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ТЕОРЕТИЧЕСКОЕ ИССЛЕДОВАНИЕ ПЕРИОДИЧЕСКОГО ЦИРКУЛЯЦИОННОГО ПСЕВДООЖИЖЕНИЯ

Целью исследования была разработка простой, но информативной модели, позволяющей оценить распределение концентрации частиц и их потоки в периодическом циркуляционном псевдоожиженном слое. Для этой цели использовалась ячеечная модель, базирующаяся на теории цепей Маркова. Подъемная и опускная части аппарата с циркуляционным псевдоожиженным слоем были представлены одномерными цепями ячеек идеального перемешивания. Эволюция содержания частиц в ячейках описана матрицами переходных вероятностей, разных для подъемной и опускной зон. Считалось, что эти переходные вероятности зависят от текущего содержания частиц в ячейках. Верхние ячейки цепей связаны через сепаратор, отделяющий частицы от газа, который мог быть идеальным или неидеальным, а нижние ячейки – через клапан, контролирующий поток частиц из опускной части в подъемную. Основной целью численных экспериментов с моделью было оценить переходные и установившиеся процессы с учетом взаимного влияния псевдоожижения частиц в подъемной части и их движения вниз в опускной части. Показано, что в некоторых интервалах изменения параметров процесса циркуляционный контур очень чувствителен к ним, а в некоторых интервалах – нет. Также исследован случай изменения скорости седиментации частиц с течением времени, и показано, что это изменение оказывает сильное влияние на эволюцию параметров процесса.

Ключевые слова: псевдоожижение, циркуляционный псевдоожиженный слой, цепь Маркова, скорость газа в свободном объеме слоя, скорость седиментации частиц, вектор состояния, матрица переходных вероятностей, кратность циркуляции

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THEORETICAL STUDY OF THE BATCH CIRCULATING FLUIDIZATION

The objective of the study is to develop a simple but informative model to estimate the particle concentration distribution and their flows in a batch circulating fluidized bed. A cell model based on the theory of Markov chains is used for that. The riser and downer of the circulating fluidized bed apparatus are presented as 1D arrays of perfectly mixed cells. The evolution of particle content in the cells is controlled by the matrixes of transition probabilities, which are different for the riser and downer. It is supposed that these transition probabilities depend on the particle content in the cells. The upper cells of the chains are connected through a gas-solid separator, which can be perfect or not, and the bottom cells are connected through a valve that controls the particle flow from the downer to the riser. The key objective of the numerical experiments with the model was to estimate the transient and steady-state processes in such apparatus taking into account the interference of the processes of particle fluidization in the riser and their downward motion in the downer. In some intervals of the process parameters variation the circulation loop was shown to be very sensitive to them, and in some intervals it is not. The case of time-varying settling velocity of particles was examined too, and it was shown that this variation had the large influence on the process parameters evolution.

Key words: fluidization, circulating fluidized bed, Markov's chain, superficial velocity, particle settling velocity, state vector, transition probabilities matrix, circulation ratio

INTRODUCTION

Fluidized bed reactors are widely used in the 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 our paper [1]. This paper gives preference to the approach based on the theory of Markov chains. It particularly emphasizes the fact that the theory of Markov chains is native to many processes in particle technology. The theory of Markov chains investigates the evolution of probabilities distribution in a sample space, and a real process with particulate solids investigates the evolution of particles distribution over operating volume, or over any other property of the particles. 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 in [2,3] and by some other authors to describe the process of fluidization but the broadest range of its application in particle technology is presented in [4,5]. However, the cited works dealt with the single stage reactors while the circulating fluidized bed reactors are of high industrial importance too. Such reactors allow intensifying the interphase exchange by means of higher gas velocity that is impossible for usual fluidized bed reactors because all particles can be blown out from the bed. A circulating fluidized bed reactor consists of the following main parts: the raiser (the fluidized bed reactor itself), the separator that separates the particulate phase from a fluidizing gas and directs the phase to the downer, the downer, in which particles move down, and the valve that controls the particle flow from the downer to the bottom of the riser. Thus, the particles can circulate in the apparatus and stay inside it as long as necessary no matter how high the gas velocity in the riser is.

The constantly developing fluidized combustion technology has become competitive with the conventional pulverized coal combustion. Circulating fluidized bed boilers can be a good alternative to pulverized coal boilers due to their robustness and lower sensitivity to the fuel quality. However, appropriate engineering tools that can be used to model and optimize the construction and operating parameters of a circulating fluidized bed reactor still require development [6].

