

Energy-based substantiation of carbon gasification modes in carbon-containing waste

Andriy Helesh^{1*}, Pavlo Saik^{2*}, Nazar Lysyi³, Vasyl Lozynskiy²,
Mykhailo Petlovanyi², Kateryna Sai²

¹ Lviv Polytechnic National University, Lviv, Ukraine

² Dnipro University of Technology, Dnipro, Ukraine

³ Lviv State University of Life Safety Educational and Research, Lviv, Ukraine

*Corresponding author: e-mail andrii.b.helesh@lpnu.ua, saik.p.b@nmu.one

Abstract

Purpose. The research aims to theoretically substantiate the energy-efficient gasification modes of carbon component in carbon-containing waste based on unified thermal self-sufficiency indicators, using coal mining waste as a basic example for model application.

Methods. Comparative material and thermal calculations for four idealized carbon conversion routes within the 660-1000°C temperature range were performed based on previously determined equilibrium constants. All specific indicators refer to 1 kg of initial C. The basic comparison was conducted at an absolute pressure of about 0.2533 MPa. The influence of pressure was investigated for steam and CO₂-gasification within the 0.1013-1.0133 MPa. The comparison of heat requirement q_{req} , resource q_{res} , difference $q_{net} = q_{res} - q_{req}$, the coefficient of potential heat coverage K_{TSS} and the lower heating value of the gas was performed taking into account the residual steam.

Findings. At $p_{abs} = 0.2533$ MPa for $C + H_2O \leftrightarrow CO + H_2$, thermal self-sufficiency coefficient for endothermic processes reaches 2.17 at 860°C, 2.19 at 960°C and slightly decreases to 2.187 at 1000°C. For $C + CO_2 \leftrightarrow 2CO$, it grows to 2.35 at 1000°C. Incomplete air oxidation provides the highest q_{net} , however, the lower heating value of the product gas (LHV_g) remains stable low in the entire examined temperature range of 4.39 MJ/nm³ due to the dilution of combustible gases with air nitrogen (N₂). Increasing p_{abs} from 0.1013 to 1.0133 MPa worsens the equilibrium values of steam and carbon dioxide conversions.

Originality. On the joint basis of 1 kg of initial C, the temperature dependences of the energy potential and heat requirement of idealized gasification routes are quantified. A single system $q_{req} - q_{res} - q_{net} - K_{TSS}$ is proposed for a correct comparison of exo- and endothermic carbon gasification routes. It is shown that the advantage of the route depends on the chosen criterion: H₂ yield, CO formation, volumetric heating value or potential thermal self-sufficiency. The relationship between pressure influence, equilibrium conversion and dilution of the product gas is clarified.

Practical implications. The criteria for pre-selection of the route are proposed, based on the intended use of gas and available heat resources. For steam gasification at the base pressure, the initial test range of 860-900°C is recommended; for target production of CO in the presence of high-temperature heat – CO₂-conversion in the 900-1000°C range is recommended. For the hydrogen product, it is advisable to combine steam gasification with water-gas shift reaction, and for the autothermal scheme – to determine the share of oxidant introduced into the system.

Keywords: coal mining waste; gasification; co-gasification; carbon conversion; synthesis gas; specific heat requirement; heat resource; temperature; absolute pressure

Nomenclature

T – temperature, °C;

p_{abs} – absolute pressure, MPa;

X_C – carbon conversion, %;

$n_{C,0}$ – initial molar amount of carbon, mol;

$n_{C,out}$ – molar amount of unreacted carbon at the outlet, mol;

V_g – specific gas volume / gas yield, Nm³ kg⁻¹ C;

$q_{h,react}$ – specific heat for reactant heating, MJ kg⁻¹ C;

$q_{h,inert}$ – specific heat for inert-component heating, MJ kg⁻¹ C;

q_{endo} – specific endothermic reaction heat, MJ kg⁻¹ C;

q_{exo} – specific exothermic reaction heat, MJ kg⁻¹ C;

q_{req} – specific heat demand, MJ kg⁻¹ C;

q_{res} – specific heat resource, MJ kg⁻¹ C;

q_{net} – net specific heat, MJ kg⁻¹ C;

K_{TSS}^{endo} – thermal self-sufficiency coefficient for endothermic processes;

K_{TSS}^{exo} – thermal self-sufficiency coefficient for exothermic processes;

LHV_g – lower heating value of product gas, MJ Nm⁻³;

ΔH_r – reaction enthalpy, kJ mol⁻¹;

Nm³ – normal cubic metre, m³;

H₂ + CO – combustible components, vol.%.

Received: 10 June 2026. Accepted: 21 September 2026. Available online: 26 September 2026

© 2026. A. Helesh et al.

Mining of Mineral Deposits. ISSN 2415-3443 (Online) | ISSN 2415-3435 (Print)

This is an Open Access article distributed under the terms of the Creative Commons Attribution License (<http://creativecommons.org/licenses/by/4.0/>), which permits unrestricted reuse, distribution, and reproduction in any medium, provided the original work is properly cited.

1. Introduction

The accumulation of coal mining waste remains one of the most complex environmental and resource problems of mining regions. Rock dumps contain mineral fraction, residual carbon, sulfur-containing compounds and impurities of metals, so their long-term storage is accompanied by risks of spontaneous combustion, dust pollution, acid drainage and secondary pollution of soils and water bodies [1]-[4]. At the same time, the residual carbon component of such waste can be considered as a secondary energy resource suitable for thermochemical conversion.

Thermochemical processing of coal mining waste makes it possible to combine the ecological purpose of waste disposal with the resource use of the carbon phase [5], [6]. Recent characterization of coal mining waste has also shown that the predominance of medium and fine-sized fractions can provide favorable conditions for heat and mass transfer during thermochemical conversion [7]. A two-stage scheme involving low-temperature thermal decomposition of the organic fraction and high-temperature gasification of the carbon residue, is promising to produce a gaseous energy carrier [8], [9]. The proposed approach can be applied not only to coal mining waste, but also to other carbon-containing waste, in which after thermal conversion a solid carbon residue is formed, suitable for further gasification.

Experience gained from underground coal gasification demonstrates that the conversion of solid carbonaceous material is governed by a combination of thermochemical and technological factors. Previous studies have addressed thermal and material balances, the formation of temperature fields and their interaction with the surrounding medium, as well as the influence of gasifier configuration and gasifying-agent supply on process performance [10]-[13].

Further development of gasification technologies has also shown that process intensification and reaction evolution require coordinated control of operating conditions [14]. Although these studies concern in-situ coal conversion, their findings emphasize the need to distinguish thermochemical characteristics from reactor and technological effects when assessing gasification modes for carbon-containing materials.

Against this broader technological background, previous studies by the research team focused on the thermodynamic substantiation of coal mining waste gasification, analysis of pressure influence and general assessment of thermochemical utilization of such waste [15]-[19]. The relevance of this approach is consistent with recent reviews on carbon-containing waste gasification, where the determining factors are temperature, pressure, the nature of gasifying agent, the mineral composition of raw materials and catalytically active impurities [20], [21]. Of particular importance is the combination of pyrolysis stages, gasification of carbon residue CO_2 or H_2O and subsequent adjustment of the H_2/CO ratio through a water-gas shift reaction, which makes it possible to move from simple waste disposal to controlled formation of fuel or process gas [22]. However, the very fact of thermodynamic probability of a reaction does not yet determine the energy feasibility of the process. For the technology of coal mining waste gasification, it is necessary to assess not only the temperature range in which the equilibrium shifts towards CO and H_2 , but also the maximum possible specific heating value of the gas, the specific heat balance of the process and the quality of the resulting gas.

In this context, not only equilibrium indicators are important, but also criteria of technical-economic feasibility. Studies of small-scale gasification energy systems show that the economic result depends significantly on the cost of raw materials, capital costs, operation mode, gas quality, the need for its purification and the potential for heat recovery [23]. For the initial selection of modes, it is advisable to compare the increase in the energy potential of gas with the additionally required heat. Further economic assessment should separately take into account the cost of equipment, preparation of raw materials, gasifying agents, gas purification and heat recovery of products.

