Simulation and control of the oxidation of sulfur dioxide in a micro-structured reactor \star

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Abstract: For producing sulfur trioxide, sulfur dioxide is oxidized to sulfur trioxide in the presence of a catalyst. Industrially, sulfur trioxide is produced by the contact process which consists of several catalytic beds and inter-cooling stages between them. This process is highly energy intense and inflexible due to long periods of start-up and shut-down caused by the thermal inertia. In order to improve the flexibility of the production process, a new approach for producing sulfur trioxide is currently being investigated. A micro-structured reactor has been constructed by the Karlsruhe Institute for Technology (KIT) which consists of only one cooling passage. In our work, a three dimensional dynamic reactor model was developed that describes the dynamic behavior of the micro-structured reactor. This model was used to simulate the efficiency and the distribution of temperature and reactants in the system and in particular for investigating the start-up of the reactor. A control strategy was developed and tested in simulations in order to react on disturbances and to improve the start-up time. The simulations showed that a conversion of approx. 98.5 % can be reached and by a suitable control strategy the start up time can be improved significantly.

Keywords: Micro-structured reactor, start-up, SO_3, gPROMS, heat transfer, contact process, PI controller

1. INTRODUCTION

Sulfur trioxide (SO_3) is an important raw material for many processes in the chemical industry. It is used for the sulfonation of organic compounds, in the production of detergents and as a raw material for chlorosulfuric acid and sulfuric acid, for flue gas conditioning, in coal fired power plants to improve dust removal and for many other applications. Today, SO₃ is mainly obtained by producing SO_2 via the combustion of elemental sulfur which is then oxidized to sulfur trioxide. The conversion of SO_3 from SO_2 is an exothermic and equilibrium controlled reaction. The equilibrium is such that higher temperatures are desired at the entry of the reactor to speed up the reaction and lower temperatures are used towards the end of the reactor to reach almost full conversion (see figure 1 B). Figure 1 A shows the principle how SO_3 is produced industrially where SO_2 is oxidized to SO_3 in a series of catalytic beds. The oxidation beds are separated by intercooling stages that lower the temperature of the mixture after its adiabatic rise in each catalytic bed. This concept for the production of sulfur trioxide has been used since the 1930s (Jähnisch (2004)).

A new reactor for the production of sulfur trioxide has been developed by the Karlsruhe Institute for Technology (KIT) in cooperation with the Institute of Chemical Process Fundamentals at the Academy of Sciences of the Czech Republic (UCHP) and Procter & Gamble (P&G) within the European research project IMPULSE and F3 (F3H (2009), imp (2009)) with the goal to increase the flexibility of the production of SO_3 and to circumvent the need of several inter-stage cooling steps which are highly energy consuming. Here, only one cooling passage is used to reduce the temperature towards the outlet of the reactor. In this contribution, we describe a dynamic distributed parameter model of the new micro-structured reactor that was developed within the project F3 to compute the temperature and conversion profiles within the reactor and to optimize the operating parameters (inflow temperature of the reactant and cooling temperature). Furthermore the model was used to determine a control strategy to react to disturbances and a start-up procedure. One important motivation for this work was to validate that the reactor can indeed be brought to the desired operating conditions which is not obvious because the heat that is needed to bring the reactor to the operating temperature must partly been provided by the reaction which does not start before a certain temperature is reached. The 3D simulation model was formulated and solved in gPROMS, a modelling software developed by Process System Enterprise (PSE),

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for controller tuning, the system identification toolbox in MATLAB was used.



Fig. 1. A. Schematic description of the contact process for the production of SO₃. B. Conversion versus temperature diagram for the production of SO₃ under optimized conditions.

2. THEORETICAL BACKGROUND

The reaction of sulfur dioxide to sulfur trioxide is,

$$\mathrm{SO}_2 + \frac{1}{2}\mathrm{O}_2 \rightleftharpoons \mathrm{SO}_3, \qquad \Delta H_r = -99\mathrm{kJ}$$

The reaction is taking place over a catalyst, where in the industry the most widely used catalyst is vanadium. The reactant SO_2 is a colorless, nonflammable, toxic gas with a boiling point of -10° C at 101.3 kPa. About 98 % of SO₂ is used for the production of sulfuric acid. The production of SO_2 itself is performed by combustion of sulfur with oxygen. The product SO_3 is mainly used as an intermediate for sulfuric acid production (Müller, 2000). According to the Le Chatelier principle, high pressures tend to favor the equilibrium of the reaction towards SO_3 since there are more moles on the left-hand-side of the reaction equation. Low temperatures move the equilibrium to the desired product but low temperatures cause the reaction to proceed slowly. Industrially, the oxidation of SO_2 takes place in catalytic beds under adiabatic conditions, where the temperature increases in each bed due to the exothermic reaction, and the intercooling stages lower the temperature to achieve full conversion. This principle is sketched in figure 1.

