CO-FIRING OF GAS COAL DUST FINE PARTICLES
AND SYNTHETIC PEAT GAS.

PART 1. SIMULATION OF PROCESSES OF STEAM-AIR GASIFICATION
OF PEAT IN A FIXED BED AND COMBUSTION OF DUST
AND GAS MIX IN A STREAM

Purpose. To build a model of solid fuel gasification in a fixed bed taking into account the velocity of particles, which allows obtaining detailed information about the technological process. To develop an optimal technological scheme of co-firing of pulverized coal particles and peat gas in the TPP-210A boiler. To investigate the process of a binary mix burning.

Methodology. The objects of research were pulverized coal of Ukrainian deposits and peat from Volyn region deposits. The developed model was used to calculate the peat gasification process. To study the process of burning a binary blend, ANSYS FLUENT package was used.

Findings. A model of solid fuel gasification in a fixed bed has been developed, which differs from the existing models by taking into account the particle velocity and its change. With the aid of this model the detailed information on temperature and concentration fields of the two-phase environment is received. The process of burning binary mix in a TPP-210A boiler is investigated.

Originality. It was found that in the time interval 2200 < \tau < 3200 s a stable stationary process of thermochemical processing of peat with q_{1} = 0.13–0.3 % is formed. The profile of the dispersed phase velocity distribution resembles a mirror reflection of the profile of the particle diameter dependence.

Practical value. Using the constructed model, the process of thermochemical processing of 28 939 kg/h of raw peat in three reactors with a fixed bed at a pressure of 1.5 MPa was numerically investigated. The composition of the generator gas at the reactor outlet was determined.

Keywords: fixed bed, coal, peat, thermal conductivity, steam-air gasification, conductive and radiative heat exchange
The reactor gas to the hopper from reactor 18. The dried peat from the drying chamber 22 enters the gas generator. The coal dust from bunker 1 is fed into the mixer 4 dust feeder 2, where it is mixed with a transport agent 3, and directed into the main burner 5. Cleared from tar and dust, the synthetic peat gas 6 is mixed with evaporated moisture of raw fuel 26 and enters gas turbine 34, wherein the excess pressure is reduced to atmospheric pressure, after which the generator gas is fed into gas burners 8, located 455 mm from the bottom. The air necessary for combustion 29 is also directed there. In this case, generator peat gas contains 22.6 % H2O, which reduces its calorific value. To eliminate these disadvantages, it is proposed to discharge the evaporated moisture 26 and spent drying agent 38 by-passing the boiler through the pipeline 32. In this case, it is advisable to use a regenerative drying chamber 22. In addition, in the cleaning chamber 17 it is necessary to install two recuperative heat exchangers. The first of them is intended to cool the peat gas, resulting in condensation of moisture. The second heat exchanger serves to transfer heat to the dried generator gas extracted in the first heat exchanger. The gas turbine drives the compressor 35 in which the compression of air 28 enriched with oxygen occurs. The steam-air mixture 14 is formed in a mixer chamber 37, where steam 36 and air 28 are supplied. The composition and calorific value of the generator peat gas introduced into the reaction zone can be adjusted by changing the degree of enrichment of the reacting agent with oxygen and water vapor. Unlike the technology of combustion of binary mixtures in the furnaces with wet slag removal [1], where a similar scheme (Fig. 1) is realized, which is used in European countries, the major problem in organizing the process in the furnaces with wet slag removal 27 is to secure the steady leakage of slag through the tap-hole. This problem is solved with installation of gas burners 8 combusting synthetic gas. In the scheme of joint combustion of binary mixture in Fig. 1, the main elements are the gas generator 18 and the combustion chamber of the existing boiler TPP-210A 7. The TPP-210A boiler is equipped with three gasifiers installed in series. Such allocation minimizes the plant space.

**Unsolved aspects of the problem.** In order to obtain detailed information about aerodynamic, thermal and physical and chemical parameters of steam-air gasification of solid fuel under pressure in a fixed bed and the required design characteristics of the plant, it is necessary to develop a mathematical two-dimensional model based on the construction of a two-dimensional system of parabolic and hyperbolic equations of the reacting multiphase medium.

**Purpose.** To construct a solid fuel gasification model in a fixed bed, which differs from existing models by accounting the particle velocity and its changes, which allows obtaining detailed information about the technological process. To develop an optimal technological scheme of joint combustion of coal dust particles and peat gas in the boiler of TPP-210A. To investigate the process of binary mix combustion.

