

THE DEVELOPMENT OF CARBON MONOXIDE OXIDATION REACTOR FOR MULTI-CHAMBER FURNACES FOR BAKING ELECTRODE BLANKS

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Abstract

The article highlights the development of cheap affordable highly efficient catalytic oxidation system of harmful components of industrial flue gases of carbon graphite enterprises, its design features according to the parameters of furnace equipment, which will significantly improve the environmental safety of metallurgical and machine-building enterprises. The presented calculation and design solutions, in addition to carbon production, can be used in environmental protection technologies at other environmentally hazardous facilities to neutralize toxic emissions.

The paper presents the calculation of the catalytic CO oxidation reactor made for the real flow rate of flue gases with temperature range 270–390 °C, which come out of the combustion chamber of the Riedhammer «first fire» kiln, heated by the flue gases. For such medium exothermic processes, mass and heat transfer between gas flow and the outer surface of the catalyst grains is sufficient intense. In this case for description of the catalytic process in the reactor is sufficient to use quasi-homogeneous single-phase model.

The model of plug-flow reactor with a fixed bed of catalyst was used to calculate the flow parameters of the gas mixture through a reactor loaded with a composite zeolite-based manganese-oxide catalyst in the process of catalytic CO oxidation.

The calculation results obtained using the software CHEMCAD 7.1.5 were almost identical to the previously made calculation. The peculiarity of the suggested solution was the use of the designed catalytic reactor for treatment of large volume of flue gases with low concentrations of CO at the companies of electrode carbon graphite production.

Keywords: catalytic reactor, carbon monoxide, manganese dioxide, zeolite-clinoptilolite, electrode production, multi-chamber furnace, kiln, electrode blank, plug-flow reactor, fixed-bed reactor.

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1. Introduction

Environmental problems associated with gaseous emissions into the atmosphere have continued to be relevant for several decades. One of the most toxic and multi-tonnage compounds contained in gas emissions is carbon monoxide [1].

Research on the utilization and neutralization of CO is aimed at reducing its concentration or conversion into a less toxic product – carbon dioxide.

The analysis of scientific-technical and patent literature showed that at the present time two methods, based either on physical-chemical methods of extraction, or on thermal or catalytic oxidation are proposed for solving this problem [2, 3]. However, both physical-chemical methods and thermal oxidation are characterized by high energy consumption and capital intensity, while not guaranteeing complete CO removal. Therefore, the most relevant is a thermocatalytic method of carbon monoxide oxidation on heterogeneous catalysts [4, 5].

It is necessary to emphasize, that the technology of catalytic CO oxidation is rather well worked out only for motor transport [6, 7].

It should be noted that a number of requirements are imposed on catalysts for purification of gas emissions, the main of which are chemical and thermal stability, mechanical strength of formed granules and the optimal porous structure, that provides a high rate of mass transfer. This is due to the specific features of thermocatalytic gas purification such as possible overheating of the catalyst when the concentration of oxidizable components increases, aerodynamic overloads [8]. The versatility of modern catalytic methods allows their application for a wide class of gaseous industrial waste.

A serious problem is the replacement of scarce catalysts containing noble metals such as Pt, Pd, Sm [9–11] by cheaper and not less active ones.

Copper-zinc catalysts have confirmed their effectiveness in the processes of gas emissions purification not only from CO, but also from NO_x, ammonia [12]. They also widely applied in processes such as hydrogenation of carbonyl compounds into alcohols, removal of hydrogen sulfide and mercaptans from gaseous and liquid hydrocarbons [13, 14]. It should be noted that single-component CuO-AlO₃ catalysts are practically not used in catalysis due to the reduced stability caused by recrystallization of copper oxide under the impact of reaction media and temperature.

There are many publications on the study of the mechanism and kinetics of the catalytic CO oxidation under various temperature conditions, but there are very few results on the modeling of the process under oxidation conditions in the reactor. A reliable numerical model of the catalytic oxidation system is a powerful way to defined its parameters and optimize the performance of a catalytic reactor, having initial and desired residual CO concentrations and the temperature of gas flow.