In this work, the oxidation takes place in a microstructured reactor with only one cooling stage at the end of the reactor.

2.1 Set up of the reactor

The reactor with its piping is shown in figure 2 A. For safety reasons and to reduce heat losses to the environment, the reactor is housed in a chamber which is filled with insulation material. The reacting mixture is preheated before it flows in catalytic micro-channels from left to right. At the section of the reactor near the outlet, the reacting fluid is cooled by air which flows orthogonally to the fluid. The cooling gas inlet temperature can be regulated by a heat exchanger. Thus, the reactor inside is divided into a reaction passage and a cooling passage, as shown in figure 2 **B**. In the reaction passage, the reactor consists of 213 stainless steel reaction foils and 213 heat exchange foils of the same material which are diffusion bounded. Each reaction foil contains 250 semicircular channels with a diameter of 300 μ m. The inner surface of the reaction channels is coated by the catalyst. The total

number of reaction channels is 53,250. The heat transfer foils contain also 250 channels with a diameter of 200 μ m. The cooling channels are perpendicular to the reaction channels (see figure 3 **B**). The geometry of the reactor is shown in figure 3 **A**. The length of the reactor is 320 mm, the width and the height are 200 mm. The outside of the reactor is equipped with 12 heating cartridges where six heating cartridges are attached to the top and six are attached to the bottom of the reactor. The heating cartridges can only be switched on or off and each cartridge has an electric power of 1 kW.



Fig. 2. A. Three dimensional isometric view on the reactor. The reactor is a chamber consisting of an inlet for the fluid and and an inlet for the cooling. B. Schematic representation of the reactor.



Fig. 3. A. Geometry of the reactor. B. Schematic overview of the arrangement of the different foils. The channels in the heat exchange foil are perpendicular to the reaction channels.

2.2 Process model

The mathematical model of the reactor consists of three abstract compartments, the process fluid, the metal structure and the heat exchange (cooling) fluid. There is no mass transfer between the three compartments but heat transfer between the metal structure and each of the two fluids. As the two fluids are separated from each other by the metal structure, there is no direct heat transfer between the two fluids and the only mass balance needed is the one for the process fluid. This means that the fluids are modeled as if they were distributed in the three dimensions of the reactor. The 3D-model hence does not result from the set-up of a large number of individual channel models which would lead to an enormous demand of computational power. The three components (process fluid, metal and heat exchange fluid) are discretized along the three coordinates and the dimensions of the channels are used to calculate the velocity of the fluid, the heat exchange area and the available area for the catalytic coating. Further assumptions which were made in the model are summarized below:

- The physical properties of both fluids (heat capacity, density) depend on temperature. For the metal, these properties are constant, except for the heat conductivity.
- The heat transfer coefficients are constant.

- The heat transfer by conduction is negligible in the fluids, but in the metal the heat conduction is taken into account.
- The pressure drop along each fluid direction is linear and independent of the fluid velocity.
- Ideal gas behavior.

The convention for the geometric coordinates is as follows:

- x-axis: This axis describes the direction of the flow of the heat transfer fluid. Its length corresponds to the width of the reactor.
- y-axis: This axis is orthogonal to the direction of both fluids. Only heat transfer by heat conduction takes place in this direction and the heat cartridges are placed on both edges.
- z-axis: This axis describes the direction of the flow of the process fluid. In z-direction, the reactor is divided into two sections: the reaction passage and the cooling passage (cf. figure 2).

In the following, the model equations are presented. They were derived from mole and energy balances over a differential volume applying physical laws.

