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Combined production of synthetic liquid fuel and electricity from coal using H₂S and CO₂ removal systems



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ABSTRACT

The main aim of the research is to continue the studies on promising technologies of coal conversion into synthetic liquid fuel (methanol). The object of study is the plants for combined production of electricity and synthetic liquid fuel (PCPs), which are eco-friendly and more efficient as compared to the plants for separate production. The previous studies on PCPs consider the systems for fine cleaning of gasification products in a simplified way. This study presents the detailed mathematical modeling of the aforementioned systems and determines the values of energy consumption and investment in them. The obtained values are used to carry out the optimization studies and find the optimal parameters of PCPs with different degree of CO₂ removal from gasification products providing fine cleaning of gasification products from H₂S.

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0. Introduction

Implementation of coal projects in the electric power industry is hampered by low energy efficiency and a negative environmental impact of run-of-mine coal combustion, and high costs of delivering energy produced from coal (especially low-calorific coal) to end consumers. Therefore, there is a great interest in the problem of converting coal into synthetic liquid fuel (SLF) as an alternative energy fuel.

One of the most promising directions in coal conversion is coal gasification with subsequent SLF synthesis at plants for combined production of electricity and synthetic liquid fuel (PCPs). An interest in this technology is explained by the fact that SLF synthesis is highly efficient, quite eco-friendly, and allows production of an environmentally friendly fuel. Great attention world-wide and in Russia is paid to the integrated processing of solid fuels into synthetic high-grade fuels with CO₂ removal (Gao et al., 2009; Guang-jian et al., 2010; Hetland and Anantharaman, 2009; Larson et al., 2010; Manzolini et al., 2013; Pellegrini et al., 2011; Robinson and Tatterson, 2008; Sajo et al., 2011; Sun et al., 2010; Winkler, 2007).

Researchers from Energy Systems Institute SB RAS (Russia, Irkutsk) have been engaged in mathematical modeling of PCPs synthesizing various types of SLF (methanol, dimethyl ether, etc.) and

feasibility studies on them for a long time. The mathematical models of individual PCP modules and PCPs as a whole are developed. Optimal scheme and parameter solutions as well as conditions for PCP competitiveness are found (Kler et al., 2005; Kler and Tyurina, 2007; Tyurina et al., 2012).

Comprehensive studies on the processing of organic fuel at PCPs were carried out by such companies as Sasol, Shell, Mobil, Bechtel, Methanex. Nevertheless, the detailed analysis of the conducted studies in the aforementioned directions allows us to find out some unsettled issues.

It should be mentioned, that in the previous studies on PCPs the systems for cleaning syngas from the compounds of sulfur and carbon dioxide were considered in a simplified way using expert data on specific energy consumption and investment. At the same time the catalytic SLF synthesis requires, among other things, the absence of sulfur compounds, because they facilitate “poisoning” of the catalysts and slow down the formation of SLF. It is also necessary to remove excessive carbon dioxide from the gasification products. Since these systems are quite expensive and energy consuming, their simplified representation does not allow us to determine technical and economic indices of the PCP as a whole with the required accuracy.

Therefore, this calls for the mathematical modeling and feasibility studies on the PCP system for cleaning the gasification products.

The specific process analyzed as an example of synthetic liquid fuel production is the synthesis of methanol.

The studies have two stages. At the first stage we construct a mathematical model of the system for cleaning gasification products by the Rectisol method and conduct optimization feasibility

