalytic surface present in our system. We can, approximately, determine this lifetime as follows. When \( k_B = 0 \), the equilibration lifetime is of the order of \( \text{Max}(k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A) \纤维化 | (k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A)^{-1} \纤维化 | (k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A)^{-1} \). Similarly, when \( k_B = \infty \), the equilibration lifetime is of the order of \( \text{Min}(k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A) \纤维化 | (k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A)^{-1} \纤维化 | (k_{\text{ads}}/\gamma_i, k_{\text{des}}/k_{\text{cat}} A)^{-1} \). Using this analysis, we can always pre-determine a suitable time after which every catalyst can be assumed to be equilibrated.

Now, at time \( t = t_{\text{reactor}} + t_{\text{eq}} \), we analyse the steady state concentration profiles of the reactants and the products.

Fig. 10 shows the deviation from the linear assumption for 2 cases, \( D_A = 1.0, P_e = 10.0 \) and \( D_A = 10.0, P_e = 10.0 \) (shown by thick black lines in both figures) as probability parameters change.

In Fig. 10(a), \( p_{\text{ads}}^2 \) is fixed at 0.01 and \( p_{\text{ads}}^3 \) is varied from 0 to 0.9. \( p_{\text{ads}}^2 \) is kept constant (see Table 4). For \( p_{\text{ads}} = 0.0 \), the adsorption of \( A \) is still the slowest process and the curve almost overlaps with the base case. As \( p_{\text{ads}}^3 \) increases, the adsorption flux of product \( B \) increases along the length of the reactor. This, in turn, shifts the fractional coverage over the catalyst in favor of \( B \). The effect is visible in the concentration of \( A \) along the reactor length since more unreacted \( A \) is now convected along the longitudinal direction.

For Fig. 10(b), \( p_{\text{ads}}^2 \) is fixed at 0.5 and \( p_{\text{ads}}^3 \) is varied from 0 to 0.9. The higher desorption probability of \( B \) allows more reactant species \( A \) to adsorb, and consecutively react to form product \( B \).

4.2. Unequal species diffusivity

In the previous section, a nonlinear reaction was analyzed while ensuring that the diffusivity of the mixture in the bulk of the
reactor system always remains the same. Now, we will study the system when the reactant and product species have different diffusivity values.

With the same reactor model as defined in Section 3.5, we initiate a reaction \( A \rightarrow nB \). Consequently, \( m_A = m_A/n \) where \( n > 1 \) or \( n < 1 \). For an SRD fluid, changing the mass of particles inversely affects their diffusivity, Eq. (1). Since we are mainly interested to measure the effects of bulk diffusivity in the mixture, a first-order reaction is considered \((P_{\text{des}}^A > 0, P_{\text{des}}^B = 0.0, P_A^L = 1.0, P_{\text{des}}^B = 0.0)\). We consider the case \( D_A = 1.0, P_{\text{L}} = 10.0 \) with the same simulation parameters as used to generate the corresponding curve in Fig. 9. The rules for conversion of \( A \) to \( B \) on the catalytic surface, when mass of the two species is different, is mentioned in Appendix C. Appendix C also mentions the conservation rules applied to handle boundary conditions when particles \( A \) or \( B \) leave the periodic boundaries (as defined in Section 3.5).

When \( m_A < m_A \), the mutual diffusivity of the mixture is higher than the self-diffusivity of reactant \( A \). This translates to a lower frictional force experienced by \( A \) while traversing the reactor, thereby, reducing the residence time of species \( A \) between a volume element \( z \) and \( z + dz \). This effect is visible in Fig. 11 where for \( m_A > m_A \), more unreacted \( A \) is present along the reactor length as compared to the situation with \( m_A = m_A \). For \( m_A > m_A \), the effect is reversed and the mobility of \( A \) decreases.

It should also be noted that the above reaction system does not conserve the number of particles in the system which is a source of an internal pressure difference generated along the longitudinal direction. This pressure adds up to the external pressure difference when \( m_A < m_A(n > 1) \). Conversely, when \( m_A > m_A(n < 1) \), this internal pressure goes against the external pressure. This effect itself can be incorporated into the Maxwell-Stefan diffusivity between \( A \) and \( B \). Further information is provided in Appendix D.