Mole balance:

$$\frac{p}{R}\frac{\partial}{\partial t}\left(\frac{x_i}{T_f}\right) = -\frac{\partial}{\partial z}\left(\frac{pw_f x_i}{RT}\right) + \mu_i r_{rate} \frac{A_{xycat}}{A_{xy}} \quad i = SO_2, SO_3, O_2, N_2$$
(1)

Where

- p is the total pressure [Pa],
- R is the ideal gas constant [J/mol/K],
- x_i is the mole fraction of component *i*,
- T_f is the temperature of the fluid [K],
- μ_i is the estequiometric coefficient of component *i* $[kg/m^3],$
- w_f is the local velocity of the fluid [m/s],
- r_{rate} is the reaction rate of the fluid [mol/kg catalyst/s],
- $A_{xy,cat}$ is the cross sectional area of the catalyst $[m^2]$,
- A_{xy} is the cross sectional area of the process fluid channels $[m^2]$.

Process fluid energy balance:

$$\frac{\partial}{\partial t} \left(\rho_f c p_f T_f \right) = -\frac{\partial}{\partial z} \left(w_f \rho_f c p_f T_f \right) -
\Delta H_r r_{rate} \frac{A_{xy,cat}}{A_{xy}} + \frac{A_f k_f}{V_f} \left(T_m - T_f \right)$$
(2)

Where

- T_c is the temperature of the cooling fluid [K],
- T_f is the temperature of the fluid [K],
- cp_f is the specific heat capacity of the fluid [J/kg/K],
- ρ_f is the density of the coolant [kg/m³],
- ΔH_r is the heat of reaction [J/mol],
- V_f is the volume the fluid $[m^3]$,
- A_f is the heat exchange area for the fluid $[m^2]$,
- k_f is the heat transfer coefficient on the fluid side $[W/m^2/K].$

Heat transfer fluid energy balance:

$$\frac{\partial}{\partial t} \left(\rho_c c p_c T_c \right) = -\frac{\partial}{\partial x} \left(w_c \rho_c \right) + \frac{A_c k_c}{V_c} \left(T_m - T_c \right) \quad (3)$$

Where

- ρ_c is the density of the coolant [kg/m³],
- cp_c is the specific heat capacity of the cooling fluid [J/kg/K],
- w_c is the local velocity of the cooling fluid [m/s],
- A_c is the heat exchange area for the cooling fluid $[m^2]$, - k_c is the heat transfer coefficient on the coolant side $[W/m^2/K],$
- V_c is the volume the coolant [m³].

Metal energy balance:

$$\rho_m c p_m \frac{\partial T_m}{\partial t} = \frac{A_c k_c}{V_m} (T_c - T_m) + \frac{A_f k_f}{V_m} (T_f - T_m) + \frac{\partial^2 (K_{mx} T_m)}{\partial x^2} + \frac{\partial^2 (K_{my} T_m)}{\partial y^2} + \frac{\partial^2 (K_{mz} T_m)}{\partial z^2}$$
(4)

Where

- T_m is the temperature of the metal [K],
- ρ_m is the density of the metal [kg/m³],
- cp_m is the specific heat capacity of the metal[J/kg/K],
- V_m is the volume of the metal [m³], K_{mx} , K_{my} and K_{mz} is the conductivity along the x, y and z direction respectively [W/m/K].

2.3 Reaction mechanism

The used catalyst is 2% Pt/TiO₂. The parameters of the catalyst were fitted by a single channel model at the Institute of Chemical Process Fundamentals at the Academy of Sciences of the Czech Republic where the Karlsruher Institute of Technology provided the experimental data. The experimental data was obtained by a reactor in lab scale in which the catalyst was coated in a thin layer on microchannels. The channels were rectangular, 200 μ m width, 200 μ m high and 80 mm long. The experimental data was then fitted to the equations by the least squares method. The equations and the meaning of the kinetic parameters are summarized below.

Reaction rate:

$$r_{rate} = kp \left(x_{SO_2} \sqrt{x_{O_2} p} - \frac{x_{SO_3} \sqrt{p_{std}}}{K_{eq}} \right) \tag{5}$$

Reaction rate constant:

1

$$k = A \exp\left(\frac{-E_a}{RT}\right) \tag{6}$$

Equilibrium constant:

$$K_{eq} = \exp\left(\frac{11300}{T} - 10.68\right)$$
 (7)

Table 1. Kinetic parameters

Parameter	Meaning	Unit
A	Pre-exponential factor	mol/kg _{cat} /s/Pa ¹ .5
E_a	Activation energy	J/mol
$ ho_{cat}$	Catalyst density	kg/m^3

2.4 Boundary conditions

For the partial differential equations 1 - 4, nine boundary conditions are needed to fully specify the process. The boundary conditions are:

- 1 condition for the mole balance,
- 6 conditions for the metal heat balance,
- 1 condition for the process fluid heat balance,
- 1 condition for the heat exchange fluid heat balance.