**Methods.** It is assumed that the process of steam-air gasification is quasi-stationary, gas mixture in the gas generator consists of oxygen, carbon dioxide and monoxide, methane, water vapor, hydrogen, nitrogen, acetylene and oxygen-containing compounds, the solid phase is represented by mono-disperse particles of spherical shape, yield of volatile matter causes a decrease in the density of the solid phase at a constant diameter, and heterogeneous combustion and gasification reactions on the surface of the particles cause the decrease in their size at unchanged density, the porosity of the bed does not vary over the vertical axis z, the gas between the particles in the bed moves in the mode of ideal displacement, the stoichiometric scheme of the reactions includes four heterogeneous reactions:

\[
C + O_2 = CO_2 
\]

\[
C + O_2 = CO + C + CO_2 = 2CO
\]

\[
C + H_2O = CO + H_2
\]

\[\text{and five homogeneous reactions:}\]

\[
\text{CH}_1.193\text{O}_{0.4083} \rightarrow 0.205\text{CO} + 0.13\text{CO}_2 + 0.3326\text{CH}_4 + 0.0256\text{H}_2 + 0.375\text{H}_2\text{O}
\]

\[
\text{CO} + 0.5\text{O}_2 = \text{CO}_2
\]

\[
\text{H}_2 + 0.5\text{O}_2 = \text{H}_2\text{O}
\]

\[
\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2
\]

\[
\text{CH}_2 + \text{H}_2\text{O} = \text{CO} + 3\text{H}_2
\]

In the stoichiometric scheme of the reactions, the volatile components of the peat are represented by the oxygen-containing compound CH1.193O0.4083, which, when heated, is decomposed to gas components CO, CO2, CH4, C2H2, H2 and H2O according to the first reaction (2).

The bed is heated up with a high-temperature gas mixture to a temperature of ignition of particles (\(\geq 600 ^\circ C\)) contained in a narrow area adjacent to the bottom. This moment is taken for the beginning of the countdown \(\tau = 0\). Then the heat supply is cut and the feeding starts of a steam-air mixture enriched with O2, with a temperature \(T_{t,0}=350–450 ^\circ C\). Unlike in [2], where a non-stationary process of coal steam-oxygen gasification is studied with a singlefold (non-restorable) loading of a solid phase on a grating lattice, here a quasi-stationary process is considered of peat gasification with a continuous feeding of the raw fuel and ash and slag discharge from the gasifier. As a result, the distribution of phases and concentrations of gas components vary weakly in time.

In existing methods for calculating the fixed bed [2], the velocity of the particles is \(u_p(\tau, z) = 0\), while in the proposed model \(u_p(\tau, z) \neq 0\). Moreover, in the oxidative and reducing zones the disperse phase velocity is changing \(\partial u_p/\partial z \neq 0\) due to the decrease in diameter of particles as heterogeneous reactions (1) progress. This is the main distinction of the proposed
model from existing models [3]. The differential equation describing the change in the peat particles velocity in the oxidizing and reducing zones can be obtained from the continuity equation for the disperse phase [4]

\[
\frac{\partial \rho_{p}}{\partial t} + \frac{\partial (\rho_{p} u_{p})}{\partial z} = - \left( k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w} \right) \frac{6\mu_{c}}{8}
\]

(3)

where \( k \) – reaction rate constant; \( \mu \) – molecular mass; \( u_{p} \), \( \rho_{p} \), \( \delta \) – velocity, density and diameter of a particle; \( C_{w} \) – surface concentration of gas components; \( \tau \) – time.

The boundary condition for the ordinary differential equation (4) is \( u_{z}(z=H_{bed})=u_{0} \). The link between the diameter of the reacting peat particles \( \delta(z, \tau) \) and their velocity \( u_{p}(z, \tau) \) in the indicated zones can be obtained from the ratio of peat consumption at the outlet of the reduction zone and at the cross-section \( z \) belonging to these areas [5]

\[
\frac{(1-e)u_{0}\rho_{p}}{(1-e)u_{p}(z, \tau)\rho_{p}} = \left( \frac{\delta_{0}}{\delta(z, \tau)} \right)^{3}.
\]

(5)

where \( H_{bed} \) – bed height.

In such a setup of the problem, the addend should be added to the equations for thermal balance of the discrete phase [2], to account for the convective heat transfer. Based on this,

\[
(1-e)c_{p} \frac{\partial \rho_{p}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{p} \right) + \left[ (k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w}) \frac{6\mu_{c}}{8} \right] +
\]

(6)

\[
\frac{(1-e)}{\delta} + \frac{(k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w})}{\delta} \frac{6\mu_{c}}{8} - Q_{red} \]

(7)

where \( Q \) – thermal effect of reaction, or thermal flow; \( t_{0} \), \( t_{1} \) – particle and gas temperatures; \( c \) – heat capacity; \( h_{vap} \) – convective heat transfer coefficient.

Converting equation (6) using (3) when \( \frac{\partial \rho_{p}}{\partial t} = 0 \), we obtain a parabolic equation describing the change in thermal energy of the reacting peat particles within the oxidizing and reducing zones [6]

\[
(1-e)c_{p} \frac{\partial \rho_{p}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{p} \right) + \left[ (k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w}) \frac{6\mu_{c}}{8} \right] +
\]

\[
\frac{(1-e)}{\delta} + \frac{(k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w})}{\delta} \frac{6\mu_{c}}{8} - Q_{red}
\]

(8)

where \( V_{p} \) – mass fraction of light substances; \( A_{vap} \) – pre-exponential factor.