A preliminary thermodynamic analysis of coal mining waste shows that endothermal carbon gasification reactions exhibit a pronounced temperature dependence: with increasing temperature, the value ΔG decreases and the equilibrium constants of chemical reactions $\text{C}_{(s)} + \text{H}_2\text{O}_{(g)} \leftrightarrow \text{CO}_{(g)} + \text{H}_{2(g)}$, $\text{C}_{(s)} + 2\text{H}_2\text{O}_{(g)} \leftrightarrow \text{CO}_{2(g)} + 2\text{H}_{2(g)}$ and $\text{C}_{(s)} + \text{CO}_{2(g)} \leftrightarrow 2\text{CO}_{(g)}$ grow [15]. Therefore, the temperature range above 860°C can be considered as the range within which the equilibrium conditions for CO - and H_2 -containing gas formation are the most favorable.

At the same time, temperature intensification has a technological limit. The increase in temperature requires additional heat supply, increased thermal insulation, the use of heat-resistant materials, steam superheaters and a stable supply system for gasifying agents [24]-[26]. Therefore, the criterion "the higher the temperature, the better" for the industrial process is incorrect. From an energy perspective, a rational mode is one in which the increase in the energy value of the resulting fuel gases exceeds the additional energy expenditure required to raise the process temperature.

A similar conclusion follows from the equilibrium modeling of biomass gasification: the increase in temperature to the range of about 900°C improves the energy and environmental performance of the gas, and water vapor as a gasifying agent provides a substantially higher share of H_2 compared to air gasification [27]. At the same time, the further temperature increase should be assessed taking into account the increase in heat consumption, the risk of slagging, the lining resource and the requirements for gas purification.

The purpose of this research is to provide a theoretical substantiation of energy-efficient gasification modes of carbon component in carbon-containing waste on the basis of previously obtained equilibrium constants, summary material calculations of the composition of conversion gases and a unified system of specific heat parameters normalized to 1 kg of carbon. Coal mining waste is used as a basic model raw material for quantitative analysis.

2. Materials and methods

2.1. Calculation basis and investigated reactions

The object of theoretical analysis is the carbon component of carbon-containing waste after preliminary thermal conversion or pyrolysis. Coal mining waste is considered as a basic example of technogenic raw materials with increased residual carbon content and favorable granulometric characteristics for thermochemical conversion.

Steam, carbon dioxide and air-oxidative conversion were investigated in the $660\text{-}1000^\circ\text{C}$ temperature range. The basic comparative absolute pressure was $p_{abs} = 0.2533\text{ MPa}$. The

influence of pressure was investigated for steam gasification of carbon (R₁ route) and CO₂-gasification (R₃ route) at $p_{abs} = 0.1013; 0.2533; 0.5066$ and 1.0133 MPa.

The method is a continuation of the two previous stages: thermodynamic determination ΔH_r , ΔG and equilibrium constants [15] and material calculations of equilibrium composition, X_C and V_g depending on temperature and absolute pressure [16]. The thermochemical functions of the components were determined using established reference ratios and databases [28]–[30].

All specific indicators are referred to 1 kg of carbon, fed into the process, including its unconverted residue. The initial molar ratios are $H_2O/C = 1$ for R₁, $H_2O/C = 2$ for R₂ and $CO_2/C = 1$ for R₃. For R₄, oxygen of air is assumed to act as an oxidizer, so the corresponding ratios $O_2/C = 0.5$ and $N_2/O_2 = 3.76$ were used for calculations. Conversion rate is defined as $X_c = \frac{(n_{C,0} - n_{C,out})}{n_{C,0}}$. In the tables, it is given as a

percentage, and in formulas – as fraction of a unit. The component volumes and total gas yield are normalized to the same mass of initial C. The main reactions used in the calculations are given in Table 1. Here, R₁–R₄ routes are considered as separate idealized stoichiometric routes for comparative material and energy assessment. R₂ is the sum of the steam gasification of R₁ and the water-gas shift reaction $CO + H_2O \leftrightarrow CO_2 + H_2$. Its separate consideration characterizes the route of increased H₂ formation, rather than an equilibrium in the common gas phase independent of R₁. R₄ is an idealized limit of selective oxidation of C to CO by air. Complete conversion and absence of CO₂ in this route are model assumptions.

The specific heat values of heating were determined as the product of specific enthalpies of components at the appropriate temperatures per their masses. Table 2 gives the enthalpy values used in calculations, chosen from literary sources [28]–[30].

Table 1. The main reactions used in calculations

Route	Reaction	Heat character	Technological role
R ₁	$C_{(s)} + H_2O_{(g)} \leftrightarrow CO_{(g)} + H_{2(g)}$	endothermic, $\Delta H_r > 0$	obtaining H ₂ - and CO-containing synthesis gas
R ₂	$C_{(s)} + 2H_2O_{(g)} \leftrightarrow CO_{2(g)} + 2H_{2(g)}$	endothermic, $\Delta H_r > 0$	increase in share of H ₂ to form CO ₂
R ₃	$C_{(s)} + CO_{2(g)} \leftrightarrow 2CO_{(g)}$	endothermic, $\Delta H_r > 0$	obtaining CO-containing gas and CO ₂ utilization
R ₄	$C_{(s)} + 0.5(O_2 + 3.76N_2)_{(g)} \rightarrow CO_{(g)} + 1.88N_{2(g)}$	exothermic, $\Delta H_r < 0$	thermal compensation of endothermic processes

Table 2. Specific enthalpies of reactants in the studied temperature range, kJ/kg

Reactant	Temperature, °C								
	660	680	700	760	800	860	900	960	1000
Carbon	1145.21	1192.69	1240.14	1382.13	1476.20	1616.10	1708.47	1845.69	1941.97
Air	697.27	719.79	742.32	810.98	856.62	926.54	973.43	1043.77	1088.75
CO ₂	696.26	720.55	744.83	818.94	868.34	931.14	993.95	1070.98	1115.39
H ₂ O, $p_{abs} = 0.1013$ MPa	3836.78	3882.00	3927.22	4064.55	4157.49	4299.84	4395.30	4538.49	4628.70
H ₂ O, $p_{abs} = 0.2533$ MPa	3836.36	3881.16	3926.38	4064.55	4157.07	4299.42	4394.88	4538.49	4628.60
H ₂ O, $p_{abs} = 0.5066$ MPa	3834.69	3879.91	3925.13	4062.87	4156.24	4298.59	4394.05	4537.65	4628.00
H ₂ O, $p_{abs} = 1.0133$ MPa	3831.52	3876.86	3922.25	4060.42	4154.02	4296.67	4392.37	4536.22	4626.90

2.2. Unified thermal indicators of exo- and endothermic processes

For endothermic routes, the specific heat requirement was determined as the sum of heats for heating reactants, inert components and the endothermic effect of the reaction (positive in absolute value), with all terms normalized to the initial carbon mass:

$$q_{req}^{endo} = q_{h,react} + q_{h,inert} + |q_{endo}|. \quad (1)$$

The specific heat resource of endothermic processes was calculated as the sum of lower heating values of H₂ and CO formed:

$$q_{res}^{endo} = V_{H_2} \cdot LHV_{H_2} + V_{CO} \cdot LHV_{CO}. \quad (2)$$

To calculate the exothermic incomplete oxidation of carbon to CO_(g), the heat was determined as the sum of heats for heating carbon, air, and the exothermic effect of the reaction was included in the heat resource:

$$q_{req}^{exo} = q_{h,react} + q_{h,inert}; \quad (3)$$

$$q_{res}^{exo} = V_{H_2} \cdot LHV_{H_2} + V_{CO} \cdot LHV_{CO} + q_{exo}. \quad (4)$$

For both groups of processes, the same integral indicators were used:

$$q_{net} = q_{res} - q_{req}; \quad (5)$$

$$K_{TSS}^{endo} = \frac{q_{res}^{endo}}{q_{req}^{endo}}; \quad K_{TSS}^{exo} = \frac{q_{res}^{exo}}{q_{req}^{exo}}. \quad (6)$$

In this paper, K_{TSS} characterizes the potential coverage of the calculated heat requirement by the total energy resource. For R₁–R₃, this resource is only realized by burning the resulting combustible gas or other method of converting its chemical energy into heat. For R₄, q_{res} has two different contributions: oxidation heat in the reactor and chemical energy of the CO formed.

The calculations accepted $LHV_{H_2} = 10.79$ MJ/ Nm³ and $LHV_{CO} = 12.64$ MJ/ Nm³. The lower heating value of product gas was determined by the ratio of thermal contribution of combustible components to the total wet gas volume. CO₂, N₂ and water vapor do not form the heating value and reduce LHV_g .