The paper [7] focuses on developing a new comprehensive correlation for better prediction of the solids concentration in the fully developed region of co-current upward gas–solid flow 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, the particulate phase circulates in which, and the efficiency of a gas-solid separator were not examined.

A relatively novel hybrid Euler–Lagrange approach to model the dense gas–solid flow combined with a combustion process in a large-scale industrial circulating fluidized bed boiler is presented in [6]. Despite of 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 objectives of the study in this work.

The experimental study of drying solid materials in the riser of circulating fluidized bed covering a wide range of operating parameters is described in [8]. The effects of initial moisture content, temperature and flow rate of the heating medium, and solid circulation rate on rate of drying was critically examined. It has been observed that the solid materials used in this studt exhibits a falling rate period of drying, and the rate of drying is influenced by the temperature and flow rate of the heating medium, solids circulation rate and initial moisture content. The process of circulation was really taken into account in this work but it was a pure experimental study without an attempt to involve any theory.

The interesting experimental work is described in [9] where a phase Doppler particle analyzer was used to measure particle concentration distribution in the riser of circulating fluidized bed. The results help to understand better the local structure of fluidization but, again, only the riser was investigated.

Analysis of the mentioned above and many other publications allow making the 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 fluidized bed reactor, and practically never to investigation the circulating loop itself (in particular, to the degree of circulation). Thus, the interference of the process in the riser and downer remains not examined. On the other hand, it is known that in a closed milling circuit this interference can lead to principal changed in the circuit functioning [10]. It particular, the stability of circulation can be lost, and blockage of the mill can happen. Therefore the study of circulating fluidized bed from this viewpoint is an actual scientific and technical problem.

THEORY

The circulating fluidized bed reactor is presented schematically in Fig.1, left. It consists of the fluidized bed reactor proper 1 with a gas distributor at the bottom (the riser), the gas-solid separator 2, the tube, in which particles move downward (the downer) 3, and the valve 4 that controls the particulate flow to the riser bottom. The gas flow fluidizes the particulate solids, and if its velocity is high enough, particles reach the top of the riser. The separator separates them from the gas and directs them to the top of the downer where they, after falling down, move through the valve to the riser bottom forming the circulation loop.

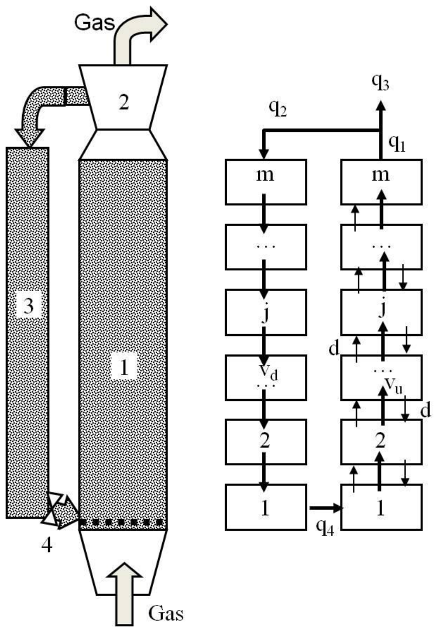

Fig. 1. Schematic presentation of a circulating fluidized bed (left) and its cell model (right)

Рис. 1. Схема циркулирующего псевдоожиженного слоя (слева) и его ячеечная модель (справа)

The structure of the cell model of the process is shown in Fig. 1, right. The riser and downer are presented as two one-dimensional arrays of m perfectly mixed cells. The particle content distribution over a chain can be described by the column state vector **S** = {S_j} of the size m×1. The state of the process is observed at discrete moments of time $t_k = (k - 1)\Delta t$ where Δt is the transition duration, and k is the transition number that can be interpreted as the discrete analogue of time. In this case, the evolution of the particle content distribution can be described by the recurrent matrix equation:

$$\mathbf{S}^{k+1} = \mathbf{P}\mathbf{S}^{k}, \tag{1}$$

where **P** is the matrix of transition probabilities that distributes S_j over the cells at each time transition. The j-th column of it belongs to the j-th cell and contains the transition probabilities to go to the neighboring cells and to stay within this cell. These probabilities have the symmetrical part d that is related to the pure quasi-diffusion and the non-symmetrical part v caused by particle transportation by the gas flow. The values of d and v can be calculated as: $d = D\Delta t/\Delta x^2$, $v = V\Delta t/\Delta x$ where D is the dispersion coefficient, V is the velocity of particle motion, Δx is the cell height. Thus, the matrix **P** is a tri-diagonal matrix. The matrix for the riser differs from the one for the downer, and the processes are to be examined separately.