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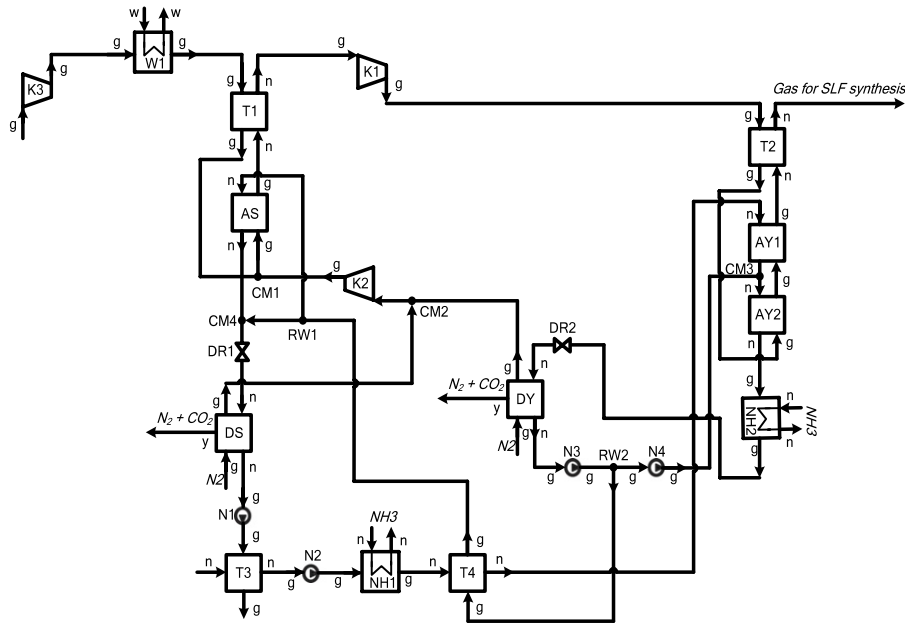


Fig. 1. Calculated scheme of a two-stage system for H₂S and CO₂ removal from gasification products based on the Rectisol method: W1—gas–water heat exchanger; T1–T4—heat exchangers; AS—sulfur absorber; AY1, AY2—CO₂ absorber of stages 1 and 2; NH1, NH2—vaporizers; DY—CO₂ desorber; DS—sulfur desorber; K1–K3—compressors; N1–N4—pumps. Notation of the flows: g—heating flow; n—heated flow; y—components removed from syngas.

studies for this model. At the second stage the developed model is included in the mathematical model of the PCP as a whole in order to make further optimization feasibility studies. The main aim of these studies is to obtain optimal parameters of PCPs with different degree of carbon dioxide removal from gasification products. This is conditioned by the fact that some part of carbon dioxide participates in the reactions of methanol synthesis, which can increase methanol production or the release of additional carbon monoxide in the blast of syngas that enters the combustion chamber of the gas turbine for liquefaction. The second circumstance can affect the generation of additional electricity.

1. Mathematical modeling of the system for cleaning gasification products by the Rectisol method

The developed mathematical model of the cleaning system rests on one of the efficient processes of cleaning gasification products from sulfur compounds and excessive carbon dioxide which is called the Rectisol process (the gas cleaning method with methanol at low temperatures). An interest in this method is explained, firstly, by the fact that during the field tests it proved to be effective in cleaning of syngas obtained by coal gasification, secondly, by the fact that PCPs have low temperature flows, which can be effectively used in the cleaning system, and thirdly, this technique ensures comprehensive removal of CO₂, H₂S, organic sulfur compounds and other impurities from gases with one and the same solvent.

Different cleaning schemes are used depending on gasification method and composition of original gas. In this paper we consider two-stage removal of H₂S and CO₂ from gasification products with cold methanol. Its design scheme (Kler et al., 2010, 2011), for which we have developed a mathematical model, is presented in Fig. 1.

When constructing a mathematical model for the system of H₂S and CO₂ removal from gasification products, we used previously developed models of the system components: heat exchangers, vaporizers, compressors, mixers, etc., and developed new components: absorber and desorber of H₂S and CO₂.

Mathematical model of the absorber includes relationships between input and output parameters of the component (equilibrium compositions and flow rates of absorbent (methanol) and solute

gas (gasification products); pressures and temperatures of flows), and the relationships between these variables and design characteristics of the device.

In the mathematical model of the absorber initial data are represented by the flow rate and composition of the solvent and solute gas, input pressure, temperature and enthalpy. Below we present a system of equations, which describe the mathematical model.