The boundary conditions for the mole balance (equation 1) that is the inlet mole concentration equals the initial concentration. Similarly, the boundary conditions for the heat balance of the process fluid (equation 2) and of the heat transfer fluid (equation 3) are the inlet temperature of the fluid $(T_{fluid,in})$ or the inlet temperature of the cooling $(T_{cool,in})$. The boundary conditions for the metal heat balance (equation 4) are given on four sides of the metal block by the heat loss to the environment which is described by an insulation model. On the two sides where the heating cartridges are situated, the boundary conditions are either also the heat loss to the environment or the heat flow generated by the heating cartridges, depending on whether the heating cartridges are switched on or off. The corresponding equations are described below.

Heating cartridges The heating cartridges are added to the reactor to overcome the heat loss and to heat the reactor at the start up. The heating cartridges can only be switched on or off together and have an available power supply of 1 kW. The boundary condition for the heating cartridges can be stated as:

$$K_{metal}\frac{\partial T_m}{\partial y} = \frac{P}{A_{xy}}.$$
(8)

Heat loss and insulation In order to simulate the heat loss to the environment, the insulation of the reactor was modeled. On the boundary between the metal surface of the reactor and the insulation layer, the following condition should be fulfilled:

$$K_{metal}\frac{\partial T_m}{\partial x} = K_{ins}\frac{\partial T_{ins}}{\partial x_{ins}}.$$
(9)

Here, $K_{insulation}$ stands for the conductivity of the insulation material, T_{ins} for the temperature of the insulation and x_{ins} for the thickness of the insulation. To use equation 9 as a boundary condition would imply to develop a fully dynamic model for the insulation on each surface for each section. For simplicity, it is assumed that the insulation reaches a steady state rapidly, which is an assumption for many solid-solid heat transfer systems (Theodore (2011)). With this assumption, the heat loss through the insulation layer is proportional to the temperature difference between the surface metal temperature and the environment temperature multiplied by the conductivity and divided by thickness of the insulation layer, which leads to the following equation:

$$K_{insulation} \frac{\Delta T_{ins}}{\Delta x_{ins}} \approx \frac{T_m - T_{env}}{\Delta x_{ins}} \tag{10}$$

Here, Δx_{ins} stands for the thickness of the layer and T_{env} for the environment temperature.

3. RESULTS

3.1 Optimization of the operating parameters

In order to achieve a high yield with respect to the conversion of SO_2 , an objective function is chosen that minimizes the ratio of the concentration of SO_2 at the outlet of the reactor to the concentration of SO_2 at the inlet of the reactor. The average concentration of SO_2 at the outlet is calculated by equation 12. Due to the nature of the reaction (see section 2.3) it is mandatory to have a high temperature at the beginning of the reactor and a lower temperature at the end of the reactor. To realize this, the inlet temperature of the process fluid and the inlet temperature of the heat transfer fluid are chosen as the degrees of freedom. The inlet temperature of the process fluid $T_{fluid,in}$ is constrained by a minimum temperature of 673 K and a maximum temperature of 820 K. The minimum temperature corresponds to the temperature which is at least required to initiate the reaction (cf. Müller (2000)). The reason for the upper bound of the temperature is the deactivation of the catalyst at higher temperatures. The heat exchange fluid temperature $T_{cool,in}$ is constrained by a minimum temperature of 320 K and a maximum temperature of 820 K. For the minimum temperature, it was assumed that this is the lowest temperature which is required. The maximum temperature was chosen for the same reason as for the process fluid. To avoid that the temperature inside the reactor is greater than 820 K, the temperature of the process fluid along the reactor has to be constrained. To do so, theoretically all discretization points inside the reactor must be constrained. To circumvent this difficulty the point in the reactor is determined where the highest temperature is expected. Here, it is assumed that the highest temperature is at the beginning of the reactor, very close to the inlet. Due to the fact that on the top and on the bottom surfaces of the reactor some heat will be lost, a point in the middle of the reactor was chosen. With these assumptions the optimization problem is given by:

$$\max_{T_{fluid,in}, T_{cool,in}} \left(1 - \frac{\dot{n}_{SO_2,out}}{\dot{n}_{SO_2,in}} \right)$$