The boundary conditions for the first order equation (8) [10], which belongs to the hyperbolic type, are obtained from the ratio: at the bed outlet \( \rho_{p} = H_{bed} = \rho_{0} \), and at that of the reducing zone \( \rho_{p} = \rho_{p} (1 - V_{p}) \) [11]. Further, convert a parabolic equation (6) based on (8)

\[
(1-e)c_{p} \frac{\partial \rho_{p}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{p}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{p} \right) + \left[ (k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w}) \frac{6\mu_{c}}{8} \right] +
\]

\[
\frac{(1-e)}{\delta} + \frac{(k_{c,CO} + 2k_{c,SO}C_{CO}^{w} + k_{c,CO}C_{CO}^{w} + k_{c,CH,O}C_{H,O}^{w})}{\delta} \frac{6\mu_{c}}{8} - Q_{red}
\]

(9)

The continuity equation for gas components can be presented in the following form.

For oxygen

\[
(1-e)c_{p} \frac{\partial \rho_{O_{2}}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{O_{2}}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{O_{2}} \right) +
\]

\[
6(k_{c,CO} + k_{c,SO}C_{CO}^{w}) \frac{(1-e)}{\delta} - 0.5c_{0,CO_{2}} + 0.5c_{0,fi,CO_{2}}
\]

(10)

carbon dioxide

\[
(1-e)c_{p} \frac{\partial \rho_{CO_{2}}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{CO_{2}}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{CO_{2}} \right) +
\]

\[
6(k_{c,CO} + k_{c,SO}C_{CO}^{w}) \frac{(1-e)}{\delta} + r_{CO_{2}} +
\]

(11)

carbon monoxide

\[
(1-e)c_{p} \frac{\partial \rho_{CO}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{CO}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{CO} \right) +
\]

\[
12(k_{c,CO} + k_{c,SO}C_{CO}^{w}) \frac{(1-e)}{\delta} - r_{CO_{2}} - r_{CO_{2}} +
\]

(12)

methane

\[
(1-e)c_{p} \frac{\partial \rho_{CH_{4}}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{CH_{4}}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{CH_{4}} \right) +
\]

\[
+ 0.33266c_{0,CH_{4}} - c_{0,CH_{4}}
\]

(13)

water vapor

\[
(1-e)c_{p} \frac{\partial \rho_{H_{2}O}}{\partial t} + (1-e)\rho_{p} \frac{\partial u_{H_{2}O}}{\partial t} = \frac{\partial}{\partial z} \left( \frac{\lambda_{ph}}{\delta_{0}^{2}} \rho_{H_{2}O} \right) -
\]

\[
6(k_{c,CH_{4}}C_{H_{2}O}^{w})(1-e) + r_{H_{2}O} - r_{CO_{2}} + 0.375c_{0,CH_{4}}
\]

hydrogen

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\[
\frac{\partial C_{\text{H}_2}}{\partial t} + \frac{\partial (u C_{\text{H}_2})}{\partial z} = \frac{\partial}{\partial z} \left[ D_{\text{H}_2,\text{con}} \frac{\partial C_{\text{H}_2}}{\partial z} \right] + \frac{6k_{-1}C_{\text{H}_2}O_{\text{H}_2}^z}{\delta} (1 - \varepsilon) - \varepsilon_{\text{H}_2} + r_{\text{CO}+\text{H}_2} + 3r_{\text{CH}_2} + 0.0256r_{\text{CH}_2,\text{con}}. \]
\]

The momentum and energy conservation equations for the gas phase are

\[
\frac{\partial (p \rho u)}{\partial t} + \frac{\partial (p \rho u^2)}{\partial z} = -\frac{\partial P}{\partial z} + \frac{\partial}{\partial z} \left[ \rho \frac{\partial u}{\partial z} \right]^2 + 150\eta_{\rho}^g(1 - \varepsilon)^2\eta_{\varepsilon}^g 1.75\alpha_{\rho}^g(1 - \varepsilon)^2\mu_{\varepsilon}^g - \frac{\partial}{\partial z} \left[ \frac{\partial \varepsilon}{\partial z} \right] \lambda_{\varepsilon}^g - \frac{\partial}{\partial z} \left[ \frac{\partial \varepsilon}{\partial z} \right] \lambda_{\varepsilon}^g \\
+ \frac{h_{\varepsilon}(t_f - t_{\text{f}_0})}{8} (1 - \varepsilon)^2 + r_{\text{CO}+\text{O}_2} \omega_{\text{CO}+\text{O}_2} + \varepsilon_{\text{H}_2} + \varepsilon_{\text{H}_2,\text{con}} + r_{\text{CO}+\text{H}_2} + r_{\text{CH}_2} \omega_{\text{CH}_2,\text{O}_2},
\]