• *Material balance of the process*

$$\Sigma G_{(i)} = G_{L(i)}^{in} + G_{g(i)}^{in},$$

where $\Sigma G_{(i)}$ —total flow rate of the component distributed from gaseous phase to solution in the whole device;

$G_{L(i)}^{in}$ —flow rate of the absorbent (methanol), which contains distributed components, at the inlet to the section;

$G_{g(i)}^{in}$ —flow rate of the gas component at the inlet to the section.

• *Heat balance of the process*

$$G_g^{av} \cdot c_g \cdot (t_g^{in} - t_g^{out}) - G_L^{av} \cdot c_L \cdot (t_L^{out} - t_L^{in}) + Q^{dif} = 0,$$

where Q^{dif} —differential heat of gas dissolution;

G_g^{av} , G_L^{av} —average flow rates of phases in absorber section;

c_L —relative thermal capacity of absorbent;

c_g —relative thermal capacity of solute gas;

t_L^{out} , t_g^{out} —temperature of absorbent and solute gas in the given section;

t_L^{in} , t_g^{in} —initial temperature of absorbent and solute gas.

• When determining the equilibrium composition of solutions in the absorber, we use Henry's law for ideal solutions (since during the process there is no chemical interaction between gas and absorbent, and moderate pressures and low temperatures are used)

$$X_i = k_i \cdot P_i,$$

where X_i —mole fraction of extracted component in the solution;

k_i —the Henry's law constant of the component;

P_i —partial pressure of the gas component.

- Mass transfer surface

$$F = \Sigma G_{(i)} / (\mathcal{K}_g \cdot \Delta \bar{Y}_{av}),$$

where $\Sigma G_{(i)}$ —total flow rate of the component distributed from gaseous phase to solution in the whole device;

\mathcal{K}_g —mass transfer coefficient;

$\Delta \bar{Y}_{av}$ —motive force in concentration units of gaseous phase.

- Mass transfer coefficient

$$\mathcal{K}_g = 1 / (1/\beta_g + m/\beta_L),$$

where β_L and β_g —mass transfer coefficients in liquid and gaseous phases, respectively, $\text{kg}/(\text{m}^2 \cdot \text{c})$;

m —distribution coefficient, kg of methanol/ kg of gas.

- Height of the absorber is determined by geometric relation

$$H = F / (0,785 \cdot a \cdot d_{st}^2 \cdot \psi_a),$$

where F —mass transfer surface;

a —specific surface area of packed bed;

d_{st} —standard diameter of the absorber;

ψ_a —share of the active packed bed surface.

The system of equations is solved by the Newton iteration method with regard to the constraint on heat balance. By solving this system we determine the design characteristics of the absorber.

Mathematical model of the desorber includes the relationships that are necessary to determine equilibrium composition of two-phase mixtures, heat and material balances. The non-negativity constraints on temperature differences are taken into account. Design characteristics are determined (mass transfer surface, height of the desorber stage, etc.).

Thus, *the mathematical model of the system for cleaning gasification products by the Rectisol method* is aimed at the engineering design of PCP components: determination of heating surfaces of heat exchangers, mass transfer surfaces of absorbers and desorbers, driving power of pumps and compressors, thermodynamic parameters, flow rates, etc.

Based on the developed mathematical model of the system for cleaning gasification products by the Rectisol method, we conducted the optimization studies aimed at determining energy consumption and capital costs, depending on the degree of CO_2 removal providing fine cleaning of gasification products from sulfur compounds.

To this end, we solved the problems of non-linear mathematical programming that imply calculation of the parameters (mass transfer surfaces of absorbers and desorbers, etc.), which provide minimum capital investment in the cleaning system at a given price of ceramic acid resistant packing (Raschig rings) and take account of physical and technical constraints on the parameters of the cleaning system and energy consumed for H_2S and CO_2 removal.