Catalytic fixed-bed reactors are widespread in heterogeneous catalytic processes in various areas of the chemical, petrochemical, and refining industries. Thus, the use of a one-dimensional model of the catalytic oxidation of carbon monoxide on hopcalite, taking into account the material and energy balance, allowed to calculate the conversion efficiency of CO to CO₂ depending on the initial temperature of the gas mixture at the entry into the reactor [15]. For this purpose, a system of partial differential equations was composed and solved by the backward Euler method.

A tubular reactor and a two-dimensional model were used to simulate the mass and heat transfer processes in the oxidation of CO to produce H₂ for fuel cells by methane reforming. The mathematical model was described by a series of partial differential equations that were solved numerically using finite element methods with the commercial software package Comsol Multiphysics [16].

The objective of this work was to determine the parameters of equipment for post-combustion of flue gases from industrial kilns at electrode production. For this, the research tasks were

to build a model of CO oxidation in a tubular reactor at composite manganese-zeolite catalyst and based on it to calculate the physical dimensions and hydraulic parameters of the reactor for catalytic CO oxidation for application with the Riedhammer kiln.

2. Materials and methods

A batch of zeolite for building materials with particle size 3–5–8 mm was used in studies of CO oxidation process [17] (Fig. 1, *a*). According to the information from the company, the used zeolite batch was produced from zeolitized clinoptilolite-type tuffs with clinoptilolite $(\text{Na,K})_6[(\text{Al}_6\text{Si}_30)\text{O}_{72}] \cdot 20\text{H}_2\text{O}$ content about 77 %. In order to obtain composite catalyst, its dried 150 g samples were contacted in static mode with 150 cm³ model solutions of Mn (II) and Mn (VII) salts according to the following procedure: a zeolite sample was treated with 0.475 M MnCl_2 solution for 1 hour, then MnO_2 was precipitated on zeolite by adding 0.316 M KMnO_4 solution by reaction:



and kept for 24 hours. The modified zeolite was then separated from the aqueous phase by decantation, washed and air-dried to constant weight (Fig. 1, *b*).

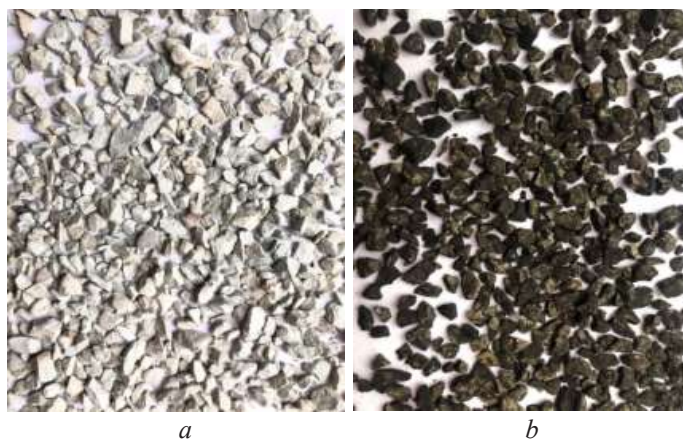


Fig. 1. Appearance of zeolite of Sokyrnytsia deposit: *a* – natural zeolite; *b* – zeolite modified with MnO_2

The calculation of the catalytic CO oxidation reactor was performed for the real flow rate of flue gases of multi-chamber furnaces for baking electrode blanks $V = 18500 \text{ m}^3/\text{h}$ with the temperature 270–390 °C, coming from the «first fire» furnace chamber heated by flue gases. The flue gas passed through 7 fire channels of the furnace chamber. The composition of flue gases was as follows: $[\text{CO}] = 0.13 \text{ \% vol.}$, $[\text{O}_2] = 14.4 \text{ \% vol.}$, $[\text{CO}_2] = 3.71 \text{ \%}$, $[\text{N}_2] = 81.76 \text{ \% vol.}$

The residual CO content in the emission should be less than $250 \text{ mg}/\text{m}^3 = 0.02 \text{ \% vol.}$ Thus, the degree of oxidation of CO should be: $X = (0.13 - 0.02)/0.13 = 85 \text{ \%}$.