where \( D_\rightarrow \text{cm} \) – diffusion coefficient of the \( j \) gas component into the mixture; \( u_j, \eta_j \) – velocity and kinematic gas viscosity; \( g \) – acceleration of gravity; \( \lambda_{\varepsilon}^g \) is the effective coefficient of thermal conductivity, \( C_{\text{H}_2}^w, C_{\text{H}_2}^w, C_{\text{H}_2}^w \), and \( C_{\text{H}_2}^w \) – concentration of gas components on the surface of the particle, \( r_{\text{H}_2,\text{O}_2}, r_{\text{CO}+\text{H}_2}, r_{\text{CH}_2}, r_{\text{CH}_2,\text{O}_2}, r_{\text{CH}_2,\text{con}} \) are the reaction rates; along with the yield of volatile matter from peat, they are determined in accordance with [12].

Boundary conditions for parabolic equations (7), (9–20) at the bed inlet (\( z = 0 \)) are determined

\[
\frac{\partial \rho}{\partial z} = C_{\text{CO}_2} = C_{\text{CO}}, \quad \frac{\partial C_{\text{CH}_2}}{\partial z} = \frac{\partial C_{\text{H}_2}}{\partial z} = \frac{\partial C_{\text{O}_2}}{\partial z} = \frac{\partial C_{\text{H}_2,\text{con}}}{\partial z} = 0; \quad \varepsilon_{\text{H}_2} = \varepsilon_{\text{H}_2,\text{con}} = 0; \quad C_{\text{H}_2,\text{O}_2} = C_{\text{H}_2,\text{O}_2}; \quad C_{\text{N}_2} = C_{\text{N}_2}; \quad u_j = u_{\text{f}_0}; \quad t_{\text{f}_0}.
\]

At the outlet from the bed (\( z = H_{\text{bed}} \))

\[
\frac{\partial t_{\text{f}}}{\partial z} = \frac{\partial \rho}{\partial z} = \frac{\partial C_{\text{CO}_2}}{\partial z} = \frac{\partial C_{\text{CO}}}{\partial z} = \frac{\partial C_{\text{CH}_2}}{\partial z} = \frac{\partial C_{\text{H}_2}}{\partial z} = \frac{\partial C_{\text{CH}_2,\text{O}_2}}{\partial z} = \frac{\partial C_{\text{H}_2}}{\partial z} = 0; \quad t_{\text{f}} = t_{\text{f}}(z = H_{\text{bed}}).
\]

The first and second order equations are integrated applying the marching method of direct and reverse runs on an uneven grid, which is thickened in the lower part of the bed [13, 14]. In this case, for the solution of parabolic equations (7), (9–20), the method of iterations was used, while there was no need to use iterations for hyperbolic equations. The pressure gradient was excluded with the well-known SIMPLE method [15].
Second moments \( \{C_{ij}u'v'\} \), \( \{u'^2\} \), \( \{v'^2\} \), appearing in equations (24–26) are calculated on the basis of the Boussinesq hypothesis:

\[
\begin{align*}
\{C_{ij}u'v'\} &= -D_{ij} \frac{\partial C_{i}^{\tau}}{\partial x_{j}}; \\
\{u'^2\} &= -\eta_{x} \left( \frac{\partial u^{'2}}{\partial x_{j}} + \frac{\partial u^{'2}}{\partial y_{j}} \right) + \frac{2}{3} \kappa_{j} \delta_{ij}; \\
\{v'^2\} &= -\eta_{y} \left( \frac{\partial v^{'2}}{\partial x_{j}} + \frac{\partial v^{'2}}{\partial y_{j}} \right) + \frac{2}{3} \kappa_{j} \delta_{ij}.
\end{align*}
\]

The equation of the change in mass, motion and energy of the solid phase along the trajectories of individual particles in the Lagrange representation are written as:

\[
m\frac{\partial V_{j}}{\partial t} = 0.5C\rho \frac{\pi \delta_{j}^{2}}{4} \left[ V_{j} - V_{g} \right] + 0.125\rho_{g} \delta_{j}^{2} \left[ V_{j} - V_{g} \right] \sigma_{j} + F_{v,sg} + F_{v,fr} - \frac{\partial \dot{m}_{j}}{\partial t}.
\]

\[
\frac{\partial \dot{m}_{j}}{\partial t} = \frac{\partial \dot{m}_{j,c,O}}{\partial t} + \frac{\partial \dot{m}_{j,c,SO2}}{\partial t} + \frac{\partial \dot{m}_{j,c,CO}}{\partial t}
\]

\[
+ \frac{\partial \dot{m}_{j,c,H2O}}{\partial t} + \frac{\partial \dot{m}_{j,rad}}{\partial t} + \frac{\partial \dot{m}_{j,super}}{\partial t},
\]

where the first four terms in the left hand side of equation (31) characterize the transition of the solid phase into the gas phase through the heterogeneous reactions \( C + O\_2 = CO\_2 \), \( C + 0.5O\_2 = CO \), \( C + CO\_2 = 2CO \) and \( C + H\_2O = CO + H\_2 \). These components are determined similar to (3). The fifth and sixth terms of the equation are the weight loss rate due to the yield of pyrolysis gases [16] and the evaporation of moisture [17].