Mathematical statement of the problem is as follows

$$\min \Delta \mathcal{K}_{CY}(x, y, \mathcal{K}^{TO}, \mathcal{K}^{AS}, \mathcal{K}^{DS}, \mathcal{K}^{KR}, \Delta N_{CY})$$

subject to

$$H(x, y) = 0;$$

$$G(x, y) \geq 0;$$

$$x_{\min} \leq x \leq x_{\max},$$

where x —vector of independent optimized parameters;

y —vector of dependent (calculated) parameters;

H —vector of equality constraints (equations of material and energy balances, heat transfer, etc.);

G —vector of inequality constraints;

x_{\min}, x_{\max} —vectors of boundary values of optimized parameters;

$\Delta \mathcal{K}_{CY}$ —investment in the system of H_2S and CO_2 removal;

\mathcal{K}^{TO} —investment in heat exchangers;

\mathcal{K}^{AS} —investment in absorbers;

\mathcal{K}^{DS} —investment in desorbers;

\mathcal{K}^{KR} —investment in compressors;

ΔN_{CY} —energy consumption in the cleaning system.

In total we optimize 20 parameters (methanol flow rate in the cleaning system, changes in the enthalpy of cold flows in vaporizers, nitrogen consumption at the desorber inlet, etc.). The system of constraints includes conditions for non-negativity of end temperature differences in heat exchangers, constraints on design temperatures and mechanical stress of heat exchanger tubes, mass transfer surface of absorbers and desorbers, etc. (135 constraints, altogether).

Table 1 presents optimal parameters of the main components of the system for cleaning gasification products by the Rectisol method, depending on the degree of CO_2 removal.

The results of the optimization studies on the cleaning system allow us to construct approximation relationships (Fig. 2) and determine investment and energy consumption for the cleaning system, depending on the degree of CO_2 removal.

The studies on the system for cleaning gasification products by the Rectisol method showed that the lower the degree of CO_2 removal the less the number of absorption stages. This can be explained by a decrease in the flow rate of absorbent required to absorb CO_2 from gasification products (the variant, in which CO_2 removal is 40%, has one absorption stage). For the sake of comparison, the table presents the variant with CO_2 removal equal to 65% (shown in gray), which shows the degeneration of the second absorption stage.

The obtained approximation relationships (Fig. 2) testify to an increase in energy consumption and investment in the cleaning system which is caused by the expansion of mass transfer surfaces of absorbers and desorbers after an increase in the share of CO_2 removal.

The developed mathematical model of the system for removal of H_2S and excessive CO_2 from gasification products by the Rectisol method is included in the mathematical model of PCP as a whole (Fig. 3) in order to carry out further studies.

2. Optimization studies on the plant for combined production of electricity and synthetic liquid fuel (methanol synthesis) with regard for H_2S and CO_2 removal

The mathematical model of PCP is generally intended for engineering design of the plant components and contains about 2000 variables and several hundreds of algebraic and transcendental equations. The system of equations is solved by the Seidel method.

The purpose of optimization studies based on the mathematical model of PCP with a system for cleaning gasification products from H_2S and CO_2 is to obtain optimal thermodynamic and flow-rate parameters of PCPs and their technical and economic indices, depending on the degree of CO_2 removal from gasification products in the cleaning system.

The optimization was performed based on the criterion of minimum prices of the produced SLF with the given levels of internal rate of return and prices of consumed fuel and supplied electricity and with regard for physical and technical constraints on the plant parameters and for costs of the H_2S and CO_2 removal system.

The optimized parameters are enthalpy, pressure and flow rate of live steam, catalyst volume in PCP, etc. The system of constraints contains non-negativity constraints of the end temperature heads of heat exchangers and differential pressures along the once-through part of steam and gas turbines, constraints on design

Table 1

Optimal parameters of the system for cleaning gasification products by the Rectisol method, depending on the degree of CO₂ removal (notation of components corresponds to the notation in Fig. 1).