The process in the riser

It transforms the state vector $\mathbf{S_r}$ of the riser. The basic form of the matrix for the riser $\mathbf{P_r}$ belongs to the closed ergodic chain. Its elements can be calculated as follows:

$$P^k_{r,j,j+1} = v^k_j\left(1 - \frac{S^k_{r,j+1}}{S_{max}}\right)e + d\,,\ j=1,\ldots,m-1 \tag{2}$$

$$P^k_{r,j+1,j} = \left|v^k_j\right|\left(1 - \frac{S^k_{rj}}{S_{max}}\right)(1-e) + d\,,\ j=1,\ldots,m-1 \tag{3}$$

$$P^k_{r,j,j} = 1 - \sum_{i=1,i\neq j}^{m} P^k_{r,i,j}\,,\ i=1,\ldots,m, \tag{4}$$

where e = 1 if $v_j^k>0$ and e = 0 if $v_j^k<0$.

Here, v_j^k is the rate of particles motion in the j-th cell. This is the difference between the upward gas velocity and particle settling velocity v_s. However, if the gas superficial velocity is equal to w_s (the gas velocity in the empty riser), the velocity of flow around particles in higher than w_0 due to smaller hydraulic section because of the presence of particles in the cell.

Suppose that each cell can contain the maximum volume S_{max} of dense packed particles and their porosity in this state is equal to ε. In this case the value of v_j^k can be calculates as:

$$v^k_j = w_s\left(\frac{1}{1 - \varepsilon\dfrac{S^k_j}{S_{max}}} - \frac{v^k_s}{w_0}\right) \tag{5}$$

However, this velocity coincides with the corresponding transition probability only if the particulate outflow from the cell j occurs into an empty neighboring cell. If the neighboring cell is filled with particles the transition occurs only into the void volume in the cell. This factor is taken into account by the multiplier in the parenthesis in Eqs. (2, 3). If the neighboring cell is completely filled with particles, the transition into it becomes forbidden. Thus, the evolution of the state vector in the closed chain for the riser after the k-th time transition is described by the recurrent matrix equation

$$\mathbf{S}^{k+1}_{\mathbf{r}} = \mathbf{P}(\mathbf{S}^{k}_{\mathbf{r}})\mathbf{S}^{k}_{\mathbf{r}}, \tag{6}$$

where the transition matrix is state dependent, i.e., the model is non-linear.

The outflow from the upper cell m of the riser (i.e., the part of particles that leaves it during one time transition) is to be calculated separately:

$$q^{k+1}_1 = S^{k+1}_{r,m} v^k_m$$

After that, the state $S_{r,m}^{k+1}$ becomes

$$S^{k+1}_{r,m} := S^{k+1}_{r,m} - q^{k+1}_1, \tag{7}$$

where := is the assignment operator.

This flow q_1^{k+1} goes to the separator, in which its part φ leaves the apparatus, and the part (1-φ) goes to the upper cell of the downer (φ = 0 for the perfect separator).

The process in the downer

It transforms the state vector $\mathbf{S_d}$ of the downer. The basic form of the matrix for the riser $\mathbf{P_d}$ belongs to the closed ergodic chain. The only difference between the processes is that there is no upstream gas flow in the downer, and particles move down with their settling velocities. The elements of the matrix $\mathbf{P_d}$ can be calculated as follows:

$$P^k_{d,j,j+1} = d\,,\ j=1,\ldots,m-1 \tag{8}$$

$$P^k_{d,j+1,j} = v^k_s\left(1 - \frac{S^k_{dj}}{S_{max}}\right) + d\,,\ j=1,\ldots,m-1 \tag{9}$$

$$P^k_{d,j,j} = 1 - \sum_{i=1,i\neq j}^{m} P^k_{d,i,j}\,,\ i=1,\ldots,m \tag{10}$$

The bottom cell of the downer is connected with the bottom cell of the riser through the regulated valve. This valve takes the z-th part of particles in the cell 1 of the downer and directs it into the cell 1 of the riser. Thus, we get:

$$q^{k+1}_4 = S^{k+1}_{d,1} z \tag{11}$$

$$S_{d,1}^{k+1} := S_{d,1}^{k+1} - q_4^{k+1} \quad (12)$$

$$S_{u,1}^{k+1} := S_{u,1}^{k+1} + q_4^{k+1}, \quad (13)$$

where := is the assignment operator.

