DOI: 10.6060/ivkkt201962fp.5852

УДК: 66.096.5-932.2

ПРОСТАЯ МОДЕЛЬ ДЛЯ ОЦЕНКИ И СРАВНЕНИЯ ЭФФЕСТИВНОСТИ РЕАКТОРА С КИПЯЩИМ СЛОЕМ ПРИ ОТСУТСТВИИ И НАЛИЧИИ ЦИРКУЛЯЦИИ

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Предложена простая модель для оценки и сравнения эффективности реактора с кипящим слоем при отсутствии и наличии циркуляции частиц. Модель базируется на детерминированных дифференциальных уравнениях движения одиночной частицы, где ее масса меняется с течением времени благодаря тому или иному физическому или химическому процессу в реакторе. Изменение массы частицы описано уравнением кинетики реакции первого порядка. Постоянная скорости реакции считается пропорциональной поверхности частицы и скорости ее обтекания газом. Численные эксперименты с моделью выполнены для случая, когда объем частицы остается постоянным, но ее плотность уменьшается. Это соответствует, например, сушке частицы без ее сжатия. Скорость процесса обработки частицы оценена временем, необходимым для уменьшения способной вступить в реакцию массы частицы на 95%. Показано, что скорость преобразования частицы растет с ростом скорости газового потока. Однако, в то же время, время пребывания частицы в реакторе уменьшается, и степень завершения реакции уменьшается для прямоточного реактора. Из этого следует, что в прямоточном реакторе преимущественна обработка частиц в слое, близком к плотному. В реакторе с циркуляционным кипящим слоем частица, достигшая его вершины, направляется вниз реактора и может участвовать в процессе несколько раз до тех пор, пока не будет достигнута заданная степень завершения реакции. В этом случае предпочтительна высокая скорость газа, поскольку при ней интенсифицируется процесс обмена между газом и частицей, и эффективность всего процесса может быть значительно повышена.

Ключевые слова: кипящий слой, циркуляция, частица, переменные во времени свойства, скорость осаждения, скорость газа, степень завершения реакции

A SIMPLE MODEL TO ESTIMATE AND COMPARE EFFICIENCY OF FLUIDIZED BED REACTOR WITHOUT AND WITH CIRCULATION

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A simple model to estimate and compare efficiency of fluidized bed reactor without and with circulation is proposed. The model is based on the deterministic differential equations of a single particle motion in an upstream gas flow where the particle mass varies with time due to this or that physical or chemical process in the reactor. The particle mass variation is described by the equation similar to the equation of the first order chemical reaction kinetics. The constant of the reaction rate is proportional to the particle surface and the local velocity of gas flow around the particle. Numerical experiments with the model were done for the case when the particle volume remains constant but the particle density is decreasing. It corresponds, for example, to the process of particle drying without its shrinking. The rate of the process of particle treatment was estimated as the time, which is necessary to covert 95% of particle mass, which is capable to enter the reaction. It is shown that the rate of particle conversion grows with the gas flow velocity. However, at the same time, the particle residence time is getting smaller, and the degree of the reaction completion becomes less for direct-current reactor. It follows from that that in a direct-current reactor the bed close to the dense one is preferable. In a circulating fluidized bed reactor, the particle after reaching the top of it is directed to the bottom of the reactor and can participate in the process several times until it reaches the required degree of reaction completion. In this case much higher gas flow velocity is preferable because it intensifies gas-particle exchange, and the efficiency of the process can be considerably improved.

Key words: fluidized bed, circulation, particle, time-varying properties, particle settling velocity, gas flow velocity, degree of reaction completion

Для цитирования:
Мизонов В.Е., Митрофанов А.В., Tannous K., Camelo A. Простая модель для оценки и сравнения эффективности реактора с кипящим слоем при отсутствии и наличии циркуляции. *Изв. вузов. Химия и хим. технология*. 2019. Т. 62. Вып. 3. С. 83–88

For citation:
Mizonov V.E., Mitrofanov A.V., Tannous K., Camelo A. A simple model to estimate and compare efficiency of fluidized bed reactor without and with circulation. *Izv. Vyssh. Uchebn. Zaved. Khim. Khim. Tekhnol*. 2019. V. 62. N 3. P. 83–88

INTRODUCTION

Fluidized bed reactors are widely used in chemical and other industries. Development of computer technologies gave an impulse to modeling the process for engineering practice. However, the complexity of the process brought about a broad variety of approaches to modeling it. The comparative analysis of different approaches is presented in [1]. This paper gives preference to the approach based on the theory of Markov chains. The matrix of transition probabilities, which is the main operator of this theory, can be compared to a mathematical image of a real process. This approach was successfully used by Dehling et al. [2, 3] and by some other authors to describe the process of fluidization. The broadest range of its application in particle technology is presented in [4]. However, the cited works only dealt with the single-stage reactors while the circulating fluidized bed reactors are also of high industrial importance. A circulating fluidized bed reactor consists of the following main parts: the riser (the fluidized bed reactor itself), the separator which separates the particulate phase from a fluidizing gas and routes the phase to the downer, the downer, in which particles move down, and the valve that controls the particle flow moving from the downer to the bottom of the riser. Thus, the particles can circulate in the apparatus and stay inside of it as long as necessary no matter how high the gas velocity in the riser is (assuming the separator is a perfect one).