Component	Index		Degree of CO ₂ removal(%)			
			95	80	65	40
1	2		3	4	5	6
T1	Temperature of the cooled flow, K	Inlet	298.1	298.1	298.1	298.1
		Outlet	266.3	287.2	285.7	277.7
	Temperature of the cooling flow, K	Inlet	242.2	225.1	232.3	236.1
		Outlet	285.1	240.2	250.7	265.1
	Pressure of the cooling flow, kg/cm ²		30.0	30.0	30.0	30.0
	Pressure of the cooled flow, kg/cm ²		30.0	30.0	30.0	30.0
Total area of heat exchangers, m ²		36280.2	5798.2	8794.5	2342.0	
T2	Temperature of the cooled flow, K	Inlet	298.1	298.1	297.8	–
		Outlet	264.7	287.3	285.9	–
	Temperature of the cooling flow, K	Inlet	247.2	257.4	266.6	–
		Outlet	285.5	269.1	278.8	–
	Pressure of the cooling flow, kg/cm ²		50.0	50.0	50.0	–
	Pressure of the cooled flow, kg/cm ²		50.0	50.0	50.0	–
Total area of heat exchangers, m ²		16994.9	2769.6	5959.7	–	
T3	Temperature of the cooled flow, K	Inlet	254.8	275.6	285.2	250.7
		Outlet	231.2	232.7	241.8	237.7
	Temperature of the cooling flow, K	Inlet	227.4	228.8	229.2	237.8
		Outlet	246.8	239.9	241.7	224.1
	Pressure of the cooling flow, kg/cm ²		29.8	29.8	29.8	30.0
	Pressure of the cooled flow, kg/cm ²		30.0	30.0	30.0	20.0
Total area of heat exchangers, m ²		34776.2	5972.8	5620.0	2073.4	
T4	Temperature of the cooled flow, K	Inlet	261.1	264.5	266.9	–
		Outlet	237.6	220.4	227.2	–
	Temperature of the cooling flow, K	Inlet	219.0	192.1	220.2	–
		Outlet	242.0	234.9	259.9	–
	Pressure of the cooling flow, kg/cm ²		50.0	50.0	50.0	–
	Pressure of the cooled flow, kg/cm ²		30.0	30.0	30.0	–
Total area of heat exchangers, m ²		39151.5	5483.1	24857.9	–	
NH1	Temperature of the cooled flow, K	inlet	231.9	233.7	242.5	238.1
		Outlet	219.0	192.1	220.2	231.5
	Temperature of the cooling flow, K	Inlet	107.0	82.5	88.5	83.9
		Outlet	184.0	139.9	193.7	109.2
	Pressure of the cooling flow, kg/cm ²		2.0	1.7	1.7	2.0
	Pressure of the cooled flow, kg/cm ²		50.0	50.0	50.0	30.0
Total area of heat exchangers, m ²		4970.1	2620.5	2851.3	205.9	
NH2	Temperature of the cooled flow, K	Inlet	254.1	263.1	270.4	–
		Outlet	252.9	258.9	257.2	–
	Temperature of the cooling flow, K	Inlet	99.7	118.1	86.2	–
		Outlet	185.4	172.8	156.7	–
	Pressure of the cooling flow, kg/cm ²		2.0	2.1	2.1	–
	Pressure of the cooled flow, kg/cm ²		50.0	50.0	49.6	–
Total area of heat exchangers, m ²		955.8	509.7	1019.8	–	
AS	Absorbent temperature, K	Inlet	237.6	220.3	227.1	231.3
		Outlet	242.3	237.6	242.7	245.0
	Gas temperature, K	Inlet	266.3	240.0	245.1	247.4
		Outlet	242.3	225.3	232.1	236.3
	Absorber mass transfer surface, m ³		21190.7	13417.5	14541.7	9600.5

(continued on next page)

Table 1 (continued)