2.3. Assumptions and limits for application

Calculations are theoretical in nature and describe idealized gasification routes within the accepted set of reactions and components. The model does not take into account competing reactions, kinetic and diffusion constraints, reactor hydrodynamics, the influence of the mineral fraction of raw

material, as well as heat losses to the environment [31]–[33]. The physical heat of the product gas was also not taken into account, since the degree of its use depends on the specific scheme of heat recovery, steam superheating and preheating of raw materials. Therefore, the above values should be considered as comparative model estimates, and the determined temperature ranges should be considered as guidelines for further experimental verification.

The actual thermal self-sufficiency of the plant is determined by the total heat balance, heat transfer efficiency, the degree of heat recovery, heat losses and the fraction of product gas used for internal heat supply [34], [35]. In this regard, $K_{TSS} > 1$ is only a theoretical indicator of potential thermal self-sufficiency, and $q_{net} > 0$ is the estimated heat resource margin, not the guaranteed useful heat yield of the plant. At the same time, these indicators make it possible to accurately compare alternative gasification routes, determine the rational temperature range and assess their potential for autothermal implementation under the adopted model assumptions and calculation basis.

3. Results and discussion

3.1. Summary results of material recalculations

The aggregated values for X_C , V_g and the composition of gases are formed based on previously published thermodynamic and material calculations [15], [16]. They are the starting point for new thermal recalculations and make it possible to determine what portion of the increase in q_{res} is caused by an increase in conversion rate, and what portion – by a change in gas composition. Summary results of material recalculations depending on the temperature are given in Table 3.

For R_1 , the conversion rate increases from 39.71 to 96.48%, and the total share of $H_2 + CO$ – from 56.85 to 98.21 vol.%. For R_3 , the most pronounced change is observed after 760°C: the share of CO increases from 53.98 vol.% at 760°C to 80.13 vol.% at 860°C and 95.47 vol.% at 1000°C. Incomplete air oxidation has a constant theoretical composition of 34.72 vol.% CO and 65.28 vol.% N_2 , so the temperature rise does not improve the LHV_g of this gas.

Table 3. Summary results of material recalculations depending on the temperature

No.	Route/ reaction	Indicator	Temperature, °C								
			660	680	700	760	800	860	900	960	1000
1	$C + H_2O \leftrightarrow CO + H_2$	X_C , %	39.71	45.67	51.73	68.79	77.89	87.34	91.33	95.00	96.48
		V_g , Nm ³ /kg C	2.608	2.719	2.832	3.151	3.321	3.497	3.571	3.640	3.668
		H ₂ , vol.%	28.43	31.35	34.09	40.75	43.79	46.62	47.73	48.72	49.10
		CO, vol.%	28.43	31.35	34.09	40.75	43.79	46.62	47.73	48.72	49.10
		H ₂ O, vol.%	43.15	37.30	31.82	18.49	12.43	6.76	4.53	2.56	1.79
2	$C + 2H_2O \leftrightarrow CO_2 + 2H_2$	X_C , %	57.55	60.13	62.56	69.03	72.67	77.23	79.75	82.89	84.61
		V_g , Nm ³ /kg C	4.808	4.856	4.901	5.022	5.090	5.175	5.222	5.281	5.313
		H ₂ , vol.%	44.69	46.23	47.65	51.32	53.30	55.72	57.02	58.60	59.46
		CO ₂ , vol.%	22.35	23.12	23.83	25.66	26.65	27.86	28.51	29.30	29.73
		H ₂ O, vol.%	32.96	30.65	28.52	23.02	20.05	16.43	14.48	12.10	10.81
3	$C + CO_2 \leftrightarrow 2CO$	X_C , %	14.45	17.88	21.85	36.96	48.99	66.85	76.70	87.03	91.34
		V_g , Nm ³ /kg C	2.136	2.200	2.275	2.557	2.781	3.115	3.298	3.491	3.572
		CO, vol.%	25.26	30.33	35.86	53.98	65.76	80.13	86.81	93.06	95.47
		CO ₂ , vol.%	74.74	69.67	64.14	46.02	34.24	19.87	13.19	6.94	4.53
4	$C + 0.5(O_2 + 3.76N_2) \rightarrow CO + 1.88N_2$	X_C , %	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
		V_g , Nm ³ /kg C	5.376	5.376	5.376	5.376	5.376	5.376	5.376	5.376	5.376
		CO, vol.%	34.72	34.72	34.72	34.72	34.72	34.72	34.72	34.72	34.72
		N ₂ , vol.%	65.28	65.28	65.28	65.28	65.28	65.28	65.28	65.28	65.28

3.2. Influence of temperature on thermal indicators

The graphical dependences presented in this section are based on the unified indicators of q_{req} , q_{res} , q_{net} and K_{TSS} at $p_{abs} = 0.2533$ MPa. As the temperature increases, q_{req} increases for all routes, however, the reasons and rates of this increase differ (Fig. 1). For R_1 – R_3 , an endothermic contribution is added to the heating of the flows, the value of which depends on the conversion rate. R_2 has the highest expenditures. At 860 and 1000°C, they are 20.79 and 22.70 MJ·kg^{−1} C, respectively. This is due, in particular, to the double stoichiometric water supply relative to R_1 . For R_3 , the intensive growth of q_{req} in the medium- and high-temperature regions is due to a significant increase in the growth rate of the converted C (Table 3). For R_1 , within the 860–1000°C temperature range, the intensity of increase in heat requirement decreases as the conversion approaches its maximum value and the energy required to compensate for the endothermic nature of the process decreases. The lowest value of q_{req} is observed for the exothermic R_4 route, since the process and heat are only spent on heating the reactants (C and air).

For endothermic routes, q_{res} is determined by the number of combustible components formed, not only by their volume fraction (Fig. 2). In the R_1 , the conversion increase raises the yields of H_2 , CO and q_{res} to 38.20 MJ·kg^{−1} C at 860°C and 42.20 MJ·kg^{−1} C at 1000°C. At the top of the range, the increase slows down. R_3 has the most pronounced relative increase within the examined range: the increase in CO yield is ensured by $q_{res} = 31.55$ at 860°C and 43.10 MJ·kg^{−1} C at 1000°C. At temperatures above 970°C, the specific heat resource value on the R_3 route exceeds that for R_1 route. Route R_2 forms more H_2 , but does not form CO; therefore, at high temperatures, its total energy potential is lower than that of R_1 and R_3 . For R_4 , the resource is almost constant, since the conversion and yield of CO in the studied temperature range remain virtually unchanged.

Parameter q_{net} represents the difference between the accepted resource and heat requirement (Fig. 3). For R_1 , it increases from 20.62 MJ·kg^{−1} C at 860°C to 22.90 MJ·kg^{−1} C at 1000°C, which is approximately by 11.1%. For R_3 , over the same range, the increase is about 42.4%: from 17.39 to 24.77 MJ·kg^{−1} C.

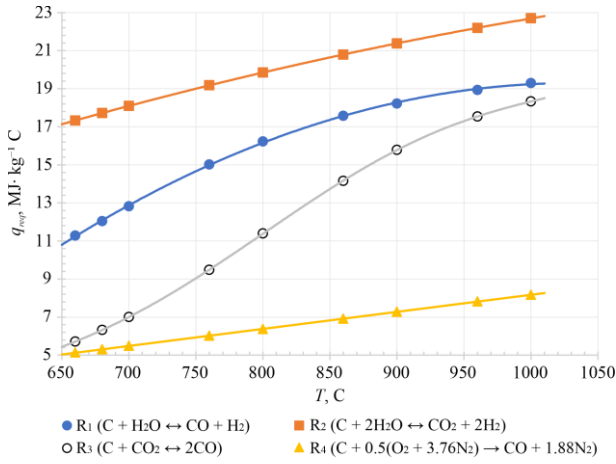


Figure 1. Dependence of the specific heat requirement q_{req} on temperature for routes R1-R4 at $p_{abs} = 0.2533$ MPa; calculation for 1 kg of initial C

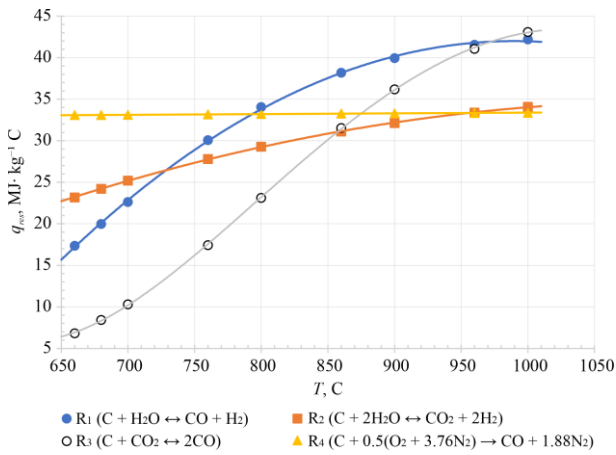


Figure 2. Dependence of the specific heat resource q_{res} on temperature for routes R1-R4 at $p_{abs} = 0.2533$ MPa