A relatively novel hybrid Euler-Lagrange approach to modeling the dense gas-solid flow combined with a combustion process in a large-scale industrial circulating fluidized bed boiler is presented in [5, 6]. Despite the fact that calculated results are in good correlation with the industrial experimental data, the circulation itself and its influence on the hydrodynamic state of the apparatus were not among the objectives of the study in this work.

The study by Qi et al. [7] focused on developing a new comprehensive correlation for better prediction of the solids concentration distribution in the fully developed region of co-current upward gas-solid flow occurring in circulating fluidized bed risers. It is shown that the correlation works well for a wide range of operating conditions, particle properties and riser diameters. However, the interconnection between the riser and downer, in which the particulate phase circulates, was not examined.

The experimental study of drying solid materials in the riser of the circulating fluidized bed covering a wide range of operating parameters is described in [8]. The influence of initial moisture content, temperature, and flow rate of the heating medium, and the solid circulation rate on the rate of drying was critically examined. Although the process of circulation was indeed taken into account in this work, it was a purely experimental study that did not imply any attempts to implement any theories. An interesting experimental work is described in Van den Moortel et al. [9] where a phase Doppler particle analyzer was used to measure the particle concentration distribution in the riser of a circulating fluidized bed. Its results help to understand better the local structure of fluidization but, again, only the riser was investigated.

The analysis of the above-mentioned and many other publications allows us to make a conclusion that the circulating fluidized bed is the objective of many research works. However, these works are mostly devoted to experimental (more seldom theoretical) investigation of the process in separate parts of a circulating fluidized bed reactor, and practically never to the entire circulating loop. Thus, the interference between the processes in the riser, separator and downer remains unexamined. On the other hand, it is known that in a closed milling circuit this interference can lead to circulation stability loss and subsequent blockage of the mill [10]. In particular, the paper [11] shows that, other conditions being equal, optimizing the position of the circulation flow input over the length of a tube mill can increase the fineness of the end product.

Therefore, the study of the circulating fluidized bed in terms of particulate flows formation is of actual scientific and technological value. Some approaches based on the theory of Markov chains to build such models are described in [12, 13] where some interesting and unexpected result were found. Nevertheless, all these models remain too complex and computational time consuming to use them for preliminary estimation of the role of particles with variable properties circulation in a circulating fluidized bed. It can be done on the basis of a model of a single particle motion in an upstream gas flow. Being rather approximate such model allows taking into account the effect of the first order importance and separating a range of the process parameters for more detailed modeling. The single particle models are successfully used in different fields of research [14-17]. In particular, it was used to model of single plastic particle conversion in suspension [18], for numerical simulation of high-temperature fusion combustion characteristics for a single biomass particle [19], for investigation of effects of size and loading rate on the mechanical properties of single coral particles [20]. Namely this approach inspired developing the model presented below.

THEORY

Fig. 1 shows schematically a simplified presentation of the process. It is one-dimensional motion of a particle with time-varying properties in an upstream gas flow. At the initial moment of time, the particle is placed on a gas distributor. It enters into a physical or chemical interaction with the gas flow that leads to its properties variation. While the particle settling velocity is less than the gas flow velocity, the particle remains of the gas distributor surface. Then it begins to move up with the gas flow. After reaching the top of a fluidized bed reactor it can leave the process (direct current fluidized bed reactor), or can be directed in the circulation loop to appear the bottom of the reactor and enter to the interaction with the gas flow again (circulating fluidized bed reactor).

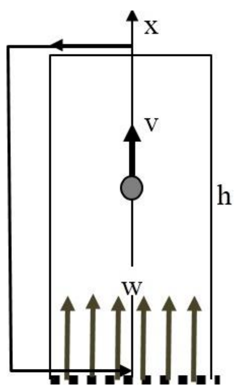

Fig. 1. The computational scheme of the model
Рис. 1. Расчетная схема модели

The governing equations of the model and the initial conditions are listed below:

$$\frac{dv}{dt} = -g + \frac{1}{m} c_f \frac{\pi d^2}{4} \frac{\rho_g (w-v)^2}{2} \quad (1)$$

$$\frac{dx}{dt} = v \quad (2)$$

$$\frac{dm}{dt} = -\beta(m - m_2) \quad (3)$$

$$v(0)=0,\ x(0)=0,\ m(0)=m1 \quad (4)$$

where v is the particle velocity, x is its co-ordinate, w is the gas flow velocity, m is the particle mass, d is the particle diameter, g is the gravity acceleration, ρ_g is the gas density, c_f is the drag force coefficient.