The process flow diagram shown in Fig. 2 was used to calculate the catalytic CO oxidation.

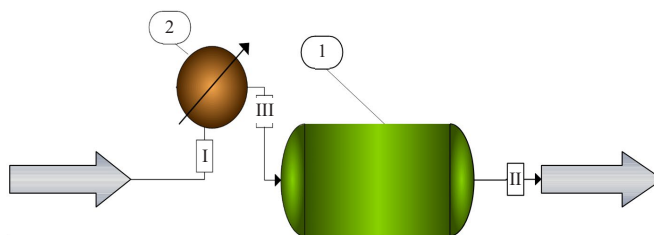


Fig. 2. Process diagram of catalytic oxidation: 1 – kinetic reactor; 2 – heat exchanger; I – gas composition at normal conditions before feeding into kinetic reactor; II – gas composition after the catalyst; III – gas composition at the outlet of heat exchanger at the temperature 350 °C

This flow diagram was also used to confirm the results obtained using the software CHEM-CAD 7.1.5 [18]. The heat exchanger 2 in the scheme was conditionally introduced to heat up the gas to the required temperature.

3. Results and discussion

A catalytic fixed-bed reactor is a heterogeneous system consisting of two phases: solid particles of the catalyst and spaces between them filled with a moving gas. Both chemical transformations on the surface of the catalyst and physical processes, such as transfer of reactants and reaction products inside the bed, heat transfer etc., occur simultaneously in the reactor.

A microcell of a catalyst layer is a volume occupied by approximately one catalyst grain. Such an elementary cell is repeated many times in the layer, and the nature of the processes taking place in the cell largely determines the nature of the processes in the layer. If the size of the elementary cell is much smaller than the size of the layer, the elementary processes occurring in it can be averaged and the average values for each element of the layer can be used. In this case, fluctuations in the values of transfer rate coefficients along the layer height and radius in relation to their average values, which depend on the physical properties of the medium, may be small, so they can be neglected and the coefficients can be considered constant. Oxidation of carbon monoxide can be expressed by the reaction:



It is known that this reaction is exothermic ($\Delta H^\circ_{298} = -282.6$ kJ/mol) and almost irreversible up to 1500 K ($\Delta G^\circ_{298} = -256.7$ kJ/mol; $\Delta S^\circ_{298} = -86.5$ J/(mol·K)).

The full mathematical model of the catalytic fixed-bed reactor takes into account chemical processes and mass and heat transfer processes resulting from concentration and temperature gradients:

- 1) convective transfer of heat and the components of reaction mixture;
- 2) processes of mass and heat transfer within a porous catalyst particle;
- 3) adsorption of reagents and chemical transformation on the inner surface of the catalyst, and desorption of products into the gas phase;
- 4) mass and heat exchange between the gas flow and the outer surface of the catalyst particles;
- 5) longitudinal and radial transfer processes (heat transfer in the solid particles; heat and mass transfer in the gas phase);
- 6) heat removal through the wall (from the flow and catalyst grains).

Mass and heat exchange between the gas flow and the outer surface of the catalyst grains is rather intensive for many medium-exothermic processes; the temperature and concentrations on the outer surface of the catalyst grain are equal to the temperature and concentrations in the flow. A quasihomogeneous single-phase model can be successfully applied in this case to describe the catalytic process in the reactor.

The mechanism of longitudinal and radial mass and heat transfer in the catalyst bed as well as the mechanism of mass and heat transfer inside a grain is described within the diffusion model using the fundamental laws of molecular transfer in a stationary gas medium – Fick's and Fourier's laws with effective diffusion and heat conductivity coefficients.