Eqs. (11)-(13) close the circulation loop and allow calculating the particle content distribution in the entire system.

RESULTS AND DISCUSSION

The objective of the section is to show in numerical experiments how the model works.

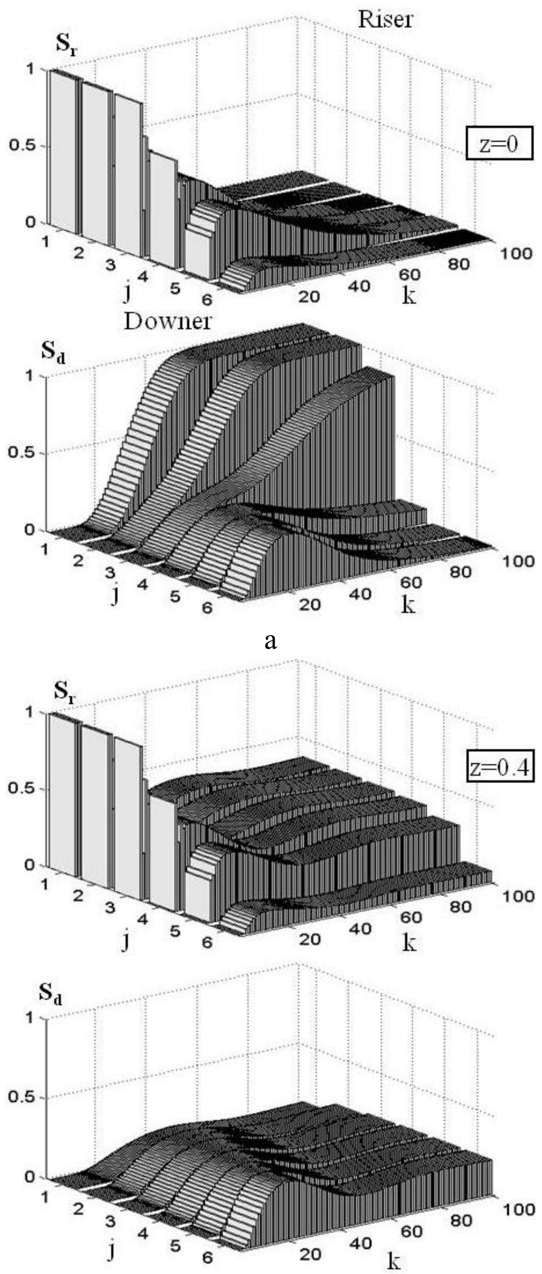

Fig. 2. Particle content distribution at closed (z=0) and partly open (z=0.4) valve: w_s=0.4; v_s=0.3; d=0.1

Рис. 2. Распределение содержания частиц при закрытом (z=0) и частично открытом кране (z=0,4): w_s=0,4; v_s=0,3; d=0,1

Fig. 2 illustrates the evolution of particle content distribution in the riser and downer at closed and partly open valve.

The calculations were done for m=6 cells. Initially, the particles occupy 3 cells at the bottom of the riser. After beginning of gas supply the bed in the riser expands, reaches the top cell 6, and begins to enter to the top of the downer, in which the particles move down to its bottom. If the valve is closed (the upper graphs) the bed should transit to the bottom of the riser little by little. It can be seen at the graphs: the riser is actually empty but the bottom of the downer is almost fully filled with particles; only small part of them had no enough time to settle.

If the valve is partly or fully open (the lower graphs), the circulation begins, and the steady-state distribution comes asymptotically. The kinetics of the process can be seen at the graph. The steady-state distribution is being reached when the flows q_2 and q_4 become practically equal. The graph of these flows variation with time is shown in Fig. 3.