Component	Index		Degree of CO ₂ removal(%)			
			95	80	65	40
1	2		3	4	5	6
AY1	Absorbent temperature, K	Inlet	242.5	–	–	–
		Outlet	248.1	–	–	–
	Gas temperature, K	Inlet	257.9	–	–	–
		Outlet	247.0	–	–	–
Absorber mass transfer surface, m ³		8810.7	–	–	–	
AY2	Absorbent temperature, K	Inlet	247.1	252.1	261.6	–
		Outlet	254.2	263.2	270.5	–
	Gas temperature, K	Inlet	264.6	265.6	272.9	–
		Outlet	257.9	257.1	266.7	–
Absorber mass transfer surface, m ³		7508.6	5009.7	916.7	–	
DS	Absorbent temperature, K	Inlet	212.5	218.3	222.3	225.6
		Outlet	226.3	227.0	228.7	236.6
	Gas temperature, K	Inlet	315.4	346.3	336.5	266.8
		Outlet	257.3	291.8	291.7	248.6
Desorber mass transfer surface, m ³		13301.6	7760.3	6120.1	4961.0	
DY	Absorbent temperature, K	Inlet	250.7	253.3	256.2	–
		Outlet	261.3	263.3	266.2	–
	Gas temperature, K	Inlet	347.8	368.4	311.6	–
		Outlet	255.7	258.3	261.2	–
Desorber mass transfer surface, m ³		28430.4	6689.4	4858.9	–	
K1	Flow pressure, kg/cm ²	Inlet	29.9	29.9	29.9	–
		Outlet	50.0	50.0	50.0	–
	Flow temperature, K	Inlet	285.2	240.8	250.8	–
		Outlet	311.4	311.0	310.7	–
Consumed power, MW		11274.6	10295.8	10412.1	–	
Total area of built-in gas–water heat exchangers, m ²		839.6	1047.3	1039.6	–	
K2	Flow pressure, kg/cm ²	Inlet	1.2	1.2	1.1	–
		Outlet	30.0	30.0	30.0	–
	Flow temperature, K	Inlet	242.0	222.7	223.5	–
		Outlet	392.4	379.2	380.2	–
Consumed power, MW		2814.8	3859.5	3858.1	–	
Total area of built-in gas–water heat exchangers, m ²		222.0	282.6	281.5	–	
K3	Flow pressure, kg/cm ²	Inlet	20.0	20.0	20.0	20.0
		Outlet	30.0	30.0	30.0	30.0
	Flow temperature, K	Inlet	290.0	290.0	290.0	290.0
		Outlet	308.6	308.5	308.5	308.5
Consumed power, MW		9712.6	9712.6	9712.6	9712.6	
Total area of built-in gas–water heat exchangers, m ²		934.9	934.9	934.9	934.9	
Total investment, mln. dollars			136.9	108.8	128.3	65.1
Total electricity consumption, MW			27.406	24.913	25.053	10.237

temperatures and mechanical stresses of the heat exchanger pipes, constraints on minimum and maximum gasification temperatures, etc. Initial technical and economic information was chosen on the basis of the studies previously conducted at Energy Systems Institute SB RAS (Russia, Irkutsk), which were devoted to the investigation into the technologies of converting solid fuel into synthetic liquid and gaseous fuels and on the basis of the analysis of estimates of expenditures made for process and power plants (Kler

et al., 2005; Kler and Tyurina, 2007; Tyurina et al., 2012). Fuel is gasified in gas generators with a fluidized bed and a dry slag removal system on the basis of steam-oxygen blast under a pressure of 2 MPa. Such a gas generator is analogous to the gas generator Winkler, which was quite thoroughly studied and commercially introduced. These gas generators are most often used in PCPs. The internal rate of return equals 15%, which corresponds to the world practice of studying large-scale projects.