The catalytic process in the reactor can be considered under the following assumptions:

- longitudinal mass and heat transfer have no significant effect on the concentration and temperature gradients along the layer height;
- diffusion and thermal conductivity coefficients vary insignificantly along the radius of the layer;
- velocity of the gas flow and the heat capacity of the gas flow are constant along the height and cross-section of the reactor.

The mathematical model of the reactor in this case has the following form.

Equations for concentrations of key substances:

$$\frac{D_r^e}{r} \frac{\partial}{\partial r} \left(r \frac{\partial x_i}{\partial r} \right) - u \frac{\partial x_i}{\partial l} - (1 - \epsilon) \sum_{j=1}^{N_r} \mu_{ij} \frac{M_i}{\rho_g} W_j = 0, \quad i = 1, N_k. \quad (1)$$

The temperature equation:

$$\frac{\lambda_r^e}{r} \frac{\partial}{\partial r} \left(r \frac{\partial T}{\partial r} \right) - u c_p \rho_g \frac{\partial T}{\partial l} - (1 - \varepsilon) \sum_{j=1}^{N_r} Q_j W_j = 0. \quad (2)$$

Boundary conditions:

$$r = 0: \frac{\partial T}{\partial r} = \frac{\partial x_i}{\partial r} = 0;$$

$$r = R: \lambda_r^e \frac{\partial T}{\partial r} = \alpha_w (T_w - T), D_r^e \frac{\partial x_i}{\partial r} = 0; \quad (3)$$

$$l = 0: T = T_{in}, x_i = x_i^{in},$$

where D_r^e is the effective diffusion coefficient along the tube radius; u is linear velocity at the full cross-section of the tube; ε is the porosity of the layer (fraction of free layer volume); M_i is the molecular weight of the i -th substance; μ_{ij} is the stoichiometric coefficient of the i -th substance in the j -th reaction; ρ_g is density of gas mixture; N_r is number of reactions; N_k is number of key substances; W_i is observed rate of formation (consumption) of the i -th substance, referred to a unit volume of the grain (taking into account the diffusion resistance of the reaction); x_i, x_i^{in} is concentration of the i -th substance in the layer and at the inlet to the layer; λ_r^e is the effective heat-conductivity coefficient along the tube radius; c_p is heat capacity of the gas mixture; Q_j is heat effect of the j -th reaction; T, T_w, T_{in} is temperatures of the catalyst bed, refrigerant and the initial reaction mixture, respectively; α_w is the coefficient of heat exchange with the refrigerant; l is coordinate along the length of the tube; r is coordinate along the tube radius; R is radius of the tube.

The first term in the material balance equation (1) characterizes the diffusive transport of the i -th substance along the radius of the layer, the second term characterizes the transport of the substance with the flow, and the third term characterizes the rate of chemical transformation on the catalyst grain (taking into account the pore-diffusion resistance). Formula (2) is a heat balance equation. The first term in this equation describes heat transfer along the radius of the layer within the framework of diffusion model, the second term describes heat transfer by flow, and the third term describes heat release or heat absorption accompanying chemical transformation on the catalyst surface.

As will be shown below, the first term in equation (1) can be neglected; the possible presence of diffusion restrictions is taken into account by the experimentally found effective rate constant of the process. Taking into account the low concentrations of CO in the gas flow, as well as the absence of temperature restrictions (moreover, a temperature increase contributes to an increase of the CO oxidation rate), the solution of the heat balance equation (2) can be neglected.

The mathematical model uses the observed kinetics:

$$w_j = \eta W_j^*,$$

where η is the degree of utilization of the inner surface of the catalyst; W_j^* is the rate of formation (consumption) of the i -th substance in the kinetic area.

In practice, the reaction proceeds in the kinetic area in case for layers with small-sized catalyst grains, and $\eta = 1$ can be assumed. For the conditions where pore-diffusion resistance exists in the catalyst grain, η is estimated on the basis of data on the catalyst activity, grains size, and physical properties of the reaction mixture and is averaged over the bed height.