5. Discussion and conclusion

We have presented here a multiscale modeling technique that is able to accurately resolve different processes occurring on time scales that differ on orders of magnitude. We have highlighted the basic modeling techniques involved in modeling a simple reactor system (involving convection, diffusion and reaction) using SRD. The analysis was first carried out for a linear reaction assumption, followed by the incorporation of a reaction cycle following Langmuir Hinshelwood kinetics. The ability of the SRD model to also account for multicomponent diffusion opens up a whole new domain of complex reaction system that can be modeled.

To model a real chemical reactor, the number of processes occurring at different time scales might be much larger than 3. Even if kinetics is given by the simple Langmuir Hinshelwood model, intermediary processes like adsorption, desorption, surface reaction become important. As the number of participating components increases, the system complexity increases and resolution of each and every process becomes challenging.

The generation of different product species in such a multicomponent nonlinear scheme will directly affect the behavior of the bulk fluid as well. Different species in the bulk will have different self-diffusivity. The pairwise interaction between species can be incorporated in a mutual diffusivity expression, which has been presented in this work, using SRD, for a binary mixture but can be extended for any \( n \) component solution. Unequal moles of reactants and products might lead to an internal pressure force originating from or away from the reaction site, the contribution of which has been incorporated in the Maxwell-Stefan diffusion coefficients, see Appendix D.
Another time scale of importance is the equilibration time of the catalytic strip. For all cases with linear reaction assumption, we assumed that the catalyst equilibrates or achieves a pseudo steady state instantly. When this is not observed, regions of low density (depletion regions) may form near the catalytic strip which can be a source of another driving force (only for higher values of Sc). When reactions occur inside complex geometries, a new time scale will be present which relates to the time taken for the reactant fluid to move through a pore towards the reaction site (pore diffusion).

The combined effect of the above described processes is reflected in the output concentration of any species from the reactor system. A macroscale reactor might involve length scales much larger than the length scales of individual processes which will result in the incorporation of another time scale in the system.

With a large number of processes to resolve at the hydrodynamic and reaction level, SRD has been shown to be quite effective in resolving particle–particle interactions quickly for a large number of components. The computational efficiency of the algorithm therefore makes it easy to couple this scheme with any reaction-based scheme for complex geometries, as long as the reaction steps conform to the mean-field assumption.

We would like to end the discussion for this work by guiding the readers interested in further resolving the system of interest. For reaction schemes where a mean-field assumption does not hold true (island formation or phase transition at the catalytic sites), a stochastic treatment to the individual reaction steps has to be done which will add up to the simulation time. For such complex reaction schemes, one may use Machine Learning to develop an algorithm that observes a stochastic reaction cycle. The algorithm is then able to predict the output fractional vacancies of different species based on input of (a) reaction rate constants at time \( t \), (b) fractional vacancy of surface sites at time \( t \), and (c) the hydrodynamic time step \( \Delta t_c \). This will be the topic of our future work.

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### Appendix A. Reactions on a sphere

To model reactions over a more complex geometry like a sphere, the lattice cells in an SRD fluid near boundaries should be updated and the catalytic surface needs to be modified.

To simulate a catalytic sphere with periodic boundaries, the surface of the sphere will be no-slip and reactive. As mentioned before, a grid-shift procedure helps achieve Galilean-invariance in the system where lattices are shifted in space before collision by a random number between 0 and \( \varphi \) in each Cartesian direction. Due to this, all lattice cells inside the sphere which are at a distance \( \sqrt{3} \varphi \) from the surface need to be checked for ghost–cell boundary conditions. For all cells near boundaries, a modified ghost boundary cell approach was used that conserves not the number density of particles but the average centre of mass momentum of particles in the bulk (Sengar et al., 2017).

In presence of convection, the pressure over the sphere will be different at different positions. A single averaged catalytic surface will then give wrong predictions. To tackle this, the surface of the sphere can be divided into \( \pi \) equal parts along any Cartesian axis will result in an equal surface area of each portion.