The law of the particle mass variation presented by Eq. (3) is take as the analogy to the first order chemical reaction where m_2 is the residual mass that do not enter to the reaction. The reaction rate is supposed proportional to the particle surface S and to the velocity of the particle flow around (w-v):

$$\beta = \alpha S(w - v)^z \qquad (5)$$

where α and z are constant parameters.

Eq. (3) can be rewritten as follows:

$$\frac{d(\rho_p V_p)}{dt} = -\beta(\rho_p V_p - \rho_{p_2} V_{p_2}) \qquad (6)$$

where ρ_p and V_p are the particle density and its volume.

Two end cases can be examined here. The first one is related to the case V_p = const. For instance, it concerns drying a particle with the rigid skeleton, i.e., without particle shrinking during the process. In this case

$$\frac{d(\rho_p)}{dt} = -\beta(\rho_p - \rho_{p_2}) \qquad (7)$$

The other case corresponds to a reaction, in which the solid product if the reaction leaves the particle surface with the gas, and the nuclear of the particle remains unchangeable. In this case

$$\frac{d(V_p)}{dt} = -\beta(V_p - V_{p_2}) \qquad (8)$$

The drag force coefficient can be described by the one term equality

$$c_f = \frac{a}{Re^n} \qquad (9)$$

where a = 24 and n = 1 for the Stokes law (small Reynolds number), a = 13 and n = 0.5 for the Allen's law (average Reynolds number), and a = 0.48 and n = 0 for the Newton's law (high Reynolds number). Substitution of Eq. (9) in Eq. (1) gives the following equation:

$$\frac{dv}{dt} = -g + \frac{3}{4} a \frac{\rho_g}{\rho_p} \frac{v^n}{d^{1+n}} (w - v)^{2-n} \qquad (10)$$

Under these assumptions, the particle settling velocity can be calculated as follows:

$$v_s = \left(\frac{4}{3}\frac{g}{a}\frac{\rho_p}{\rho_g}\frac{d^{1+n}}{v^n}\right)^{\frac{1}{2-n}} \qquad (11)$$

The necessary duration of the process can be estimated by the time t_{95} when 95% of the reactive mass completed the reaction:

$$\frac{m(t_{95}) - m_2}{m_1 - m_2} = 1 - 0.95 \qquad (12)$$

The model allows describing the particle motion and conversion when it moves in the upstream gas flow belonging to a fluidized bed reactor.

RESULTS AND DISCUSSION

The numerical analysis of the described above set of equations for the case V_p = const is presented below. The calculations were done with d_p = 2mm, ρ_{p1} = 1000kg/m^3, ρ_{p2} = 500kg/m^3, a = 13, n = 0.5, z = 0.8, h = 5m.

Fig. 1 shows the graphs of the particle density variation with time for different gas flow velocity. The process goes like follows. If the initial particle settling velocity v_{s1} is less than the gas flow velocity, the particle remains on the distributor surface (x = 0). The value of v_s is decreasing with time due to the particle density decrease according to Eq. (11). When it becomes equal to the gas flow velocity the particle begins to move up with increasing velocity but with decreasing the flow around velocity (w-v) that leads to the decrease of the exchange reaction rate. If $v_s(t_{95})$ is still less than w, the particle remains on the distributor surface during the entire process.

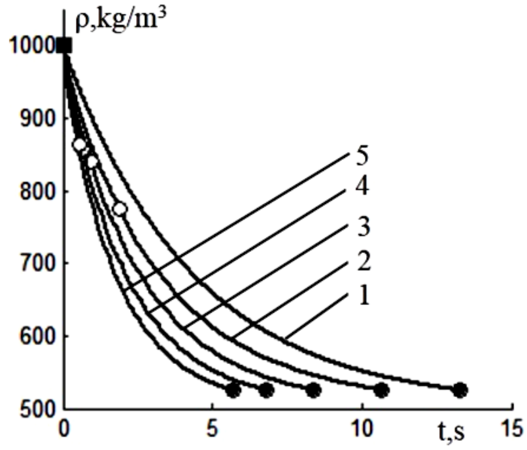

Fig. 2. Particle density versus time for different gas flow velocity w, m/s: 1 – 5; 2 – 10; 3 – 15; 4 – 20; 5 – 25

Рис. 2. Зависимость плотности частицы от времени при различных скоростях газа w, м/с: 1 – 5; 2 – 10; 3 – 15; 4 – 20; 5 – 25

The black circles are related to the 95% reaction completion. It can be seen that the rate of density variation grows considerably with the increase of w. However, these circles correspond to unlimited height of the reactor. The white circles correspond to the moments of time when the particle reaches the height x = h = 5m (the particle residence time). If we deal with a direct-current reactor, the particle leaves it with the degree of reaction completion, which becomes less with the gas velocity growth. There is no white circle on the curve 1. It means that the particle remains at the position x = 0 all time long if w = 5 m/s. It can be concluded from the graphs that in a direct-current reactor treatment of particles is more effective if they are treated in the state, which is close to the dense bed.