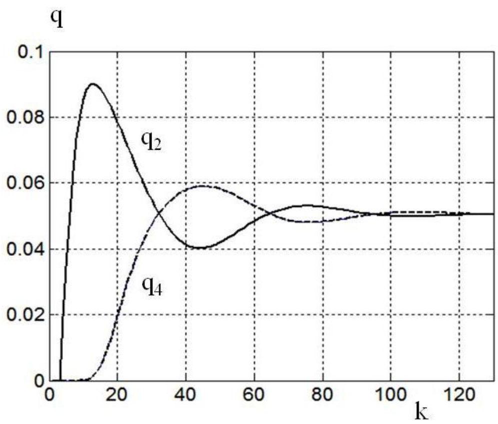

Fig. 3. Stabilization of outflows from the riser and downer: w_s = 0,4; v_s = 0,3; d = 0,1; z = 0,4

Рис. 3. Стабилизация выходящих потоков в кипящем слое и возврате: w_s = 0,4; v_s = 0,3; d = 0,1; z = 0,4

It can be seen that stabilization of outflows from the riser and downer goes like damped oscillations with time delay of one graph with respect to another. The process is practically stabilized after 130 time transitions.

The steady-state hold-up in the riser Q_r and downer Q_d can be calculated as follows:

$$Q_r = \sum_{j=1}^{m} S_{r,j}^{\infty} \quad \text{and} \quad Q_d = \sum_{j=1}^{m} S_{d,j}^{\infty} \quad (14)$$

and the ratio

$$K_c = \frac{Q_d}{Q_r} \quad (15)$$

can be called as the degree of circulation.

Fig. 4 shows how the degree varies with the gas flow velocity at various position of the valve. The fluidization begins at w_s = 0.122; at lower velocity the particles remain in the dense packed state. Until w_s = 0.212 the bed expands but still remains within the riser. At higher velocity the circulation begins, and at partly or fully open valve the system reaches the steady state. It is clear physically that the more valve is closed the higher degree of circulation is. For instance, at z = 0.1 and w_s = 0.45 the degree of circulation K_c = 1, i.e., the riser and downer contain equal amount of particles. On the other hand, the degree of circulation grows with the growing gas flow velocity. It is also clear because at high gas velocity the particles pass the riser quickly but they are settling in the downer with the same settling velocity, i.e., the rate of their settling does not grow with the growth of w_s.

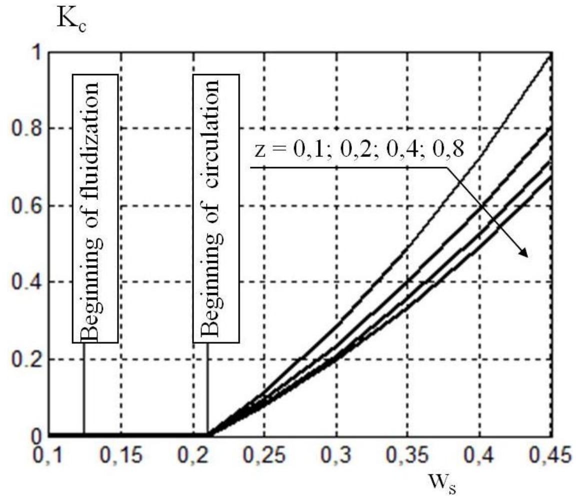

Fig. 4. Influence of the gas flow velocity on the degree of circulation at various z: v_s=0.3; d=0

Рис. 4. Влияние скорости газового потока на степень циркуляции при различных z: v_s=0,3; d=0

If the separator is imperfect, the bed looses particles. Fig. 5a-c illustrates how the flows q_2 and q_4 vary with time at different value of φ. Fig. 5a repeats the graph in Fig. 3 because φ = 0 in this case, and the system reaches the steady-state circulation. However, if φ ≠ 0, the both flows gradually decrease with time tending to zero asymptotically (Fig. 5b,c). It means that asymptotically the bed looses all particles that were loaded into it, the higher φ, the faster at that. At last, Fig. 5d how the hold-up in the riser decreases with time at different value of φ.

The particle settling velocity can vary with time due to thermal or chemical treatment of particles. For example, wet mineral porous particles can change their mass during drying practically without changing their diameter. It means that their effective density