### Appendix B. Coupling reactions with hydrodynamics

In our previous work, Sengar et al. (2017), we outlined the need to rewrite the rate equations to simultaneously simulate adsorption, desorption and reaction steps. Let’s assume adsorption of \( A \) on a catalytic strip, the rate equations when coupled with the fluid in the bulk can be written as

\[
\frac{\Delta \rho_{A}}{\Delta t_c} = k_1 \theta - k_2 \theta_A - k_3 \theta_A + k_4 \theta_B \quad (B.1a)
\]

\[
\frac{\Delta \rho_{B}}{\Delta t_c} = k_2 \theta + k_3 \theta_A - k_4 \theta_B - k_5 \theta_B \quad (B.1b)
\]

where \( \Delta t_c \) is the hydrodynamic time step of the fluid and \( \theta = 1 - \theta_A - \theta_B \) by the conservation of surface sites. The hydrodynamic time step is set such that the adsorption rates \( k_1 \) and \( k_6 \) (of \( A \) and \( B \) respectively) can be assumed constant over this time period. If this is not the case, \( \Delta t_c \) needs to be reduced further. Sequential treatment of the reaction steps might lead to overestimation or underestimation of a surface species concentration. This occurs if the change produced by any intermediary step in the reaction cycle within one time step is much higher as compared to the other steps (which generally occurs when reaction occurs much faster than diffusion). For an accurate simultaneous representation of the reaction system, one solution is to reduce the time step \( \Delta t_c \). This however increases the computational time. In this work, we use a technique that modifies the instantaneous rates at every time step to avoid any error.

#### B.1. Adaptive rate method

We define the fastest intermediary step between time interval \( t \) and \( t + \Delta t_c \) as \( \psi(t) = \max(\Delta k \theta) \). The method involves rewriting the rate equations with respect to the process \( k \) acting on \( \theta \). The modified rate equations then become

\[
\frac{\Delta \theta_A}{\Delta t_c} = \psi(t)(k_1 \theta - k_2 \theta_A - k_3 \theta_A + k_4 \theta_B) \quad (B.2a)
\]

\[
\frac{\Delta \theta_B}{\Delta t_c} = \psi(t)(k_2 \theta + k_3 \theta_A - k_4 \theta_B - k_5 \theta_B) \quad (B.2b)
\]

where \( k_i = k_i(\psi(t)) \). For every time step \( t \), the rate constants of all intermediary steps are modified by \( \psi(t) \). The reaction cycle is initiated by the first occurrence of \( \psi(t) \) following which a sequential treatment of the other processes with modified rate occurs. This is repeated until all the occurrences of \( \psi(t) \) are exhausted.

As discussed in Sengar et al. (2017), the adsorption probabilities derived from the adsorption rates remain independent of the steaming step of the fluid, \( \rho_{ads} = c k_{ads} \), where \( c \) is a constant that depends on simulation parameters and fluid properties (Table A1). The modified adsorption probability for the scheme discussed becomes \( \rho_{ads} = c k_{ads} - \psi(t) \). The probability of desorption and reaction/\(-\)conversion to other surface species is linked to the rate constants as \( \rho_{des} = -\log(1 - k_{des} \Delta t_c) \). The modified probability then becomes \( \rho_{des} = -\log(1 - k_{des} \Delta t_c) \).

A reaction cycle involving nonlinear reaction pathways, for example \( k \theta \theta \), can be easily modified using the above approach. The modified rate can be written as \( k = k(\theta(t)) \), and a probability parameter can be obtained for the modified rate constant. Since \( k \leq k \), the linearization approach used to model with scheme Eq. (B.2) will always be numerically more accurate that sequential treatment in Eq. (B.1).
B.2. Matrix formulation

The second method involves writing the rate of change in surface species involved in matrix formulation and solving it numerically over the streaming time step of the fluid.

$$\begin{bmatrix}
\Delta\theta/\Delta t_c \\
\Delta\theta_i/\Delta t_c \\
\Delta\theta_j/\Delta t_c \\
\end{bmatrix} = \begin{bmatrix}
-k_1 - k_6 & k_2 & k_3 \\
k_1 & -k_2 - k_3 & k_4 \\
k_6 & k_3 & -k_4 - k_5 \\
\end{bmatrix} \begin{bmatrix}
\theta \\
\theta_i \\
\theta_j \\
\end{bmatrix}$$

The above matrix calculates the surface species concentration at the end of time step. With this information, we can numerically predict the total number of adsorptions, desorptions and reactions to occur over the catalytic strip between time \( t \) and \( t + \Delta t_c \). If \( \Delta \theta_j \) represents the number of occurrences of reaction step \( k \), acting over \( \theta_j \) between time \( t \) and \( t + \Delta t_c \), the modified probability for any process becomes \( \Delta \theta_j/\theta_j(t) \). If \( \Delta \theta_{i,j} \) is the numerically calculated number of adsorptions of species \( A \) between \( t \) and \( t + \Delta t_c \), the modified probability of adsorption for time \( t \) is \( \Delta \theta_{i,j}/\theta_j(t) \).