Equation of particles energy:

\[
c_{p} \frac{\partial T}{\partial t} = h_{an,l}(t_{l} - T) \pi \delta^{2}_{j} + a_{r} \sigma_{j}^{2} \sigma_{j}(0 - T) + Q_{\Delta}; \]

\[
\theta_{l} = (G/4\pi \sigma_{j})^{1/4}; \quad G = \int_{0}^{\infty} \Omega \, d\Omega;
\]

\[
Q_{\Delta} = \frac{\partial \dot{m}_{j,c,O}}{\partial t} Q_{c,O} + \frac{\partial \dot{m}_{j,c,SO2}}{\partial t} Q_{c,SO2} + \frac{\partial \dot{m}_{j,c,CO}}{\partial t} Q_{c,CO} + \frac{\partial \dot{m}_{j,c,H2O}}{\partial t} Q_{c,H2O};
\]

where \( \Omega \) — angular velocity; \( \Omega \) — direct cosine of radiation; \( \delta_{j} \) — Kronecker symbol; \( \sigma_{j} \) — Stefan-Boltzmann constant; \( \gamma_{j} \) — Reynolds stress tensor component; \( C_{D} \) — aerodynamic drag coefficient; \( m \) — mass; \( k_{f} \) — emissivity factor; \( l \) — length of the mixing path.

**Results.** We will discuss four options of calculating gasification of particles of Volyn peat with a diameter \( \delta_{0} = 0.009 \) m in a steam-air mixture enriched with oxygen, in a stationary bed with the content of fixed carbon (per working mass) \( C_{f} = 18.5124 \% \), volatile matter \( C_{var} = 17.4876 \% \), sulfur \( S = 0.69 \% \), hydrogen \( H = 3.58 \% \), oxygen \( O = 19.6 \% \), nitrogen \( N = 1.36 \% \), moisture \( W = 18.9 \% \), ash \( A = 19.9 \% \), volatile substances (per dry ash mass) \( V = 69.8 \% \), temperature of ash deformation and softening \( t_{d} = 1360 ^\circ \text{C} \) and \( t_{a,soft} > 1400 ^\circ \text{C} \), with particles density \( \rho_{p} = 1450 \) kg/m\(^3\), porosity \( z = 0.4 \) under pressure \( P_{p} = 1.5 \) MPa. Option 1 (per particle gasifier): steam consumption \( B_{H2O} = 667 \) kg/h, oxygen \( B_{O2} = 1795 \) kg/h, nitrogen \( B_{N2} = 2667 \) kg/h, \( u_{O2} = -0.000036 \) m/s, \( u_{N2} = -0.0039 \) m/s, \( D_{bed} = 3.446 \) m, \( H_{bed} = 1.1064 \) m; Option 2 (per particle gasifier): steam consumption \( B_{H2O} = 1795 \) kg/h, \( B_{O2} = 667 \) kg/h, oxygen \( B_{O2} = 2667 \) kg/h, \( u_{O2} = -0.000036 \) m/s, \( u_{N2} = 0.042 \) m/s, \( D_{bed} = 3.446 \) m, \( H_{bed} = 1.1064 \) m; Option 3 (per gasifier): \( B_{H2O} = 667 \) kg/h, \( B_{O2} = 1795 \) kg/h, \( B_{N2} = 2667 \) kg/h, \( u_{O2} = -0.000036 \) m/s, \( u_{N2} = 0.042 \) m/s, \( D_{bed} = 3.446 \) m, \( H_{bed} = 1.1064 \) m; Option 4 (per gasifier): steam consumption \( B_{H2O} = 1795 \) kg/h, \( B_{O2} = 2667 \) kg/h, \( u_{O2} = -0.000036 \) m/s, \( u_{N2} = 0.042 \) m/s, \( D_{bed} = 3.446 \) m, \( H_{bed} = 1.1064 \) m. Temperature of particles at the outlet of the reactor \( t_{p} = 300 ^\circ \text{C} \).

In Fig. 2, temperature profiles of phases are illustrated for Option 1, for the inlet fuel particles velocity \( u_{f,i} = -0.000036 \) m/s. It is seen that the maximum heat generation zone is adjacent to a grate. In this zone, the temperature of gas and that of particles are steeply increasing and reach the maximum values of \( t_{p,max} = 1114–1126 \) and \( t_{p,max} = 799.5–866 ^\circ \text{C} \) (Fig. 2). In the range of temperature under discussion, \( t_{p,max} \) is lower than the temperature of the deformation of ash \( t_{d} = 1360 ^\circ \text{C} \), which allows organizing the process of peat gasification with wet slag removal in the reactor.