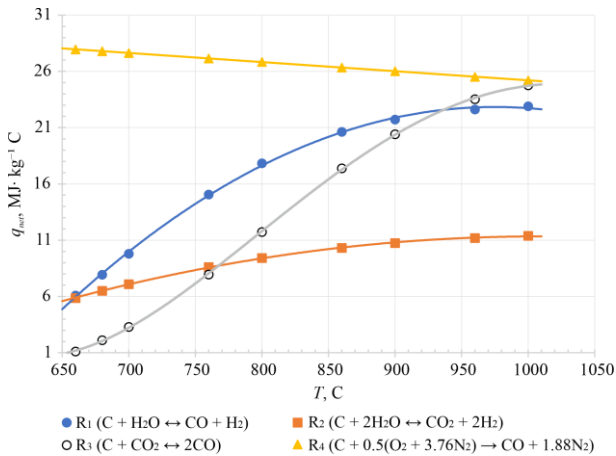


Figure 3. Dependence of the net specific heat q_{net} on temperature for routes R1-R4 at $p_{abs} = 0.2533$ MPa

Therefore, temperatures above 900°C have a more tangible energy motivation for CO₂-conversion than for R₁. In the temperature range of 900-960°C, the R₃ curve passes above R₁ (at 960°C and the value q_{net} is 23.53 and 22.62 MJ·kg⁻¹ C, respectively). For R₂, the additional resource is largely compensated by heat requirement, so q_{net} at 860-1000°C remains low: 10.3-11.4 MJ·kg⁻¹ C. R₄ is characterized by the highest values of q_{net} , but the curve shows a downward trend due to increased energy consumption for heating at almost constant q_{res} .

The q_{res}/q_{req} ratio makes it possible to assess how quickly the increase in energy resource compensates for the increase in heat requirement (Fig. 4). For R₁, K_{TSS} increases from approximately 1.54 at 660°C to 2.17 at 860°C, while remaining close to 2.19 in the 900-1000°C range. Such a plateau means almost identical relative increases in the numerator and denominator, rather than a cessation of q_{net} growth. For R₂, the coefficient is close to 1.50. Additional water supply and lack of CO in the accepted set of products limit the energy advantage of this hydrogen route. For R₃, the value increases from 1.19 at 660°C to 2.23 at 860°C and 2.35 at 1000°C. At base pressure, the K_{TSS} value on the R₃ route exceeds that for the R₁ route at temperatures of 830-1000°C. For R₄, the coefficient decreases from 6.44 to 4.08.

The LHV_g parameter characterizes the energy of the unit of normalized mixture volume and complements q_{res} , referred to the mass of C (Fig. 5).

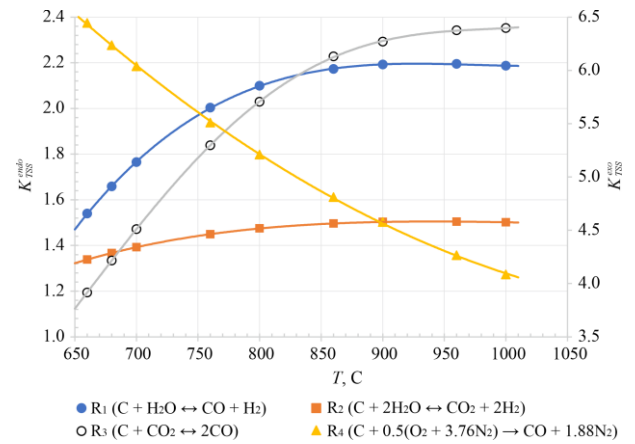


Figure 4. Temperature dependences of the potential thermal self-sufficiency coefficients K_{TSS}^{endo} and K_{TSS}^{exo} at $p_{abs} = 0.2533$ MPa; left axis – R1-R3, right axis – R4

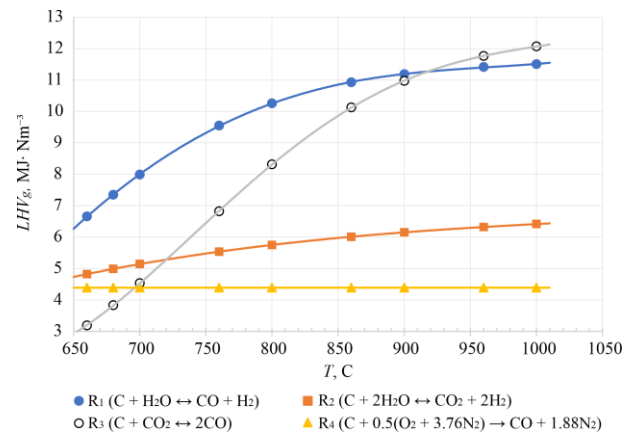


Figure 5. Dependence of the lower heating value of product gas LHV_g on temperature for routes R1-R4 at $p_{abs} = 0.2533$ MPa

In the R₁, the increase in temperature decreases the share of residual steam and increases the share of H₂ + CO; at 860°C, it is 93.24 vol.% (Table 3), and $LHV_g = 10.92$ MJ·Nm⁻³. Above 860-900°C, the increase slows down, approaching the maximum value of a dry equimolar H₂-CO mixture of about 11.72 MJ·Nm⁻³. In the R₂, hydrogen is diluted with CO₂ and residual steam; therefore, at 860-1000°C, LHV_g is only 6.01-6.42 MJ·Nm⁻³, although the share and yield of H₂ is

higher than that of R₁. In the R₃, the CO₂ to CO conversion increases the LHV_g from 3.19 at 660°C to 12.07 MJ·Nm⁻³ at 1000°C. In the range of 920-1000°C, this route is ahead of R₁. In the R₄, the constant ratio of CO:N₂ = 1:1.88 determines constant heating value of about 4.39 MJ·Nm⁻³. Thus, the highest heat resource of the process q_{res} and the highest gas quality in terms of LHV are not identical criteria.

The generalization of dependences given in Figures 1-5 reveals that the high-temperature range is the most informative for comparing the routes; in this range, significant conversion rates have already been reached for endothermic processes, and relatively high values of q_{res} , q_{net} , K_{TSS} and LHV_g are formed. However, the nature of the change in the parameters for individual reactions differs significantly: for R₁ after 860-900°C, the K_{TSS}^{endo} approaches a plateau, for R₂ the temperature effect is relatively weak, while for R₃ the

temperature rise to 1000°C continues to provide a noticeable improvement in energy performance. For R₄, on the contrary, with the increase in temperature, the thermal advantage of the process decreases due to an increase in energy consumption for heating, while the heat resource of the products remains virtually unchanged.

Therefore, for a more visual quantitative comparison of processes, Table 4 provides the main thermal and energy indicators at 860, 960 and 1000°C. The temperature of 860°C is chosen as the lower boundary of the intensive formation of CO- and H₂-containing gas, 960°C – as the characteristic temperature of the high-temperature regime, and 1000°C – as the upper boundary of the examined range, which makes it possible to assess the feasibility of further temperature increase for each route within the adopted comparative calculation framework.

Table 4. Key thermal indicators at 860, 960 and 1000°C

Route/reaction	T, °C	q_{req} , MJ kg ⁻¹ C	q_{res} , MJ kg ⁻¹ C	q_{net} , MJ kg ⁻¹ C	K_{TSS}	LHV_g , MJ Nm ⁻³
R ₁ / C + H ₂ O ↔ CO + H ₂	860	17.58	38.20	20.62	2.17	10.92
	960	18.93	41.55	22.62	2.19	11.41
	1000	19.30	42.20	22.90	2.19	11.51
R ₂ / C + 2H ₂ O ↔ CO ₂ + 2H ₂	860	20.79	31.11	10.32	1.50	6.01
	960	22.19	33.39	11.20	1.50	6.32
	1000	22.70	34.08	11.39	1.50	6.42
R ₃ / C + CO ₂ ↔ 2CO	860	14.16	31.55	17.39	2.23	10.13
	960	17.54	41.07	23.53	2.34	11.76
	1000	18.33	43.10	24.77	2.35	12.07
R ₄ / C + 0.5(O ₂ + 3.76N ₂) → CO + 1.88N ₂	860	6.92	33.26	26.35	4.81	4.39
	960	7.82	33.34	25.52	4.27	4.39
	1000	8.17	33.37	25.20	4.08	4.39

The specific heat requirement of all processes increases with temperature. For R₁, it increases from 11.28 MJ kg⁻¹ C at 660°C to 19.30 MJ kg⁻¹ C at 1000°C. At the same time, q_{res} grows faster to 860-900°C, causing K_{TSS}^{endo} to rise from 1.54 to 2.19, and then reaches a plateau.