Now suppose that the particle moves in a circulating fluidized bed reactor. It means that after reaching the point x = h it appears in a circulating loop, which turns it back to the state x = 0, v = 0. It is supposed that no transformation occurs with the particle in the loop. It is obvious that the particle already has the settling velocity that is less than gas flow velocity, and it immediately begins its upward motion. The next example was calculated for w = 7m/s. The graphs of v(t) and x(t) are shown in Fig. 3.

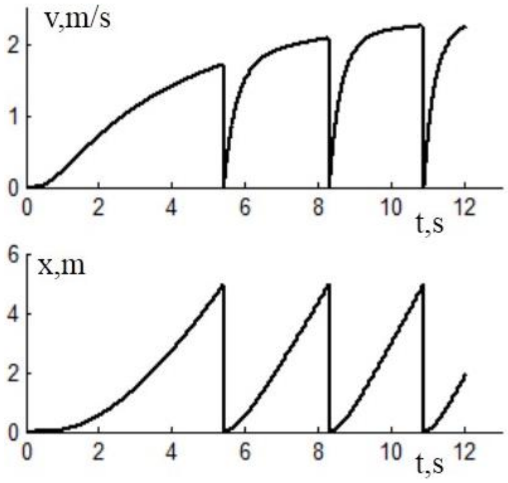

Fig. 3. Graphs of the particle circulating motion
Рис. 3. Графики циркуляционного движения частицы

It can be seen that during the certain time interval (about 0.3s) the particle has to wait on the gas distributor surface until its settling velocity becomes less than the gas flow velocity. The necessary treatment time t_{95} = 12s is less than one for the unlimited height reactor, which is 13.3 s. It can be explained by the fact that the particle periodically appears in the state x = 0, v = 0 when the velocity of flow around has maximum.

Fig. 4 shows how the particle density varies with its position in the circulating reactor. Between the black square and triangle the particle lays on the distributor because its v_s is still higher than w.

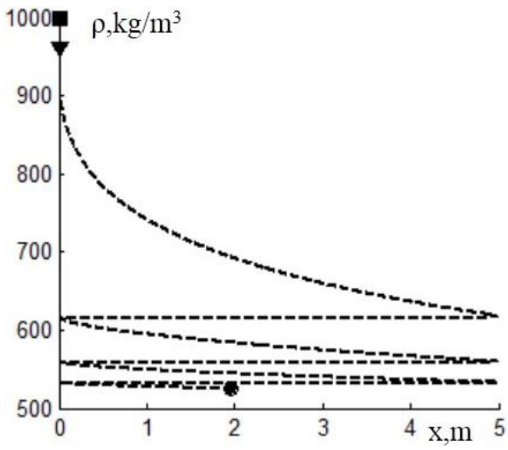

Fig. 4. Particle density versus its position in the circulating reactor
Рис. 4. Зависимость плотности частицы от ее положения в циркуляционном реакторе

It can be seen that in order to reach 95% completion of the reactions (black circle) the particle has to circulate 3 times.

CONCLUSIONS

A simple model to estimate and compare efficiency of fluidized bed reactor without and with circulation is proposed. The model is based on the differential equations of a particle motion in an upstream gas flow where the particle properties vary with time due to this or that physical or chemical process in the reactor. In particular, the particle settling velocity also varies with time. If it is higher than the gas flow velocity the particle is located on the gas distributor until the settling velocity becomes smaller than the gas flow velocity. It is shown that the rate of the particle conversion grows with the gas flow velocity. However, at the same time, the particle residence time is getting smaller, and the degree of the reaction completion becomes less for direct-current reactor. It follows from that that in a direct-current reactor the bed close to the dense one is preferable. On the other hand, if the reactor has a circulation loop, it is preferable to have much higher gas velocity (developed fluidized bed) because the particle can participate in the process several times, and the efficiency of the process can be considerably improved.

ACKNOWLEDGEMENT

This work is supported by the Russian Foundation for Basic Research (project 18-08-00028).

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Поступила в редакцию 05.06.2018
Принята к опубликованию 29.08.2018

Received 05.06.2018
Accepted 29.08.2018