The time interval \( 0 < \tau < 2200 \) s is characterized by non-stationary (unsteady) process of thermochemical processing of Volyn peat with distributions of phase temperature and concentration of gas components varying in time. In the oxygen zone \( 0 < z < 0.00146 \) m, there is a sharp increase in the \( t_{p} (\tau, z) \) dependence due to prevailing of heat dissipation in oxidative heterogeneous reactions \( C + O_{2} = CO_{2} \) and \( C + 0.5O_{2} = CO \) and afterburning of gasification products CO and H\(_2\) (CO + \( + 0.5O_{2} = CO \) and \( + 0.5O_{2} = H_{2}O \) over the heat absorption...
in reducing reactions $C + CO_2 = 2CO$ and $C + H_2O = CO + H_2$, resulting in that the temperature of the reacting particles at the outlet of the zone reaches maximum value of $t_{p,max} = 1126^\circ C$ (curve $\beta$ in Fig. 2).

The contents of $O_2$ and $H_2O$ in gas drop from 30 and 19.6 % to 24.9 and 16.9 %, respectively, the volumetric fraction of gas components $CO$, $CO_2$ and $H_2$ increase and achieve the values: $R_{CO} = 0.106$, $R_{CO_2} = 8.33$ and $R_{H_2} = 1.78$ % (curves $\gamma$ in Figs. 3, a, b, a).

In the oxidation-reducing region, $0.00146 < z < 0.0361$ m, oxygen content $R_{O_2} \to 1$ %, the heat absorption in reducing heterogeneous reactions prevails over the release of heat in oxidative reactions. The values $t_p$ and $R_{H_2O}$ decrease to 878 °C and 1.54 %, the volumetric fractions of gas components increase and reach the following values $R_{CO} = 2.1$, $R_{CO_2} = 47.704$ and $R_{H_2} = 12.26$ %.

Gasification zone, $0.0361 < z < 0.171$ m, is characterized by the absence of pyrolysis gases, reduction of the volume ratios of water vapor, carbon dioxide and of the temperature of peat particles to the levels of $R_{H_2O} = 0.043$ %, $R_{CO} = 1.86$ %, and $t_p = 484.5$ °C, the weak growth of values $R_{CO} = 50.23$ and $R_{H_2} = 13.4$ %. Thus it follows that the most extended gasification zone with a small degree of carbon conversion deems inefficient.

In the dry distillation zone, $0.171 < z < 1.1064$ m, heating of peat particles to a temperature of $168\ldots484.5$ °C (curve $\gamma$ in Fig. 2) occurs due to interphase convective heat transfer and radiation-conductive heat transfer of the discharge phase. For this reason, there is an intense yield of volatile substances in the form of an oxygen-containing compound $CH_{1.193}O_{0.4083} \to 0.205CO + 0.13CO_2 + 0.3326CH_4 + 0.1625C_2H_2 + 0.0256H_2 + 0.375H_2O$ with its subsequent partial decomposition to gas components $CO$, $CO_2$, $CH_4$, $C_2H_2$, $H_2$ and $H_2O$, which provides growth of the functions $R_{CH_2}(z, \tau)$, $R_{CH_2O}(z, \tau)$, $R_{CH_2OH}(z, \tau)$ and $R_{CO_2}(z, \tau)$ and downslope in dependencies $R_{CO}(z, \tau)$ and $R_{H_2}(z, \tau)$ in the pyrolysis zone (curves $\gamma$ in Figs. 3, a, b, 4, a and 5). Such behavior of the curves can be explained by the following reasons. In the dry distillation zone, there is a mixing of two gas streams: of pyrolysis gas ($CO$, $CO_2$, $CH_4$, $C_2H_2$, $H_2$, $H_2O$ and $CH_{1.193}O_{0.4083}$), in which the amount of $CH_4$ and $H_2O$ prevails over $CO$ and $H_2$ (First Reaction (2)), and of the synthesis gas coming from the reduction zone, which predominantly consists of $CO$, $N_2$, $H_2$ and a small volumetric fraction of $R_{CO}$.

Therefore, upon mixing of two gas flows the decrease occurs of values $R_{CO}$ and $R_{H_2}$, while the values $R_{CH_2}$, $R_{CO_2}$, $R_{H_2O}$ and $R_{CH_2O}$ are increasing.

Intense yield of volatile matter contributes to a decrease in particle density with unchanged diameter and velocity of the discharge phase (curves $\beta$ in Figs. 4 and 6). In the time interval $0 < \tau < 2200$ s in the area of $0.192 < z < 1.1064$ m the function $p_{\rho}(\tau, z)$ decreases dramatically due to the yield of volatile matter from a value of 1050 to 556.6 kg/m^3, while in the range $0.0361 < z < 0.171$ m the particle density does not change $p_{\rho}(\tau, z) = 556.6$ kg/m^3 (curve $\gamma$ in Fig. 6). In the range of $2200 \leq \tau < 3200$ s, where a dry distillation area is absent, the density $p_{\rho}(\tau, z) = 556.6$ kg/m^3 (curves $\delta - \lambda$).