For R₂, the K_{TSS}^{endo} coefficient stabilizes at around 1.50. This route yields an increased H₂ fraction, but CO₂ and residual steam reduce the LHV_g to 6.01-6.42 MJ Nm⁻³ in the range of 860-1000°C.

R₃ has the highest temperature sensitivity among endothermic reactions. K_{TSS}^{endo} increases from 1.19 at 660°C to 2.23 at 860°C and 2.35 at 1000°C, and LHV_g – from 3.19 to 12.07 MJ Nm⁻³. Consequently, temperatures above 900°C can be justified for specialized CO₂-conversion schemes, but require assessment of the lining resource, the possibility of slagging, as well as consideration of actual heat losses.

For R₄, q_{res} remains almost constant – 33.08-33.37 MJ kg⁻¹ C, while q_{req} increases due to heating carbon and air. As a result, K_{TSS}^{exo} decreases from 6.44 to 4.08, and q_{net} – from 27.95 to 25.20 MJ kg⁻¹ C. Despite the high heat resource, the product has LHV_g = 4.39 MJ Nm⁻³ due to nitrogen dilution; therefore, R₄ should be considered primarily as a source of internal thermal compensation for endothermic gasification stages, rather than as an independent route for obtaining high-quality fuel gas.

A comparison of temperature dependences showed that among the routes considered, the greatest interest in further analysis of the pressure influence is two independent endothermic reactions of direct carbon conversion – R₁ steam gasification and R₃ CO₂-gasification. They determine two fundamentally different promising technological directions:

obtaining H₂- and CO-containing synthesis gas during carbon-water vapor interaction and obtaining CO-containing gas with the simultaneous involvement of CO₂ as a gasifying agent. Furthermore, for both reactions, the forward direction of the reaction is accompanied by an increase in the number of gas moles. Therefore, a change in absolute pressure directly influences the equilibrium conversion rate, the product gas composition and, accordingly, the q_{res} , q_{net} , K_{TSS} and LHV_g .

The R₂ reaction was not analyzed separately in terms of pressure, as it is the total result of the R₁ steam gasification of carbon and the water-gas shift reaction, and is therefore not considered as an independent route for the direct solid carbon conversion. Incomplete air oxidation R₄ in the adopted model was used primarily as a source of thermal compensation for endothermic stages; therefore, the pressure influence within this comparative analysis was not decisive. In view of this, further study of the pressure influence is focused on R₁ and R₃ routes.

3.3. Influence of pressure on steam and CO₂-conversion of carbon

For R₁, an increase in absolute pressure shifts the equilibrium toward a smaller number of gas moles; and since the direct reaction C + H₂O ↔ CO + H₂ occurs with an increase in the number of gases then, with unchanged initial quantities of reactants, this reduces the equilibrium yield of H₂ and CO and reduces K_{TSS} (Fig. 6). At 860°C, an increase in p_{abs} from 0.1013 to 1.0133 MPa reduces the coefficient from 2.25 to 1.90, that is, by 15.6%, while q_{net} decreases from 22.91 to 13.87 MJ·kg⁻¹ C, or approximately by 39.5%. In the lower and middle parts of the temperature range, the curves diverge

more strongly. At the lowest pressure, the plateau forms earlier, approximately in the range of 860-900°C, while at 0.5066-1.0133 MPa, a noticeable increase persists in higher temperatures. At 960-1000°C, curves converge as a result of the conversion approach to its stoichiometric limit, but the pressure influence does not disappear completely. Thus, the range 860-900°C cannot be considered technologically expedient for all the pressures considered.

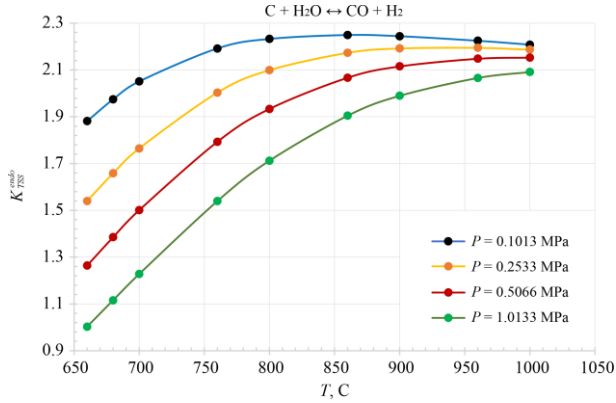


Figure 6. Temperature dependences K_{TSS}^{endo} of R_1 steam gasification at different absolute pressures p_{abs}

For the reaction $C + CO_2 \leftrightarrow 2CO$ (Fig. 7), as for R_1 , the process proceeds with an increase in the number of gas moles; therefore, an increase in pressure reduces the equilibrium CO yield and the potential heat coverage coefficient. The effect is particularly noticeable in the lower temperature range: at the highest studied pressure (1.0133 MPa), the K_{TSS} curve is below unity at 660-700°C, at a pressure of about 0.5066 MPa, the transition through the unit occurs between 660 and 680°C. These intervals characterize the sign change of the calculated q_{net} . At the lowest pressure, K_{TSS} approaches a plateau at about 900°C; at higher pressures, its increase continues up to 1000°C. At the base pressure, the value is 2.23 at 860°C and 2.35 at 1000°C. Therefore, the advantage of temperature rise for R_3 depends on the pressure: the maximum value at the upper boundary of the studied range does not prove the existence of a universal optimum specifically at 1000°C.

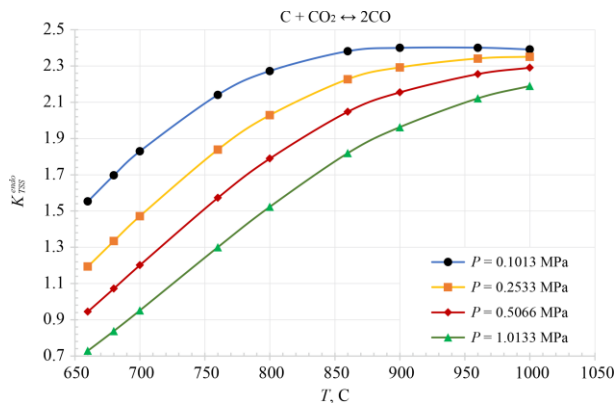


Figure 7. Temperature dependences K_{TSS}^{endo} of R_3 carbon-acid gasification at different absolute pressures p_{abs}

Along with thermal self-sufficiency, it is important to assess the pressure influence on the product gas energy quality. Therefore, the next stage analyzed the dependence of LHV_g on temperature and absolute pressure for R_1 and R_3 (Fig. 8, 9).

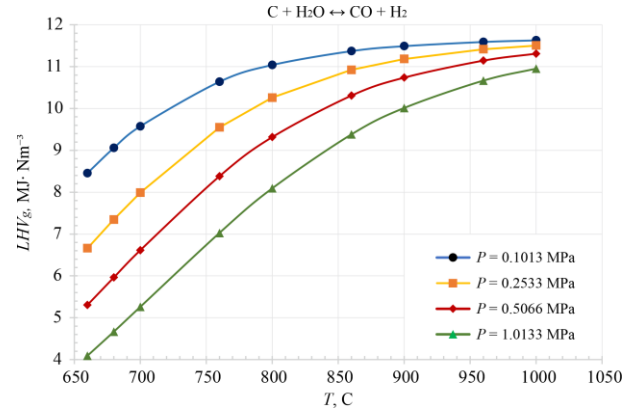


Figure 8. Temperature dependences of the lower heating value of gas LHV_g , steam gasification R_1 at different absolute pressures p_{abs}