The mathematical description (1)–(3) makes it possible to obtain the main laws of the catalytic process in a fixed-bed reactor:

- to study the influence of hydrodynamic conditions and operating parameters on the fields of concentrations and temperatures in the reactor;
- to determine the main characteristics of the process;
- to define the optimal reactor size, operating conditions for a given production capacity, etc.

The plug-flow reactor is a long channel through which the reaction mixture moves in a piston flow (Fig. 3). Each element of the flow, conditionally separated by two planes perpendicular to the channel axis, moves through it as a solid piston, displacing the previous flow elements and not mixing with either the previous or the following elements.

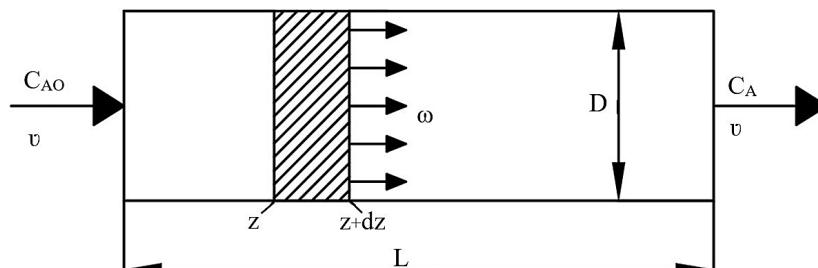


Fig. 3. The plug-flow reactor

In a chemical reaction involving two or more reactants, mixing of the reagents is a prerequisite for the reaction. In a plug-flow reactor, mixing is local, occurs in each element of the flow, and there is no mixing between elements adjacent on the reactor axis.

The following assumptions are made for a plug-flow reactor:

- 1) the moving flow has a flat profile of linear velocities;
- 2) there is no mixing in the direction of the flow axis due to any causes;
- 3) the process parameters are completely equal in every single cross-section, perpendicular to the flow axis.

In a real reactor, it is possible to approach the plug-flow mode if the reaction flow is turbulent and the length of the channel significantly exceeds its lateral dimension.

According to the above-made assumptions, the general material balance equation for the elementary volume of a flow reactor can be simplified. As an elementary volume in this case, it is possible to consider a volume bounded by two parallel planes located from each other at an infinitesimal distance dz and perpendicular to the channel axis z . In this elementary volume according to the third assumption:

$$\frac{dC_i}{dx} = 0, \text{ and } \frac{dC_i}{dy} = 0.$$

Thus, convective transport occurs only in the z -axis direction. According to the second and third assumptions, there is no diffusive transport in a plug-flow reactor. Thus, the material balance equation for a plug-flow reactor in unsteady operation mode can be presented as:

$$-\omega_z \cdot \frac{dC_i}{dz} - r_i = \frac{dC_i}{d\tau}.$$

In the stationary mode of reactor operation:

$$-\omega_z \cdot \frac{dC_i}{dz} - r_i = 0.$$

In a reactor with a constant channel cross-sectional area, the linear flow velocity ω_z will be constant and equal to the ratio of the volume flow rate v to the cross-sectional area F ($\omega_z = v/F$). Then, taking into account that:

$$F \cdot \frac{z}{v} = \frac{V}{v} = \tau,$$

the last equation can be represented as:

$$-\frac{dC_i}{d\tau} - r_i = 0.$$

It can be integrated over $\bar{\tau}$.

$$\bar{\tau} = - \int_{C_{i0}}^{C_i} \frac{dC_i}{-r_i \cdot C_i}.$$

If i is the initial reagent,

$$\bar{\tau} = C_{i0} \int_0^{x_i} \frac{dX_i}{r_i \cdot X_i}.$$

The last equation allows to calculate the contact time of the reactants with the catalyst, taking into account that the process kinetics is described by the first-order equation for CO [19]:

$$r = k_2 C_{CO}.$$

The contact time τ is calculated by equation:

$$\tau = \frac{1}{k} \ln \frac{C_0}{C} = \frac{1}{k} \ln \frac{1}{1-x}.$$

Taking the temperature of the gas mixture is 350 °C, the dynamic viscosity of the gas flow is $\mu = 2.8 \cdot 10^{-5}$ Pa·s, the calculated gas density is 0.56 kg/m³.