In Fig. 6, the distribution of peat particle velocities is presented through the height of the fixed bed versus time for Option 1. From the comparison of Figs. 4 and 6 it follows that the discharge-phase velocity profiles $u_\rho(\tau, z)$ resemble the mirror reflection of curves $\delta(\tau, z)$. Indeed, in the pyrolysis zone, the values of functions remain unchanged: $u_\rho(\tau, z) = u_{\rho 0}$ and $\delta(\tau, z) = \delta_{\rho 0}$.

In oxygen, oxidation-reducing zones and in the gasification zone, where an intensive conversion of ash-coke particles occurs with a change in their diameter and speed through heterogeneous oxidation and reduction reactions, the values $\delta_{\rho 0} = 0.009$ m and $u_{\rho 0} = -0.000365$ m/s decrease to $\delta_{\rho} = 7.24$ mm and $u_{\rho} = -0.00019$ m/s.

In the time interval 2200 \leq \tau < 3200 s profiles of each function $t_\rho(\tau, z)$, $\rho_\rho(\tau, z)$, $R_{CO}(\tau, z)$, $R_{CH_2}(\tau, z)$, $R_{CH_2O}(\tau, z)$, $R_{CH_2OH}(\tau, z)$, $R_{CO_2}(\tau, z)$, $R_{CO}(\tau, z)$, $R_{H_2}(\tau, z)$, $R_{H_2O}(\tau, z)$, $p_{\rho}(\tau, z)$, $\delta(\tau, z)$ and $q_{peach}(\tau, z)$ merge into a single curve, indicating the setting of a stationary state of the process (curves $\delta - \lambda$ in Figs. 2, 3 a, b, 4, a, b). With such an organization of the working process, the spare temperature range $t_{\rho} - t_{\rho 0, max} = 224\ldots246$ °C, and the diameter of ash-coke particles at the outlet of the bed is $\delta_{\rho} = 7.24$ mm (Fig. 4, c), which is slightly higher than the diameter of the ash particles $\delta_{ash} = 7.23$ mm, indicating a low mechanical unburnt matter $q_{as} = 0.13\ldots0.3$ % (Fig. 6, c).

At the outlet of the reactor, the generator gas with the following parameters is obtained for Option 1 (per dry mass): $R_{CO} = 33.474$ %, $R_{H_2} = 8.26$ %, $R_{H_2O} = 7.82$ %, $R_{CO_2} = 3.672$ %, $R_{CH_2} = 6.94$ %, $R_{CH_2O} = 19.81$ %, $R_{CH_2OH} = 16.63$ % and $R_{CO} = 3.39$ % at $\tau = 400$ s and $R_{CO} = 52.8$ %, $R_{H_2} = 13.21$ %, $R_{H_2O} = 33.88$ % at $\tau = 3200$ s.
The raw synthetic gas in the amount of 33005 kg/h, obtained in stationary mode (\(\tau \geq 2200\) s) of thermochemical processing of 28939 kg/h of raw peat in three reactors, is directed to chamber 17 (Fig. 1), where the condensation of heavy hydrocarbons \(\text{C}_2\text{H}_2\) in the quantity \(22\sum_{i} B_{\text{C}_2\text{H}_2} = 1754.5\) kg/h and cleaning from disperse particles occurs. Cleared from tar and dust, the peat gas is mixed with 5469.5 kg/h of evaporated moisture of the raw peat and after reset of pressure in the gas turbine enters gas burners 8 (Fig. 1). The mass fractions of the gas components of peat gas at the inlet to the furnace can be presented as: \(\text{H}_2\text{O} = 0.226, \text{H}_2 = 0.00723, \text{CO} = 0.405, \text{CO}_2 = 0.0674, \text{CH}_4 = 0.0608, \text{H}_2\text{S} = 0.00472, \text{N}_2 = 0.229\).

In the furnace of TPP-210A boiler 7, 36720 kg/h of synthetic gas (20 % by fuel energy) and 48996 kg/h of fine coal dust are supplied to the main burners 5 (Fig. 1) for joint combustion. This process in greater detail shall be examined in the third part of this work.