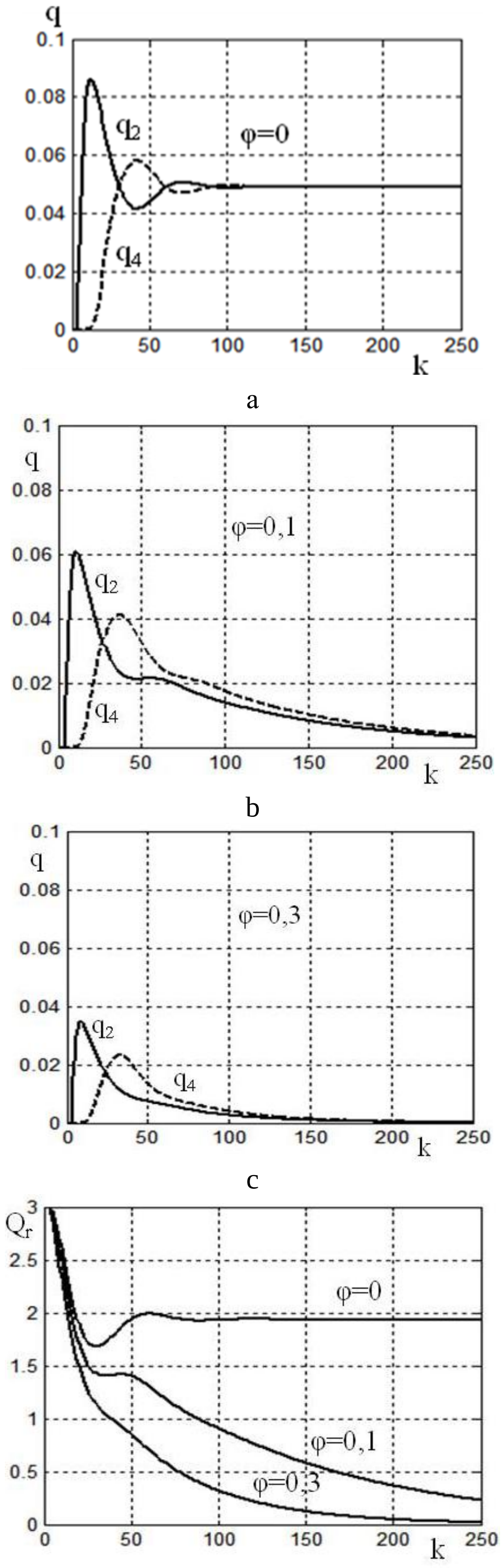

Fig. 5. Variation of outflows from the riser and downer at different values of φ (a-c) and variation of the hold-up in the riser (d): w_s=0.4; v_s=0.3; d=0.1; z=0.4

Рис. 5. Изменение выходных потоков в кипящем слое и возврате при различных значениях φ (a-c) и изменении задержки в вертикальном трубопроводе (d): w_s=0,4; v_s=0,3; d=0,1; z=0,4