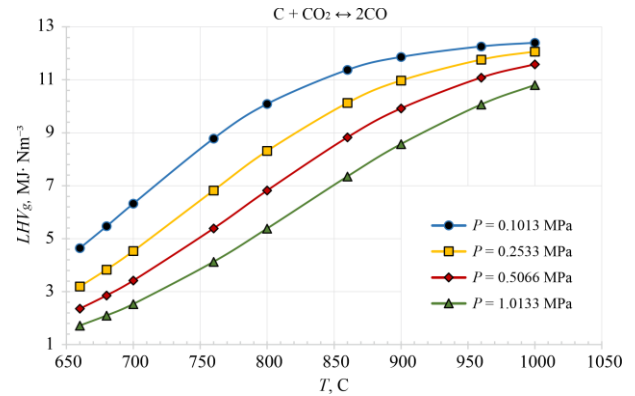


Figure 9. Temperature dependences of the lower heating value of gas LHV_g , carbon dioxide gasification R_3 at different absolute pressures p_{abs}

For both routes, an increase in absolute pressure at a constant temperature reduces the lower heating value of product gas (Fig. 8, 9), but the nature of this effect is different. In the case of R_1 steam conversion, increased pressure reduces the equilibrium carbon conversion rate, resulting in an increased share of unreacted water vapor in the gas mixture. Since, within the accepted single-reaction model, the ratio $H_2/CO = 1$, the divergence of the LHV_g curves is due mainly to the varying degree of dilution of the combustible mixture with steam. With the increase in temperature, this effect weakens. In the lower and middle parts of the studied interval, the curves for different pressures diverge markedly, while at 960-1000°C they converge. At base pressure, LHV_g is 10.92 MJ Nm⁻³ at 860°C and 11.51 MJ Nm⁻³ at 1000°C. At the same time, following the complete extraction of all water vapor, the lower heating value of the dry gas R_1 in this model would be about 11.72 MJ Nm⁻³ and would be practically independent of temperature and pressure. Therefore, Figure 8 reflects the simultaneous effects of equilibrium conversion and gas humidity and does not indicate a similar pressure influence on the dried product LHV_g .

For CO_2 -conversion (Fig. 9), the R_3 mechanism is different: an increase in LHV_g with temperature is caused by an increase in the CO_2 to CO conversion rate, while the increase in pressure counteracts this reaction and contributes to the preservation of a larger share of unreacted CO_2 in the product gas. Therefore, unlike R_1 , the reduction in heating value cannot be eliminated by simple drying, since CO_2 remains as a gaseous ballast component. At base pressure, LHV_g for R_3

increases from 10.13 MJ Nm^{-3} at 860°C to 11.76 at 960°C and 12.07 MJ Nm^{-3} at 1000°C , approaching the lower heating value of pure CO , which is 12.64 MJ Nm^{-3} . At lower pressure, this approximation occurs at lower temperatures, while at high pressure, a significant increase in LHV_g is retained at the top of the temperature range.

Thus, the increase in temperature partially compensates for the adverse equilibrium pressure influence for both routes, but the nature of the product gas dilution is fundamentally different: for R_1 , the residual H_2O content is the determining factor, and for R_3 – the unreacted CO_2 . Therefore, when choosing a steam gasification mode, it is advisable to take into account separately the possibility of gas condensation and drying, while for CO_2 -conversion, it is necessary to compare additional energy expenditures for increasing the temperature against the expenditure of CO_2 extraction or recirculation, as well as the pressure requirements at the subsequent process stage.

At the same time, the pressure influence cannot be assessed unequivocally only by equilibrium calculation. The actual gasification rate depends on the structure of carbon-containing waste, ash composition, porosity, mass transfer, and reactor hydrodynamics [36]–[38]. Increased pressure can reduce equipment dimensions, intensify phase contact and simplify gas transportation. Experimental data for carbon residues of other origin indicate that, under certain conditions, pressure can increase the conversion rate and heating value of the gas [37]. Therefore, for coal mining waste, the theoretical conclusion about the advantage of low or moderate absolute pressure requires experimental verification.

3.4. Justification of the optimal temperature range

At base pressure, the $860\text{--}900^\circ\text{C}$ range is a suitable initial range for the R_1 experimental verification. At 860°C , the model gives $X_C = 87.34\%$, the total share of $\text{H}_2 + \text{CO} = 93.24 \text{ vol.}\%$, while $LHV_g = 10.92 \text{ MJ Nm}^{-3}$ and $K_{TSS}^{endo} = 2.17$. Further temperature increase continues to increase gas and q_{net} yields, but K_{TSS} changes little. At the same time, the requirements for thermal insulation, steam superheating, heat resistance of feeding units, lining resource and slagging control, etc., are increasing. Therefore, the $860\text{--}900^\circ\text{C}$ range should be considered as a basic technologically rational range, which is a compromise between conversion and heat load, but does not prove an economic optimum or sufficient reaction rate for specific waste types.

For gas with a high H_2 content and yield, the R_2 route is preferred as a model for total steam conversion. At 860°C , the H_2 yield from Table 3 is about $2.88 \text{ Nm}^3 \text{ kg}^{-1} \text{ C}$ for R_2 , compared with $1.63 \text{ Nm}^3 \text{ kg}^{-1} \text{ C}$ for R_1 ; the share of H_2 in the wet gas is 55.72 compared with $46.62 \text{ vol.}\%$. At the same time, this advantage is accompanied by a greater heat requirement and the formation of CO_2 . Technologically, this goal should be implemented as steam gasification with a separate controlled degree of water-gas shift and CO_2 extraction. An increase in gasification temperature does not mean that the shift should be carried out at the same temperature: exothermic shift reaction $\text{CO} + \text{H}_2\text{O} \leftrightarrow \text{CO}_2 + \text{H}_2$ is thermodynamically more favorable at a lower temperature, and the catalyst accelerates the approximation to the equilibrium state.

The R_3 is preferred to obtain CO -containing gas and to use CO_2 as a reactant, especially in the available high-temperature heat and work in the $900\text{--}1000^\circ\text{C}$ range. At 1000°C , the share

of CO reaches $95.47 \text{ vol.}\%$, $LHV_g = 12.07 \text{ MJ Nm}^{-3}$, and $q_{net} = 24.77 \text{ MJ kg}^{-1} \text{ C}$. At the same time, the CO_2 to CO conversion is not its long-term binding. During the subsequent combustion of CO , CO_2 is formed. The environmental climatic effect will be determined by the source of heat, origin of C and final use of gas, etc.

From the position of the highest value of the potential heat coverage coefficient and internal exothermic contribution, R_4 is preferable. However, it forms gas with $65.28 \text{ vol.}\%$ N_2 and $LHV_g \approx 4.39 \text{ MJ Nm}^{-3}$. Such gas may be suitable for local combustion in appropriate equipment, but its low volumetric energy density increases volumetric gas flows and necessitates specialized equipment and combustion modes.

For high-quality fuel gas, the choice depends on the basis of comparison. At 860°C , with the given moisture content, R_1 has a higher LHV_g , than R_3 , and at $960\text{--}1000^\circ\text{C}$, R_3 is ahead of R_1 . After drying, the R_1 model gas has an LHV_g of about 11.72 MJ Nm^{-3} ; therefore, comparisons based on moisture content cannot be directly applied to consumers of dry gas. In addition to LHV_g , it is necessary to assess the H_2/CO ratio, compositional stability, dust, tars, sulfur-containing impurities, and the requirements of the burner, engine, fuel cell, synthesis process, etc.

For autothermal operation, a combined scheme of endothermic gasification and controlled oxidation is appropriate [39], [40]. The share of the oxidizing agent is determined from the closed heat balance at a minimum dilution and a permissible decrease in the yield of the target gas. Given strict limitations on N_2 , oxygen enrichment or indirect heat supply from a separate combustion zone should be considered, taking into account additional expenditures. Experimental and theoretical studies of two-stage gasifiers confirm the determining role of the gasifying agent composition, steam supply and available heat resource [41], [42].

Catalysis and heat recovery can expand the area of virtually suitable modes. Catalytic intensification of the water-gas and water-gas shift reactions can reduce the required temperature or reduce the time needed to reach a gas composition close to equilibrium [43]. This is particularly important for coal mining waste, as the actual rate of the process depends on the porosity, dispersion, mineral impurities and reactivity of the carbon residue. A change in the equilibrium yield is possible, in particular, when the product is removed or there is some other change in the composition of the system. To move to the waste of a particular field, it is necessary to determine the content of C, moisture content, ash content, and then check the selected modes experimentally. Heat recovery from the product gas can additionally reduce external heat demand by preheating the feedstock and gasifying agents.

The prospects for further product gas conversion are determined by its target use: obtaining H_2 , energy use or use as raw materials for synthesis; for hydrogen direction, the integration of gasification with the water-gas shift reaction and subsequent CO_2 extraction are of particular interest [44]–[46].