For a nonlinear rates like \( \theta_i/\theta_j(t) \) or \( \theta_i/\theta_j(t) \) depending on which surface species involved in matrix formulation and solving it numerically over the streaming time step of the fluid.

Appendix C. Mass and momentum conservation

The modeled system involving a reaction has two primary regions where SRD particles change to a different species while conserving mass and/or momentum. At the catalytic sites where the reaction occurs, a change in number of product particles with mass conservation occurs. Secondly, at open boundaries if particles change nature (from products to reactants or vice versa), mass and momentum conservation rules need to be applied.

C.1. Adding particles

We explore the mass and momentum conservation rules for the case where addition of \( n > 1 \) particles of different mass and velocity leads to the formation of a single new particle.

By mass conservation,

$$m' = \sum_{i=1}^{n} m_i$$

where \( m_i \) is the mass of initial particles and \( m' \) is the mass of the new particle.

Let the velocity of individual particles be \( v_i + dv_i \) where \( v_i \) and \( dv_i \) are the first and second moment of velocity respectively.

To generate \( v' + dv' \) as the velocity of new particle, we perform momentum conservation, \( m'v' = \sum_{i=1}^{n} m_i v_i \) which gives the first moment of velocity of the new particle as

$$v' = \frac{\sum_{i=1}^{n} m_i v_i}{\sum_{i=1}^{n} m_i}$$

(C.2)

Since the temperature of the ensemble of SRD particles is stored in the velocity fluctuations of each particle, we need to generate the second moment of velocity of new particle such that the total energy of the system is preserved. The total energy of all initial particles is:

$$E_{\text{initial}} = \sum_{i=1}^{n} m_i v_i^2$$

$$E_{\text{initial}} = \sum_{i=1}^{n} m_i \frac{k_B T}{m_i} = n k_B T$$

(C.3a)

By conservation of energy, \( E_{\text{initial}} = E_{\text{final}} \).

$$E_{\text{final}} = m' v'^2$$

$$m' v'^2 = nk_B T$$

$$dv' = \sqrt{\frac{k_B T}{m'/n}}$$

(C.4b)

The above equation implies that the new particle formed should be given a random velocity fluctuation from a normal distribution with standard deviation \( \sqrt{\frac{k_B T}{m'/n}} \) to conserve total energy of the system. Note that \( m'/n \) is just the average masses of the initial particles.

C.2. Splitting particles

Splitting a single particle, with mass \( m' \), velocity \( v' \) and velocity fluctuation \( dv' \), requires the same approach as above: conservation of mass, momentum and velocity fluctuations (leading to energy conservation). If \( n > 1 \) new particles are created with mass \( m_i \), by mass conservation,

$$m' = \sum_{i=1}^{n} m_i$$

(C.5)

If the velocity of \( i \)th particle is \( v_i + dv_i \), where \( v_i \) is the mean velocity and \( dv_i \) is the velocity fluctuation, by momentum conservation,

$$m' v' = \sum_{i=1}^{n} m_i v_i$$

(C.6)

The energy of the initial particle \( E_{\text{initial}} = k_B T \). The energy of final particles \( E_{\text{final}} \) (which is equal to \( E_{\text{initial}} \)) is

$$E_{\text{final}} = \sum_{i=1}^{n} m_i v_i^2 = k_B T$$

(C.7)

A solution to the above equation is \( dv_i^2 = k_B T/m_i \). Therefore, we can generate a velocity fluctuation for all the new particles from a normal distribution of standard deviation \( \sqrt{k_B T/m_i} \). The implication of this is that new particles formed are at a lower temperature than the initial particle by a fraction \( \sqrt{m_i/m'} \). Within a few collisional steps, these new particles will equilibrate according to their Maxwellian distribution with a second moment of velocity \( k_B T/m_i \) (if the environment around this particle consists majorly of particles of mass \( m_i \)).