The catalyst bulk density with a grains size 5 mm is 1.1146 g/cm³. $E = 31$ kJ/mol, $k_0 = 160$. The porosity of the catalyst is $\varepsilon = 0.53$ [20].

The contact time of the reacting components with the catalyst:

$$\tau = \frac{1}{160 \cdot e^{-\frac{31000}{8.31 \cdot (273+350)}}} \ln \frac{0.13}{0.02} = 4.68 \text{ c.}$$

The catalyst volume can be calculated by the equation:

$$V_k = \tau \cdot V_i,$$

where $V_i = V \cdot [\text{CO}] / 7 = 343.6 \text{ m}^3/\text{hour} = 0.095 \text{ m}^3/\text{c}$.

Then,

$$V_k = 4.68 \cdot 0.095 = 0.45 \text{ m}^3.$$

Because of the low concentration of CO, the heat of the oxidation reaction will be insignificant and the heat calculation can be omitted.

When calculating hydraulic resistance of porous and granular layers formed by catalyst grains, it is necessary at the beginning to determine the gas flow mode through the catalyst layer.

The equivalent diameter of intergranular channels is:

$$d_0 = \frac{2\Phi\varepsilon d}{3(1-\varepsilon)} = \frac{2 \cdot 0.8 \cdot 0.53 \cdot 5 \cdot 10^{-3}}{3(1-0.53)} = 3 \cdot 10^{-3} \text{ m,}$$

where Φ is the particle shape factor ($\Phi = 0.806$ for cube); ε is the porosity of the catalyst; d is the grain size of the catalyst, m.

The velocity of the gas in the channels can be calculated as:

$$w_0 = \frac{w}{\varepsilon}.$$

The volumetric gas velocity through one fire channel is:

$$v = \frac{18500 \text{ m}^3/\text{hour}}{7} = 2643 \text{ m}^3/\text{hour} = 0.734 \text{ m}^3/\text{s,}$$

then the linear gas velocity is equal to:

$$w_0 = \frac{v}{\varepsilon \cdot s},$$

where ε is the porosity of the layer, $\varepsilon = 0.53$; s is the cross section of the catalyst layer, $s = \pi \cdot 0.37^2 / 4 = 0.11 \text{ m}^2$, in which 0.37 m is the equivalent diameter of the container.

Then,

$$w_0 = \frac{0.734}{0.53 \cdot 0.11} = 12.6 \text{ m/s}.$$

Reynolds number:

$$\text{Re} = \frac{w_0 d_0}{\nu} = \frac{12.6 \cdot 3 \cdot 10^{-3}}{3.095 \cdot 10^{-5} / 0.57} = 696.$$

To equalize the radial concentration of CO, it is necessary that the Reynolds number is higher than 30. Thus, our assumption No. 3 for a plug-flow reactor is valid.

The one-dimensional filtration rate of a gas mixture ($\text{m}^3/\text{m}^2 \cdot \text{s}$) through a porous medium is described by the Darcy's law, which well agrees with experimental data for laminar flow:

$$u = \frac{k \Delta p}{\mu L},$$

where $\Delta p = p_1 - p_2$ is pressure drop; L is the medium length; k is the medium permeability; μ is the dynamic viscosity of the gas mixture.

To calculate the hydraulic resistance of the gas flow, the Kozeny-Carman equation was used:

$$\Delta p = \lambda \cdot \frac{H}{d_0} \cdot \frac{\rho \cdot w_0^2}{2},$$

where H is the height of the catalyst bed; λ is the resistance coefficient of the granular layer, which for the average turbulence modes (the turbulence mode occurs at $\text{Re} < 50$ when the gas flows through the granular layer) is determined by the expression:

$$\lambda = \frac{133}{\text{Re}} + 2.34 = \frac{133}{696} + 2.34 = 2.53.$$

Then,

$$\Delta p = 2.53 \cdot \frac{4.2}{3 \cdot 10^{-3}} \cdot \frac{0.57 \cdot 12.6^2}{2} = 1.6 \cdot 10^5 \text{ Pa} = 0.16 \text{ MPa}.$$

One of the constructive solutions to the problem of flue gases treatment of multi-chamber furnaces [21] for burning of electrode billets from carbon monoxide can be placement of containers with catalyst in fire channels of these furnaces in flue gas heated chambers [17]. The size of the fire channel of the furnace to place the catalyst for gas purification is as follows: the height is 5858 mm, the cross-section is rectangular 440×555 mm.