The obtained values are comparative theoretical estimates for the accepted idealized routes, and the identified temperature ranges are guidelines for planning experiments. The transition to specific carbon-containing waste requires taking into account their carbon content, ash content, moisture content, mineral composition, reactivity ability, kinetic and diffusion limitations, real heat loss and heat recovery efficiency.

4. Conclusions

A unified system of specific indicators, q_{req} , q_{res} , q_{net} and K_{TSS} , normalized to 1 kg of initial carbon is proposed, enabling endothermic and exothermic carbon conversion routes to be compared on a common calculation basis. These indicators characterize the potential heat coverage of the process and are not direct evidence of the autothermicity of the real plant.

For steam gasification (R_1 route) at $p_{abs} = 0.2533$ MPa, the temperature range of 860-900°C is a rational initial interval for experimental verification. At 860°C, the carbon conversion rate is 87.34%, the total share is $H_2 + CO - 93.24$ vol.%, $LHV_g = 10.92$ MJ Nm⁻³, and $K_{TSS}^{endo} = 2.17$. Further increase in temperature increases the gas and q_{net} yields, however, K_{TSS}^{endo} practically reaches a plateau, which reduces the energy motivation for further heating specifically for R_1 .

The R_2 route provides the largest yield and share of H_2 among the examined models but is characterized by the greatest heat requirement and lower LHV_g due to the presence of CO_2 and residual H_2O . It is therefore more appropriate to implement hydrogen route as steam gasification with a separate, controlled stage of water-gas shift and, if necessary, CO_2 extraction, rather than as a single high-temperature equilibrium process.

CO_2 – gasification R_3 has the highest temperature sensitivity among endothermic routes. At base pressure, the temperature rise from 860 to 1000°C increases K_{TSS}^{endo} from 2.23 to 2.35, q_{net} – from 17.39 to 24.77 MJ kg⁻¹ C, and LHV_g – from 10.13 to 12.07 MJ Nm⁻³. Therefore, R_3 is the preferred route for the targeted production of CO-containing gas and the use of CO_2 as a gasifying agent, especially when a high-temperature heat source is available.

Incomplete air oxidation R_4 provides the largest internal exothermic contribution and the highest q_{net} values, however, after 65.28 vol.% N_2 forms gas with $LHV_g = 4.39$ MJ Nm⁻³. Therefore, its main function in the combined scheme is to consider thermal compensation of endothermic stages, and to determine the oxidizing agent share from the closed heat balance at the minimum permissible dilution of the target gas with nitrogen.

The increase in absolute pressure from 0.1013 to 1.0133 MPa adversely affects the equilibrium characteristics of both steam R_1 and CO_2 -gasification R_3 , since the direct course of both reactions is accompanied by an increase in the number of gas moles. The negative influence of pressure is most pronounced in the lower and middle parts of the temperature range and weakens with increasing temperature.

The reasons for decrease in LHV_g with increasing pressure differ fundamentally for R_1 and R_3 . For R_1 , it is mainly associated with an increased share of unreacted water vapor; consequently, after drying the gas, this effect is largely eliminated; for R_3 , the inert component is unreacted CO_2 , which remains in dry gas. Therefore, the choice of pressure should take into account not only the equilibrium conversion, but also the scheme of further conditioning and use of product gas.

The preferred gasification route is determined by the target product: R_1 – to obtain a balanced H_2/CO -synthesis gas; R_2 or technologically $R_1 +$ water-gas shift reaction – for increased H_2 yield; R_3 – for CO-containing gas and involvement of CO_2 as a reactant; R_4 – for internal thermal compensation and formation of autothermal regime. Thus, there is no universally preferred route, and the choice should be based on a combination of the target gas composition, LHV_g , heat resource and the requirements of the next technological stage.

Author contributions

Conceptualization: AH, PS; Data curation: VL, MP, KS; Formal analysis: AH, NL; Funding acquisition: AH, PS, VL, MP, KS; Investigation: AH, PS; Methodology: AH, NL; Project administration: PS; Resources: AH, NL, VL, MP, KS; Supervision: PS; Validation: AH, PB; Visualization: AH, VL; Writing – original draft: AH, NL, PS, VL; Writing – review & editing: AH, VL, MP, KS. All authors have read and agreed to the published version of the manuscript.

Funding



This study is supported by the National Research Foundation of Ukraine (Project No. 2025.07/0393, “Development of scientific basis of physical-chemical and thermodynamic processes of coal-mining waste gasification for resource provision and sustainable development” for the call “Excellent Science in Ukraine 2026-2028”).

Acknowledgements

The authors express their sincere gratitude to the editor and the reviewers for their valuable comments and recommendations, which helped improve the manuscript. They also express their sincere gratitude to the Armed Forces of Ukraine for their courage, resilience, and continued defence of Ukraine, which enables the authors to live, work, conduct research, and contribute to the development of science.

Conflicts of interest

Authors PS, VL, MP, and KS declare that they served on the Editorial Board of *Mining of Mineral Deposits* at the time of manuscript submission. None of them was involved in the handling, peer review, or decision-making process for this manuscript. The manuscript was handled independently by an editor with no conflict of interest. This had no impact on the peer review process or the final decision. The remaining authors declare that they have no conflicts of interest.

Data availability statement

The original contributions presented in the study are included in the article, further inquiries can be directed to the corresponding author.

References

- [1] International Energy Agency. (2024). *Coal 2024: Analysis and forecast to 2027*. Retrived from: <https://www.iea.org/reports/coal-2024>
- [2] Hudson-Edwards, K.A., & Dold, B. (2015). Mine waste characterization, management and remediation. *Minerals*, 5(1), 82-85. <https://doi.org/10.3390/min5010082>
- [3] Popovych, V., Voloshchysyn, A., Lysyi, N., Petlovanyi, M., & Shuplat, T. (2024). Natural phytomelioration processes on rock dumps of abandoned coal mines (Ukraine). *Rudarsko-Geološko-Naftni Zbornik*, 39(5), 63-73. <https://doi.org/10.17794/rgn.2024.5.4>
- [4] Gao, L., Liu, Y., Xu, K., Bai, L., Guo, N., & Li, S. (2024). A short review of the sustainable utilization of coal gangue in environmental applications. *RSC Advances*, 14, 39285-39296. <https://doi.org/10.1039/D4RA06071G>
- [5] Saik, P., Petlovanyi, M., Lozynskiy, V., Sai, K., & Merzlikin, A. (2018). Innovative approach to the integrated use of energy resources of underground coal gasification. *Solid State Phenomena*, 277, 221-231. <https://doi.org/10.4028/www.scientific.net/SSP.277.221>
- [6] Dychkovskiy, R., Falshtynskiy, V., Saik, P., & Lozynskiy, V. (2025). Aspects of co-utilization of solid waste and coal through underground gasification. *IOP Conference Series: Earth and Environmental Science*, 1457(1), 012002. <https://doi.org/10.1088/1755-1315/1457/1/012002>