C.3. Updating particle positions

In case of adding particles, only particles in the vicinity of half a cubic cell length \( \delta x/2 \) are combined together since a velocity gradient present along any direction might lead to particles adding from regions of very different velocities. Following this, the coordinates of the new particle should be the coordinates of the center of mass of all the other particles.

$$m'r' = \sum_{i=1}^{n} m_i r_i$$

(C.8)

When splitting particles, since the velocity of new particles is the same as the initial particle, new particles can be positioned at the same location as the initial particle. This creates a region with high number density and low individual temperature of the formed particles momentarily until they distribute their mass in the bulk and attain the same temperature as the bulk.
Appendix D. Maxwell-Stefan diffusion: unequal consumption and generation over catalyst

Chemical reactions might involve change in pressure at the reaction site as compared to the bulk if total moles of gaseous product species is different than total moles of reactant product species. A pressure difference towards or away from the catalyst will affect the diffusivity of whole region. To demonstrate this aspect, we use the simulation method used in Sengar et al. (2017).

To summarize, a reaction $A \rightarrow nB$ occurs at a catalytic strip, placed along the $y$ axis of a 3D box $(L_x \times L_y \times L_z)$ with walls along $y$ direction and open boundaries along the $x$ and $z$ direction. The reaction leads to change in number of particles and mass of particles, leading to change in diffusivity of product particles, Eq. (1). The catalytic strip with dimensions $L_x \times L_{cat}$ is placed at the center of the box along the $y = 0$ wall such that length of strip in $z$ direction is small as compared to the length of the box, $L_z$. At $z = 0$ and $z = L_z$ product particles $B$ are removed following the conservation principles defined in Appendix C. At $x = 0$ and $x = L_x$, particles are re-entered into the system from the other end without any change in their physical property. There is no convection, only diffusion leading to transient velocity gradients that disappear at steady state. The net force on a particle of reactant or product species $i$ can be written as Taylor and Krishna (1993)

$$RTd_i = \nabla \cdot \mu_i - F_i$$  \hspace{1cm} (D.1)

where $\nabla \cdot \mu_i$ is the gradient in chemical potential at constant temperature and $F_i$ is the external force on the system. For the present case, the external force is a transient force that originates because of change in number of moles during a reaction. This leads to a pressure gradient in the system. At steady state the, external force is a transient force that originates because of change in number of moles during a reaction. This leads to a pressure gradient in the system. At steady state the, external force becomes zero but the reaction is now supported by the pressure gradient in the system. It has been shown that under forces leading to pressure gradients in the system, the force equation can be rewritten as Taylor and Krishna (1993)

$$d_i = \nabla \cdot \mu_i + \left( \frac{\mu_i}{\nabla P} - \frac{1}{P} \right) \left( J_i - \frac{\mu_i}{\nabla P} \right)$$  \hspace{1cm} (D.2)

where $\mu_i$, $\omega_i$, $\gamma_i$ are the number fraction, mass fraction and number density of species $i$ and $\nabla P$ is the gradient in total pressure in the system. At steady state, the force per particle $d_i$, in a system with $n$ different species, can be equated to the diffusive flux according to the relation:

$$\gamma_i d_i = -[B] J_i$$  \hspace{1cm} (D.3)

where $\gamma_i$ is the total number density of particles of all species, $d_i$ is the column matrix containing $d_i$, $J_i$ is the column matrix containing diffusive flux $J_i$, and $B$ is the matrix with coefficients given by

$$B_{ii} = \frac{\gamma_i}{D_{in}} + \frac{1}{n} \sum_{k \neq i} \frac{\gamma_k}{D_{ik}}$$  \hspace{1cm} (D.4a)

$$B_{ij} = -\gamma_i \left( \frac{1}{D_{ij}} - \frac{1}{D_{in}} \right)$$  \hspace{1cm} (D.4b)

where $D_{ij}$ is the Maxwell-Stefan diffusion coefficient for species $i$ and $j$ or the inverse drag-coefficient experienced by species $i$ under influence of species $j$. For a binary mixture, Eq. (D.3) can be written as

$$J_i = -\gamma_i D_{12} d_i$$  \hspace{1cm} (D.5)

For the model reaction system involving two components, at steady state, $F_i = 0$, Eq. (D.2) can be used in Eq. (D.5) to get

$$J_i = -\gamma_i D_{12} \nabla X_i$$  \hspace{1cm} (D.6)

The diffusive flux ($J$) of reactant and product species can also be calculated at steady state of the simulation by looking at the rate of disappearance of reactant species or rate of generation of product species per square meter of the catalytic strip per unit time step which is a constant determined by the reaction rate parameters and model initialization.