The significant hydraulic resistance of the catalyst bed due to the height of the catalyst bed 4200 mm and possibility of passing of the gas mixture by the catalyst for real industrial conditions (**Fig. 4, a–c**) have to be taken into account, because in the plug-flow mode the equivalent diameter of the container is $d = 370 \text{ mm}$, and the equivalent diameter of the channel is 490 mm. In order to prevent carbon monoxide slip and to reduce hydraulic resistance, the container dimensions 2000 mm high and rectangular section 420×530 mm are taken. Hydraulic resistance for the named above parameter of the equivalent channel diameter $d_c^* = 470 \text{ mm}$ and $\text{Re} = 453$ is $0.32 \cdot 10^5 \text{ Pa} = 0.032 \text{ MPa}$.

Calculation results obtained using the software CHEMCAD 7.1.5 are almost the same as the previously calculated. The difference is that in software version pressure drop over catalyst bed is 0.02 MPa and reaction temperature after catalytic reactor is 360 °C that can be observed when reactor works in adiabatic conditions.

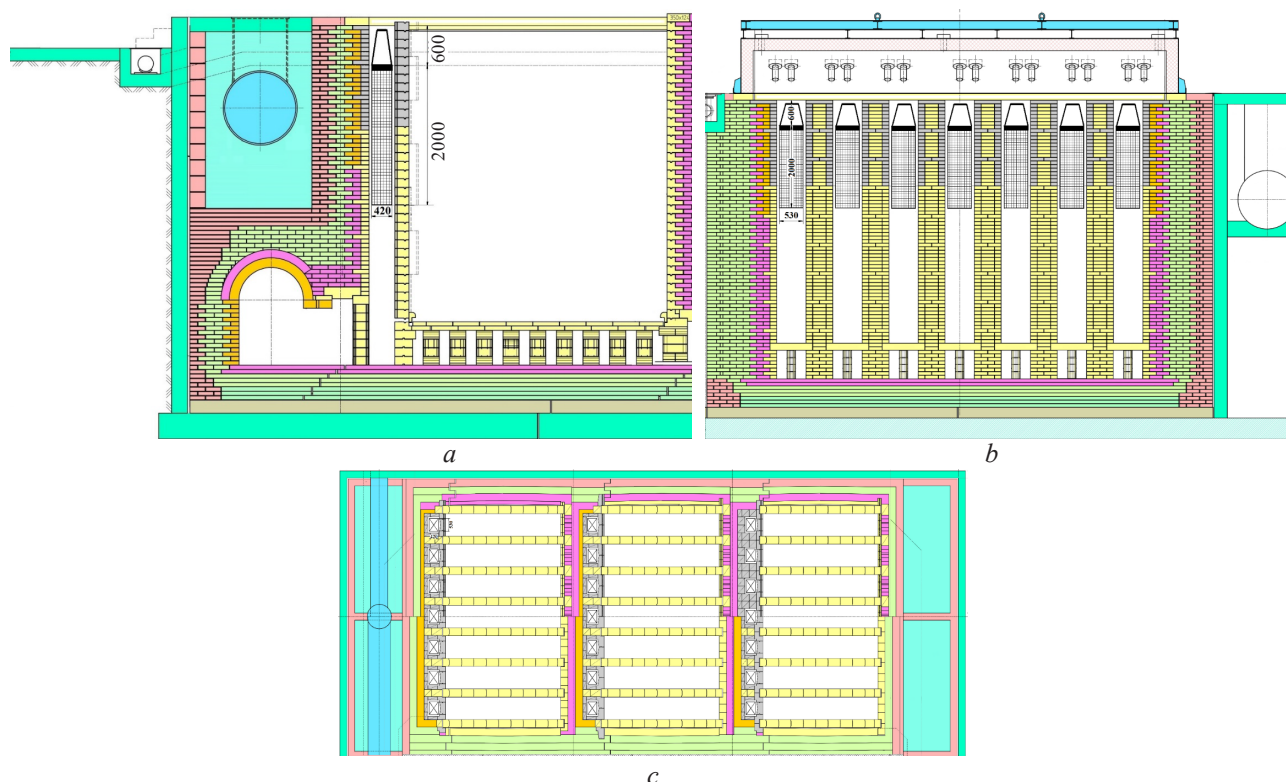


Fig. 4. View of Riedhammer furnace chamber with catalyst containers installed in the fire channels: *a* – front view; *b* – side view; *c* – top view

Constructively, in industrial conditions 7 containers with rectangular cross-section $2000 \times 420 \times 530$ mm are installed in the fire channel at 600 mm depth due to technological reasons of furnace functioning during the whole electrode burning campaign. Containers are made of mesh from stainless-steel wire ASI304, used for treatment of high-temperature flue gases in metallurgical industry according to TU 25.93.13-001-15878725-2018 «Woven metal meshes for special purpose» with a wire diameter 0.9 mm and a cell size 2.8×2.8 mm. Containers are mounted on wire handles, firmly held on cross-shaped metal bars.

It must be noted that the necessary residual concentration of carbon monoxide (250 mg/dm^3) was achieved due to the use of tubular reactors with the composite manganese-zeolite catalyst, while CO concentration in flue gases after the kilns reached 3600 mg/dm^3 . The peculiarity of the proposed solution was the use of the calculated catalytic reactor in conditions of large volumes of flue gases and low CO concentrations at the companies of electrode production. Additionally, the use of an inert material (zeolite) does not promote the loss of carbon material, while the application of manganese dioxide on the backfill of the electrodes accelerates the carbon oxidation, which is an undesirable process [19]. The advantages of the obtained oxide-metal catalyst system are cheapness and availability compared to the use of noble metals.