**Fig. 4. Distribution of volumetric fractions of gas components**

- \(H_2\) (a)
- \(H_2O\) (b)
- particle diameter (c) through the height of a stationary bed versus time for Option 1:
  - 1 – 100 s; 2 – 200 s; 3 – 300 s; 4 – 600 s; 5 – 1000 s; 6 – 1500 s; 7 – 1900 s; 8 – 2200 s; 9 – 2400 s; 10 – 2800 s; 11 – 3000 s; 12 – 3200 s

**Fig. 5. Distribution of volumetric particles of gas components**

- oxygen-containing compound (a), acetylene (b) and methane (c) through the height of a stationary bed versus time for Option 1:
  - 1 – 100 s; 2 – 200 s; 3 – 300 s; 4 – 600 s; 5 – 1000 s; 6 – 1500 s; 7 – 1900 s; 8 – 2200 s; 9 – 2400 s; 10 – 2800 s; 11 – 3000 s; 12 – 3200 s

**Conclusions.** A two-dimensional system of parabolic and hyperbolic equations is constructed describing aerodynamics, thermal and mass exchange and chemical reaction of the multiphase medium in a fixed bed. The main difference from the existing models is the accounting of the velocity of particles, its variation in the oxidative and reducing zones due to a decrease in the diameter of the disperse phase in the course of the heterogeneous reactions.

The modeling of the combustion process of gas-disperse multiphase mixture is performed using ANSYS FLUENT (Non-Premixed Combustion model). This model incorporates the Euler-Lagrange description of the medium movement: the transfer of mass, momentum and heat of the gas phase is presented in Euler variables, and the equation of the change in mass, motion and energy of the discrete phase – in Lagrange variables.

The developed model allows obtaining detailed information on the structural characteristics of the gas generator in a fixed
ic gas in amount of 33005 kg/h, generated in the stationary gasification process of the Volyn peat particles (τ ≥ 2200 s), after passing the stage of cleaning from tar and dust, is mixed with 5469.5 kg/h of evaporated moisture from the raw fuel and after resetting excess pressure in the gas turbine \( H_4 \) in the amount of 36720 kg/h enters the gas burner of the combustion chamber of TPP-210A boiler. The mass fractions of the components of this gas at the inlet to the furnace are: \( H_2O = 0.226 \), \( H_2 = 0.00723 \), \( CO = 0.405 \), \( CO_2 = 0.0747 \), \( CH_4 = 0.0608 \), \( H_2S = 0.00472 \), \( N_2 = 0.229 \). This constitutes an initial condition for a joint combustion mode of 36720 kg/h synthetic gas (20 % by fuel energy) and 48 996 kg/h of the pulverized gas coal, which will be described in the third part of this work;

b) In Option 1, where zone of maximum heat release is adjacent to the grate, within time interval 2200 s < τ < 3200 s, a stable (stabilized) stationary process of thermochemical processing of peat is formed with \( q_{g1} = 0.13 – 0.3 \% \);

c) behavior of the distribution of disperse phase velocity resembles a mirror image of the dependence of particle diameter profile. A sharp decrease in the diameter and the speed of ash-coke particles is observed in oxygen, oxidation-reducing zones and in the gasification zone, where intense conversion of ash-coke particles occurs due to heterogeneous chemical reactions.

Analyses of the results of numerical studies on gasification process in the gas generator for Options 2–4 will be given in the second part of this work.

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Спільне спалювання дрібнодисперсного пилу газового вугілля й синтетичного топф'яного газу.

Частина 1. Моделювання процесів пароповітряної газифікації торофу в нерухомому шарі та спалювання пилогазової суміші в потоці

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Meta. Побудувати модель газифікації твердого палива у фіксованому шарі з урахуванням швидкості частинок, що дозволяє отримати детальну інформацію щодо технологічного процесу. Розробити оптимальну технологічну схему спільного спалювання в котлі ТПП-210А пиловугільних частинок і товф'яного газу. Дослідити процес спалювання бінарної суміші.

Методика. Об’єктами досліджень були пилоподібне вугілля українських покладів і волинський тороф. Для розрахунку процесу газифікації товф'яного газу використовувалась розроблена модель. Для дослідження процесу спалювання бінарної суміші – ANSYS FLUENT.

Результати. Розроблена модель газифікації твердого палива у фіксованому шарі, що відрізняється від існуючих моделей урахуванням швидкості частинок та її зміни. За її допомогою отримана детальна інформація про температурні й концентраційні поля двофазного середовища. Досліджено процес спалювання бінарної суміші в котлі ТПП-210А.

Наукова новизна. Встановлено, що у часовому інтервалі 2200 < τ < 3200 с формується стійкий стаціонарний процес термохімічної переробки товфу з $q_t = 0,13–0,3 \%$. Характер розподілу швидкостей дисперсної фази нагадує дзеркальне відображення залежності профілю діаметра частинок.

Практична значимість. За допомогою побудованої моделі чисельно досліджено процес термохімічної переробки 28 939 кг/год сирого товфу у трьох реакторах із фіксованим шаром під тиском 1,5 МПа. Визначено склад генераторного газу на виході з реакторів.

Ключові слова: нерухомий шар, вугілля, тороф, тепло-предоставлення, пароповітряна газифікація, кондуктивний і радіаційний теплообмін

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