When the difference in moles of product and reactant species due to reaction is small, a pressure gradient is not present and Eq. (D.6) can be written as

$$J_i = -\gamma_i D_{12} \nabla X_i$$  \hspace{1cm} (D.7)

Appendix E. Computational setup

An in-house code has been developed using which all the above simulations have been performed. The present method development is a continuation of our previous work (Sengar et al., 2017) where a reaction-diffusion system was built. The present code has been developed using C and all libraries and functions have been developed within the group. The simulations run on a single core of a CPU and there is scope for both optimization and parallelization. The processor is an Intel®Xeon®E5-1650 v2 with 3.5 Ghz clock speed and 12 MB cache.

A typical simulation contains $N_s$ spheres in a 3D box with length $(L_x, L_y, L_z)$. Any particle undergoes the following operations: Streaming, intra-cell collision, reflection, reaction (adsorption, desorption and surface reactions).

Streaming and collision times of particles are of the same order of magnitude and scale as $O(\gamma t_{cat}/L_{cat})$ where $\gamma$ is the typical number density of particles in the system.

To make reflections efficient, we have developed the following methodology. For each sphere in the reaction system, the lattice cells that lie inside the sphere $i$ are tagged $(0,i)$ and the ones that partially lie inside the spheres are tagged $(1,i)$. This ensures an efficient calculation to predict which sphere will particles interact based on their initial and final position (unreflected) positions (a fraction of particles that will not reflect will also get a tag $(1,i)$). Having done that, the higher the value of $\epsilon$, the closer spheres will be and higher will the chance of an SRD particle to collide with the sphere will be. For increasing $\epsilon$, we ensure that particles do not end up suffering multiple collisions between 2 spheres and we limit the number of such reflections to 3. Reflections scale up approximately as $O(\gamma L_{cat} L_{cat} / (1 - \epsilon)^2)$.

Reactions occur per single hydrodynamic time step of bulk fluid. We have defined two methodologies to perform reaction in Appendix B. In the adaptive rate methodology, we propose to perform all surface reaction operations for every instance of the fastest process. For reaction system, $O(\gamma_{avg} / A_{cat} N_{cat})$ operations occur for each hydrodynamic time step where $\gamma_{avg}$ is the average number density of SRD particles interacting with catalytic strip. Using the matrix formulation, the order per time step $\Delta t$, will be $O(\gamma_{avg} / A_{cat} n_{cat} / h)$ where $h$ is the time-step of the runge-kutta solver used and $n$ is the number of different surface species.

The complete runtime of a simulation is $t = t_{start} + t_{reactor} + t_{eq}$. $t_{start}$ is the time taken to achieve mechanical equilibrium where the mean velocity across the reactor becomes fixed. Until $t = t_{start}$, only streaming, collision and reflection occur. For $t_{reactor} \approx L_v / u$, streaming, collision and reflection occur. For $t_{eq}$, the equilibration time of the catalyst, again all the processes occur.
For a linear process, \( t_{eq} = 0 \) and the simulation time generally scales because of reflections. We can also assume \( t_{\text{start}} \approx t_{\text{reactor}} \). The simulation time will scale linearly on \( L_z / u \). For \( Da = 1, Pe_L = 0.1, Re = 0.087 \), the simulation time was approximately 100,000 s (about a day). For \( Da = 10, Pe_L = 100, Re = 21 \), the simulation time was approximately 4000 s.

For nonlinear processes, the processes confined within \( L_z \) dominate the system. It took approximately 100,000 s to obtain each curve of Fig. 10 (using matrix formulation). (For the same scenario in linear reactions, it took about 20,000 s per simulation) (See Table E.5).

References