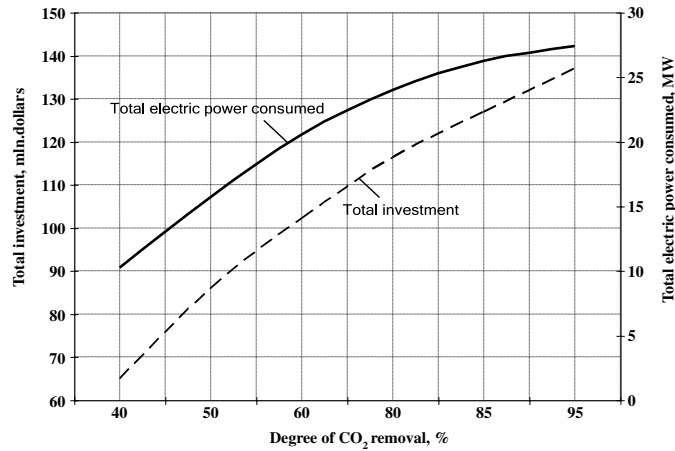


Fig. 2. Relationship between the total investment and total electricity consumption in the cleaning system and the degree of CO₂ removal.

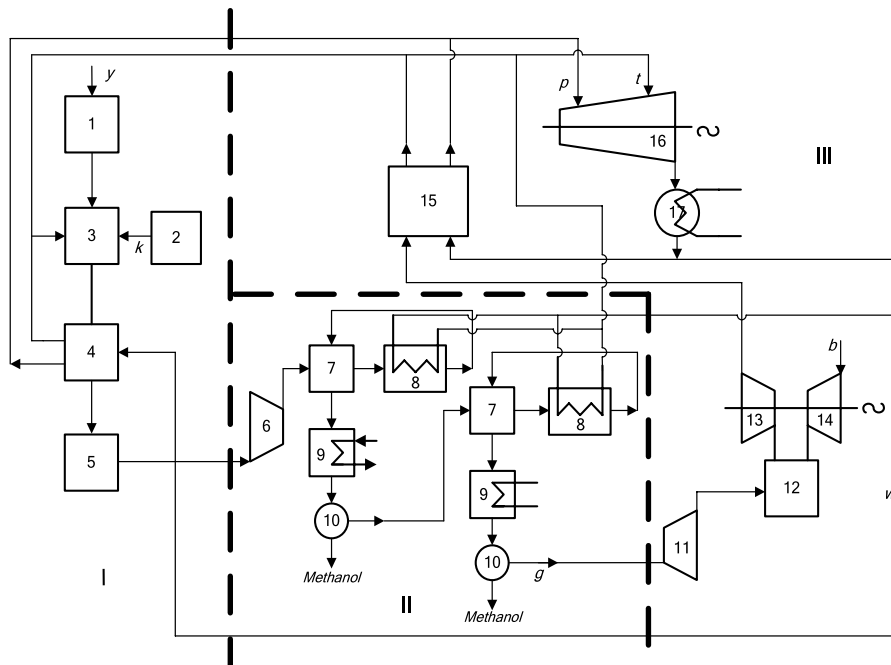


Fig. 3. Simplified process flow sheet of PCP of methanol and electricity on coal: 1—fuel preparation system; 2—air separation system; 3—gas generator; 4—module for cooling gasification products; 5—module for cleaning gasification products by the Rectisol method; 6—syngas compressor; 7—regenerative gas–gas heat exchanger; 8—catalytic reactors of methanol synthesis; 9—methanol condenser; 10—methanol separator; 11—expansion gas turbine; 12—gas turbine combustion chamber; 13—primary gas turbine; 14—air compressor; 15—waste heat boiler; 16—steam turbine; 17—steam turbine condenser. Notation of flows: *g*—gas, *b*—air, *w*—feed water, *y*—coal, *k*—oxygen, *p*—low pressure steam, *t*—high pressure steam. I—module of syngas production, II—module of methanol synthesis, III—module of electricity production.

The mathematical statement of the problem is as follows

$$\min C_{SLF}(x, y, k_m, \Delta \mathcal{K}_{CY}, \Delta N_{CY})$$

subject to

$$H(x, y) = 0;$$

$$G(x, y) \geq 0;$$

$$x_{\min} \leq x \leq x_{\max};$$

$$IRR = IRR_z,$$

where *x*—vector of independent optimized parameters;

y—vector of dependent (calculated) parameters;

H—vector of equality constraints (equations of material and energy balances, heat transfer, etc.);

G—vector of inequality constraints;

*x*_{min}, *x*_{max}—vectors of boundary values of optimized parameters;

*C*_{SLF}—methanol cost;

*k*_m—CO₂ removal coefficient;

$\Delta \mathcal{K}_{CY}$ —investment in H₂S and CO₂ removal system;

ΔN_{CY} —energy consumption in the removal system;

IRR, *IRR*_z—design and given internal rates of return.