Subject to: (11)
$$673K \leq T_{fluid,in} \leq 820K$$

$$300 \text{K} \le T_{cool,in} \le 820 \text{K}$$

 $T_{fluid}|_{x=0.2,y=0.1,z=0+} \le 820 \text{K}.$

The optimization was done in gPROMS by the sequential quadratic programming method.

Table 2 summarizes the settings of the simulation and figure 4 shows the results of the optimization. In **A** the conversion of SO₂ over the width and length of the reactor is shown and in **B** the optimized temperature profile over the width and length of the reactor can be seen. The result of the optimized temperature profile shows very well how the temperature declines along the reactor. The fluid at the entry of the rector has a high temperature ($T_{fluid,in} =$ 795.6 K) which is required to activate the catalyst. The temperature of the heat exchange fluid $T_{cool,in}$ is 593.6 K that leads to an average fluid temperature at the outlet of approx. 600 K. With these setting approx. 98.5 % conversion can be achieved.



Fig. 4. A. Optimized conversion profile. B. Optimized temperature profile.

Table 2. Settings for the simulation

Variable	Numerical value	Unit
x_{O_2}	0.2047	-
x_{SO_2}	0.0293	-
x_{N_2}	0.766	-
\dot{m}_{fluid}	8	kg/h
\dot{m}_{cool}	10	kg/h

3.2 Control structure

The control structure of the reactor is shown in figure 5. At first, the feed is preheated to a constant temperature, as mentioned in section 2.1, before the feed enters the reactor, where the oxidation of SO_2 is calculated. In the mixer behind the reactor the average mole fraction of each component (SO₂, O₂, SO₃, N₂) and the average temperature of the fluid are calculated by,

$$\overline{z}_i = \frac{\sum \sum (z_i w_f)}{\sum \sum (w_f)}.$$
(12)

Here, z_i stands either for the average mole fraction or for the temperature and w_f for the velocity of the fluid. The reason for calculating the average fractions is that the reactor consists of many parallel reaction channels which have different temperature profiles leading to different concentration profiles but the average conversion is mainly of interest. The reactor itself is controlled by a switching controller for the heating cartridges and a PI controller that controls the outlet temperature of the reacting fluid. The temperature at the inlet is not controlled because simulation studies have shown that the best variant is to preheat the process fluid to a constant temperature before the process fluid enters the reactor and to use only the PI controller to influence the conversion. The switching controller for the heating cartridges switches on the cartridges if the temperature at a certain position is less than a threshold and switches off the cartridges if the the temperature at this position is greater than a threshold. This should avoid that the temperature inside the reactor drops below a critical temperature, i.e that the reaction cannot be initiated. The position where the temperature is measured is close to the entry of the reacting fluid into the reactor. The PI controller controls the average process fluid temperature at the outlet calculated by equation 12. This corresponds to the temperature which is measured in the outlet pipe of the reactor. The set-point of the PI controller results from the optimization of the steady state, i.e. the average fluid temperature for the optimal operating conditions at the outlet is taken as the set-point.



Fig. 5. Schematic overview of the control structure of the reactor.



Fig. 6. Step response of the temperature control loop.

Tuning of the PI controller For the tuning of the controller, the transfer function of the path from the cooling temperature to the average reactor outlet temperature system was estimated from simulated step responses in MATALB using the prediction error minimization (PEM) approach to determine the transfer function coefficients. The resulting transfer function is given by:

$$G(s) = \frac{1.182}{(1+716.2449s)(1+121.4272s)}.$$
 (13)

The PI-controller was tuned such that it compensates the slow time constant of the plant transfer function and the gain was tuned in order to obtain a well-damped response, using the frequency response of the open-loop system. To avoid that the temperature of $T_{cool,in}$ is higher than 820 K, the output of the PI-controller is bounded to this upper value. The lower bound of the PI-controller is 400 K which was obtained by simulation studies by testing different lower bounds.