4. Conclusions

A model of a plug-flow catalytic fixed-bed reactor was developed based on kinetic parameters of catalytic CO oxidation, that made possible to calculate the parameters of the gas mixture flow through a reactor with a composite zeolite-based oxide-manganese catalyst in the process of catalytic CO oxidation. The calculated hydrodynamic parameters of the gas mixture flow (contact time 4.68 s with the manganese-zeolite catalyst volume 0.45 m^3 , linear flow velocity 12.6 m/s, hydraulic resistance of the catalyst bed 0.16 MPa) at a given initial and final CO concentration 0.13 and 0.02 % vol. respectively allowed to determine the parameters of containers with the catalyst.

The developed container with a catalyst provides a technologically acceptable hydraulic resistance to the movement of flue gases, and it is proposed to be placed in the fire channels of

chambers heated by flue gases of multi-chamber Riedhammer kilns for firing electrode blanks, which is an innovative technical solution to the problem of cleaning flue gases from carbon monoxide.

The developed numerical model of the catalytic reactor and design solutions can be used for the designing of gas treatment equipment at environmentally hazardous productions like carbon electrodes, coke, ferroalloys, steel production, where high CO concentration in flue gases is observed.

The conducted research and the results of numerical simulations have shown that the amount of catalyst required for effective purification of exhaust gases from CO takes the entire possible free space of the Riedhammer furnace. In the case of an increase in the concentration of CO, its breakthrough is possible, therefore, further improvement of the catalyst is necessary in order to increase its performance.

Conflict of interest

The authors declare that they have no conflict of interest in relation to this research, whether financial, personal, authorship or otherwise, that could affect the research and its results presented in this paper.

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Data availability

Manuscript has no associated data.

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