Table 2 presents optimal technical and economic indices of PCPs from coal with different degree of CO₂ removal.

Fig. 4 shows optimal PCP indices versus the degree of CO₂ removal from gasification products.

Studying PCPs with a system for cleaning gasification products as a whole, we determined the optimal share of CO₂ removal (Fig. 4). This share corresponds to the variant with 50% removal and is characterized by the lowest price of produced methanol. Both an increase and a decrease in the share of CO₂ removal are typified by a higher price of methanol.

Table 2
Optimal technical and economic indices of PCPs with different degree of CO₂ removal.

Index	Variants of PCPs with different degree of CO ₂ removal(%)		
	25	50	90
Annual consumption of fuel (coal)			
Equivalent, thous. tce	2480		
Natural, thous. t	4580		
Coal price, dollars/tce	20		
Annual production of methanol			
Equivalent, thous. tce	1323.5	1317.5	1255.2
Natural, thous. t	1852.9	1844.5	1757.3
Annual electricity supply, mln. kW h	1549.1	1508	1802
Power, MW			
Gas turbine	317.1	338.4	402.4
Steam turbine	243.01	229.5	246.9
Auxiliary	338.8	352.5	391.9
Effective	221.3	215.4	257.4
Investment in the system for cleaning gasification products, mln. dollars	53.8	76.1	126.1
Total investment in the plant, mln. dollars	1108	963.6	947.2
Thermal efficiency of methanol production, %.	64.6	64	63.5
Price of electricity supplied, cent/kW h	4		
Price of methanol production, dollars/tce	257	225	241

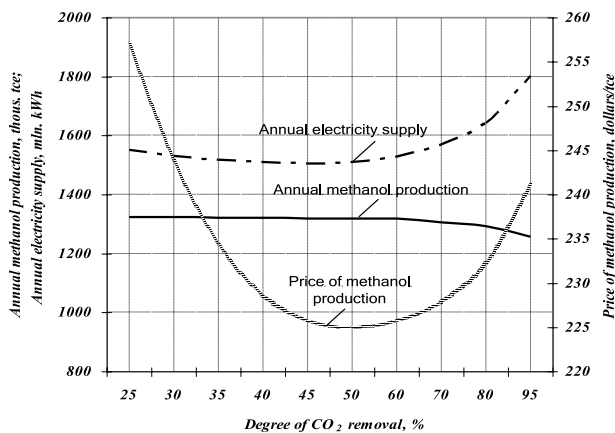


Fig. 4. Dependence of annual methanol production, annual electricity supply, and price of methanol production on the degree of CO₂ removal.

3. Conclusion

The studies on the system for cleaning gasification products by the Rectisol method showed that the lower the degree of CO₂ removal the less the number of absorption stages, which can be explained by a decrease in the flow rate of the absorbent required to absorb CO₂ from gasification products.

As a result of the conducted studies, the developed mathematical models of this system were used to construct the approximation relationships and determine the energy consumption and capital costs, depending on the degree of CO₂ removal providing fine cleaning of gasification products from sulfur compounds. The optimization feasibility studies on the PCP carried out on the basis of the obtained relationships in terms of energy consumption and capital costs for the cleaning system, depending on the degree of CO₂ removal, showed that the optimal variant characterized by the highest efficiency is the variant with 50% removal.

The obtained results and the developed mathematical models for the system for fine cleaning of gasification products from carbon dioxide and sulfur compounds by the Rectisol method can be used in the studies on both PCPs and coal-fired combined heat and power plants in terms of energy consumption and investment in the system or at pre-design and design stages of their development.

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