3.3 Start up

The start-up time denotes the time that is required to bring the process from the cold state to the steady state operation. Important aspects here are that the steady state should be reached as quickly as possible. In order to optimize the start-up time the idea is to use the heating cartridges in the initial phase and to drive the reactor to the desired outlet temperature by using the temperature controller. Figure 7 **A** shows the start-up when no heating cartridges and no controller are used. In this case, the start-up time is around 4.5 hours. As a first approach an open-loop control profile was computed by dynamic optimization. The purpose of the optimization is to determine an optimal threshold T_{max} for switching off the cartridges and an optimal cooling temperature profile such that the conversion is maximized for the startup phase. The cooling temperature is parameterized by constant values during N intervals of variable length. The cooling temperature is bounded in each interval by a lower bound of 300 K and an upper bound of 820 K. The reasons for choosing the bounds are the same as explained in section 3.1. T_{max} is bounded by a minimum temperature of 700 K, which is the temperature to initiate the reaction. The optimization problem is stated as:

$$\max_{T_{cool,in}, T_{max}, t_i} \int_{t=t_1}^{t=t_{end}} \left(1 - \frac{\dot{n}_{SO_2, out}}{\dot{n}_{SO_2, in}}\right) dt$$

Subject to:

$$700 \text{K} \leq T_{max} \leq 820 \text{K}$$

$$300 \text{K} \leq T_{cool,in,i} \leq 820 \text{K} \quad \forall i = 1, \dots, N$$

$$1 \text{s} \leq t_i \leq 1000 \text{s} \quad \forall i = 1, \dots, N$$

Here, t_i stands for the length of the control interval and N = 10. Figure 7 **B** shows the result of the optimization. The start-up time can be reduced to around 1.5 hours. The reason for the reduced start-up time is that $T_{cool,in}$ is used in the initial phase as a further heating source until the temperature inside the reactor is nearly 700 K before it drops significantly to obtain the optimal temperature profile. When the number of time intervals is changed, the characteristics of the start-up do not change significantly suggesting that N = 10 is sufficient. As a second step, in order to obtain a feedback control strategies using the PI controller, the start-up strategy is formulated as a dynamic optimization problem as follows. In the first step, the initial cooling temperature $T_{cool,1}$ that is applied from t_0 to t_1 was optimized such that at time t_1 the conversion reaches its maximum value by minimizing the time for heating. Here, $T_{cool,1}$ was bounded between 300 and 800 K. In the second step, $T_{cool}(t)$ and T_{max} were optimized in two constant segments from t_1 to a sufficiently large time t_{end} . The optimization problem can be formulated as:

$$\max_{T_{cool,in}, T_{max}, t_2} \int_{t=t_1}^{t=t_{end}} \left(1 - \frac{\dot{n}_{SO_2, out}}{\dot{n}_{SO_2, in}}\right) dt$$

Subject to:

$$700 \text{K} \le T_{max} \le 820 \text{K}$$

 $300 \text{K} \le T_{cool.in} \le 820 \text{K}$

This optimization was performed for the model of the reactor with the switching controller for the cartridges. The result was an open loop input profile of $T_{cool}(t)$ after t_1 . This profile was then approximated by switching on the PI outlet temperature controller at time t_1 with a suitably chosen lower bound on T_{cool} of 400 K. The resulting start-up sequence is shown in figure 7 C. With this sequence, the start-up can also be reduced to around 1.5 as shown in **B**. This shows that the PI controller is a good choice to control the process, because it is able to react fast enough to drive the system to the steady state. A controller basing on optimizing control inputs, e.g. an NMPC controller, would improve the efficiency only marginally but it would be significantly more complex.



Fig. 7. Simulation results when the reactor is started up **A**. without controller and heating cartridges, **B**. with an optimized open-loop strategy, **C**. using the PI controller.

4. SUMMARY AND OUTLOOK

A three dimensional dynamic distributed parameter model that describes the oxidation of SO_2 to SO_3 in a microstructured reactor was developed. A controller that controls the temperature of the process fluid at the end of the reactor was designed and tested in simulations. A control strategy which reduces the start-up time significantly was developed by dynamic optimization. The next steps of the work are the comparison of the simulation results to experimental data both in steady state and for the startup. The model can also be used for simulations of alternative reactor designs, e.g. different catalysts or modified dimensions in order to improve the efficiency of the reactor and to reduce the cost of manufacturing and of operation.

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