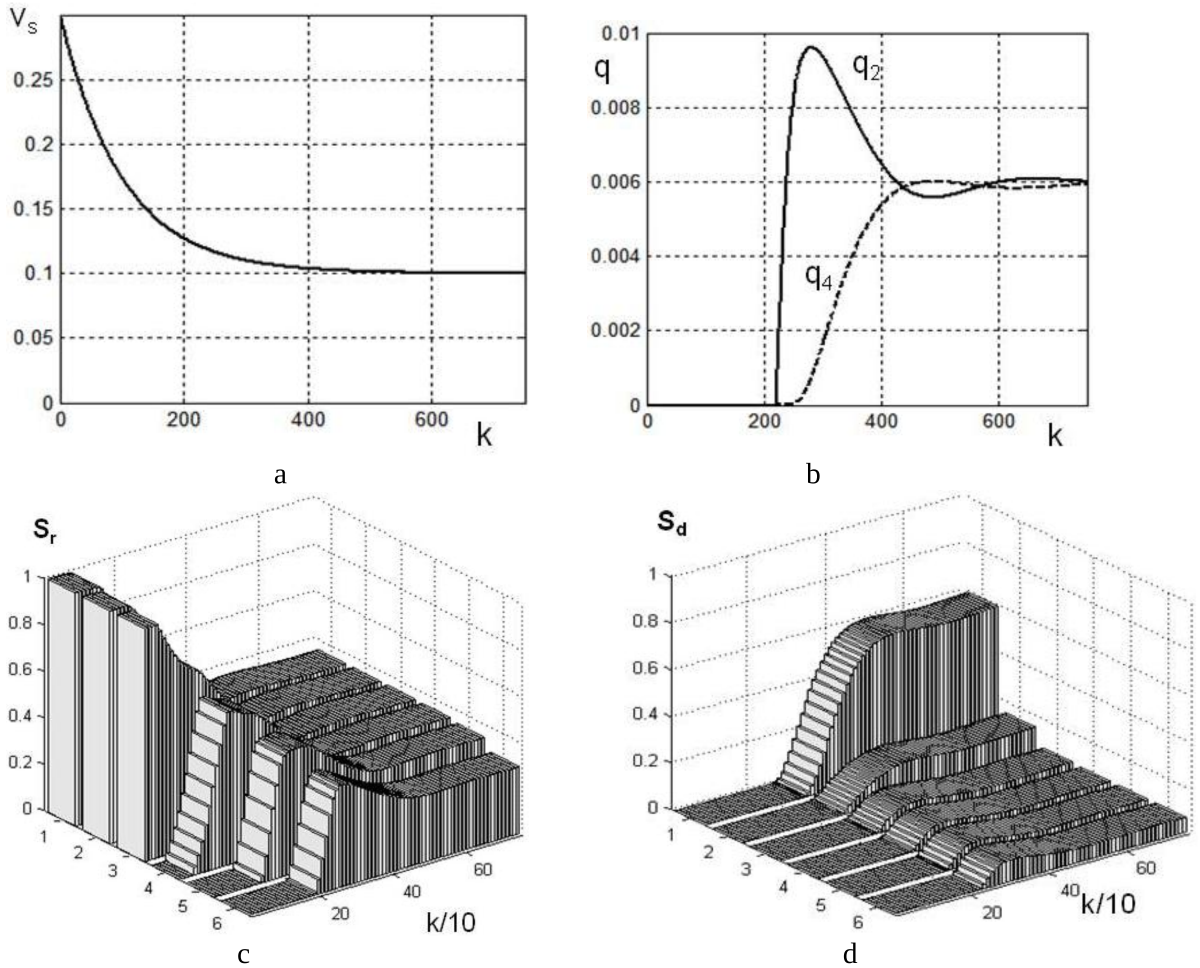

Fig. 6. Process with variable settling velocity: a) settling velocity versus time; b) outflows from the riser and downer; c) evolution of particle content distribution in the riser; d) evolution of particle content distribution in the downer: w_s=0.1; d=0.1; z=0.01

Рис. 6. Процесс с переменной скоростью осаждения: a) скорость осаждения как функция времени; b) выходящие потоки из кипящего слоя и возврата с) эволюция распределения содержания частиц в кипящем слое; d) эволюция распределения содержания частиц в возврате: w_s=0,1; d=0,1; z=0,01

changes during drying that leads to decreasing of their settling velocity. Suppose that the settling velocity varies with time according to the following formula

$$v_s^k = v_{s2} + (v_{s1} - v_{s2})e^{-b(k-1)}, \qquad (16)$$

where v_{s1} and v_{s2} is the initial and asymptotical settling velocity, b is the proportional coefficient.

An example of modeling of such process kinetics is shown in Fig. 6 where the following parameters of the process were used: v_{s1} = 0.3, v_{s2} = 0.1, b = 0.01, d = 0.1, z = 0.01, φ = 0 and w_s = 0.1 that was taken small enough to demonstrate all specific feature of the process. Fig.6a simply illustrates how the settling velocity varies with time. It can be seen from Fig.6c that the bed remains in the dense packed state during 40 time transitions because the settling velocity is much higher than the local gas flow velocity. After 40 time transitions fluidization begins, and the bed expands gradually without circulation until k = 215. At this moment of time the bed reaches the upper cell of the riser and the circulation begins. After that the system gradually transits to the steady-state that corresponds to the asymptotical settling velocity equal to 0.1. Stabilization of outflows from the riser and downer is shown in Fig. 6b.

Thus, the proposed Markov chain model of circulating fluidized bed allows describing practically all characteristics of the process that are needed for industrial application both for the transient and steady-state conditions.

CONCLUSIONS AND PERSPECTIVES

The Markov chain model proposed earlier for a direct flow fluidized bed reactor [1] is generalized to the case of circulating fluidized bed reactor, which can be found in many industries. This model deals with the reactor as a system with distributed parameters and allows describing the transient and steady-state processes in the bed for particulate solids of constant and variable properties. The apparent advantage of such cell models is low computational time. It takes several minutes to calculate a transient process until it reaches a steady-state (if exists). It opens the way of fast comparison of various variants of the process at different regime and design parameters to find the most appropriate ones for industrial design. The prospective of the model development is its generalization to the case of polydispersed particulate solids and different conditions of particle settling velocity variation in the riser and downer.

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