- [7] Petlovanyi, M.V., Lozynskiy, V.H., Saik, P.B., & Sai, K.S. (2026). Research into the granulometric, chemical and mineralogical characteristics of coal mining waste as incoming raw materials to the gasification process. *Mineral Resources of Ukraine*, 3, 41-46. <https://doi.org/10.31996/mru.2026.3.41-46>
- [8] Yang, L., Zhang, X., Yan, S., & Luo, Y. (2024). Research on the resource treatment and comprehensive utilization of carbon containing wastes using pyrolysis-gasification two-stage recycling. *Processes*, 12(2), 361. <https://doi.org/10.3390/pr12020361>
- [9] Haydary, J. (2018). Aspen simulation of two-stage pyrolysis/gasification of carbon based solid waste. *Chemical Engineering Transactions*, 70, 1033-1038. <https://doi.org/10.3303/CET1870173>
- [10] Pivnyak, G., Dychkovskiy, R., Bobyliov, O., Cabana, E.C., & Smoliński, A. (2018). Mathematical and geomechanical model in physical and chemical processes of underground coal gasification. *Solid State Phenomena*, 277, 1-16. <https://doi.org/10.4028/www.scientific.net/ssp.277.1>
- [11] Falshtynskiy, V.S., Dychkovskiy, R.O., Lozynskiy, V.G., & Saik, P.B. (2013). Determination of the technological parameters of borehole underground coal gasification for thin coal seams. *Journal of Sustainable Mining*, 12(3), 8-16. <https://doi.org/10.7424/jsm130302>
- [12] Falshtynskiy, V.S., Dychkovskiy, R.O., Saik, P.B., Lozynskiy, V.H., & Cabana, E.C. (2017). Formation of thermal fields by the energy-chemical complex of coal gasification. *Naukovy Visnyk Natsionalnoho Hirnychoho Universytetu*, 5, 36-42.
- [13] Buktukov, N.S., Gumennikov, E.S., Mashataeva, G.A. (2019). In-situ gasification of steeply dipping coal beds with production hole making by supersonic hydraulic jets. *Mining Informational and Analytical Bulletin*, 9, 30-40. <https://doi.org/10.25018/0236-1493-2019-09-0-30-40>
- [14] Lozynskiy, V., Falshtynskiy, V., Saik, P., & Lozynska, M. (2025). Substantiation of the parameters for underground gasifier combustion face advance considering the influence of magnetic fields on the gasification process intensification. *IOP Conference Series: Earth and Environmental Science*, 1491(1), 012041. <https://doi.org/10.1088/1755-1315/1491/1/012041>
- [15] Lysyi, N., Helesh, A., Popovych, V., Saik, P., & Dmytruk, O. (2025). Thermodynamic research of coal mining waste gasification processes. *Mining of Mineral Deposits*, 19(3), 132-143. <https://doi.org/10.33271/mining19.03.132>
- [16] Helesh, A., Lysyi, N., & Popovych, V. (2025). Doslidzhennia vplyvu tysku na protses oderzhannia syntez-hazu konversiei vuhletsu vodiannoi paroiu. *Khimichna Tekhnologhiia ta Inzheneriia*, 65-70.
- [17] Saik, P., Lysyi, N., Dmytruk, O., & Lozynskiy, V. (2026). Thermochemical utilization of coal mining waste: challenges and technological prospects. *Scientific Journal of Metinvest Polytechnic. Series: Technical Sciences*, 6, 211-221. <https://doi.org/10.32782/3041-2080/2026-6-26>
- [18] Saik, P., Dmytruk, O., & Lysyi, N. (2025). On the issue of utilization of carbon-containing mining waste. *Collection of Research Papers of the NMU*, 82, 296-308. <https://doi.org/10.33271/crpnmu/82.296>
- [19] Lysyi, N.R., Helesh, A.B., & Popovych, V.V. (2024). Gasification of coal-containing waste. *Proceedings of the International Scientific Conference: Actual Problems of Natural Sciences Development Amidst the Evolution of Artificial Intelligence*, 30-34. <https://doi.org/10.30525/978-9934-26-521-1-7>
- [20] Dai, F., Zhang, S., Luo, Y., Wang, K., Liu, Y., & Ji, X. (2023). Recent progress on hydrogen-rich syngas production from coal gasification. *Processes*, 11(6), 1765. <https://doi.org/10.3390/pr11061765>
- [21] Segneri, V., Ferrasse, J.H., Trinca, A., & Vilardi, G. (2022). An overview of waste gasification and syngas upgrading processes. *Energies*, 15(17), 6391. <https://doi.org/10.3390/en15176391>
- [22] Saik, P., Dychkovskiy, R., Lozynskiy, V., Falshtynskiy, V., Cabana, E.C., & Hrytsenko, L. (2021). Chemistry of the gasification of carbonaceous raw material. *Materials Science Forum*, 1045, 67-78. <https://doi.org/10.4028/www.scientific.net/msf.1045.67>
- [23] Porcu, A., Sollai, S., Marotto, D., Mureddu, M., Ferrara, F., & Pettinau, A. (2019). Techno-economic analysis of a small-scale biomass-to-energy BFB gasification-based system. *Energies*, 12(3), 494. <https://doi.org/10.3390/en12030494>
- [24] Dong, X., Wang, J., Yang, L., Liu, F., & Zuo, Z. (2025). Mechanistic insights into copper slag as a thermal carrier and catalyst in lignite steam gasification: Experimental and DFT study. *Energy*, 335, 138310. <https://doi.org/10.1016/j.energy.2025.138310>
- [25] Gao, J., Su, W., Hou, J., Song, X., Wang, J., Lv, P., Xu, G., & Yu, G. (2024). High-temperature corrosion behavior of high chromia refractory by alkali metal sodium with different presence states under gasification conditions. *Energy*, 309, 133093. <https://doi.org/10.1016/j.energy.2024.133093>
- [26] Han, F., Niu, Y., Zhang, X., Guo, Z., Duan, S., Liu, H., Lu, B., & Chen, H. (2024). Performance analysis of a pilot gasification system of biomass with stepwise intake of air-steam considering waste heat utilization. *Renewable Energy*, 236, 121498. <https://doi.org/10.1016/j.renene.2024.121498>
- [27] Cvetinović, D., Erić, A., Milutinović, N., Petrov, N., Anđelković, J., & Bakić, V. (2024). Thermodynamic equilibrium modeling of biomass gasification: Effects of operating conditions on gasifier performance. *Journal of King Saud University – Science*, 36(10), 103370. <https://doi.org/10.1016/j.jksus.2024.103370>
- [28] Smith, J.M., Van Ness, H.C., Abbott, M.M., & Swihart, M.T. (2018). *Introduction to chemical engineering thermodynamics*. New York, United States: McGraw-Hill Education, 769 p.
- [29] Linstrom, P.J., & Mallard, W.G. (2026). *NIST Chemistry WebBook (NIST Standard Reference Database Number 69)*. National Institute of Standards and Technology. <https://doi.org/10.18434/T4D303>
- [30] Wagner, W., & Kretzschmar, H.-J. (2008). *International steam tables: Properties of water and steam based on the industrial formulation IAPWS-IF97*. Heidelberg, Germany: Springer, 391 p. <https://doi.org/10.1007/978-3-540-74234-0>
- [31] Lozynskiy, V. (2023). Critical review of methods for intensifying the gas generation process in the reaction channel during underground coal gasification (UCG). *Mining of Mineral Deposits*, 17(3), 67-85. <https://doi.org/10.33271/mining17.03.067>
- [32] Dychkovskiy, R., Falshtynskiy, V., Saik, P., Lozynskiy, V., Sala, D., Hankus, Ł., Magdziarczyk, M., & Smoliński, A. (2025). Control of contour evolution, burn rate variation, and reaction channel formation in coal gasification. *Scientific Reports*, 15(1), 9075. <https://doi.org/10.1038/s41598-025-93611-3>
- [33] Lozynskiy, V. (2024). Numerical simulation of carbonaceous raw material combustion in a coal seam channel. *Mining of Mineral Deposits*, 18(4), 109-124. <https://doi.org/10.33271/mining18.04.109>
- [34] Lozynskiy, V., Saik, P., Petlovanyi, M., Sai, K., Malanchuk Z., & Malanchuk, Ye. (2018). Substantiation into mass and heat balance for underground coal gasification in faulting zones. *Inżynieria Mineralna*, 19(2), 289-300. <http://doi.org/10.29227/IM-2018-02-36>
- [35] Saik, P., & Berdnyk, M. (2022). Mathematical model and methods for solving heat-transfer problem during underground coal gasification. *Mining of Mineral Deposits*, 16(2), 87-94. <https://doi.org/10.33271/mining16.02.087>
- [36] Gil, S., Bialik, W., Moeck, P., Rimár, M., Kizek, J., & Polivka, N. (2023). Investigation of the kinetics of pressure coal char hydrogasification. *Energies*, 16(13), 4937. <https://doi.org/10.3390/en16134937>
- [37] López, F.A., Centeno, T.A., Alguacil, F.J., Lobato, B., López-Delgado, A., & Feroso, J. (2012). Gasification of the char derived from distillation of granulated scrap tyres. *Waste Management*, 32(4), 743-752. <https://doi.org/10.1016/j.wasman.2011.08.006>
- [38] Bazaluk, O., Lozynskiy, V., Falshtynskiy, V., Saik, P., Dychkovskiy, R., & Cabana, E. (2021). Experimental Studies of the effect of design and technological solutions on the intensification of an underground coal gasification process. *Energies*, 14(14), 4369. <https://doi.org/10.3390/en14144369>
- [39] Li, Y., Xing, G., Liu, S., Guo, S., Ge, Z., & Guo, L. (2025). Thermodynamic and environmental analysis of the hydrogen and combustible gas production in supercritical water partial oxidation system of coal. *Energy*, 326, 136399. <https://doi.org/10.1016/j.energy.2025.136399>
- [40] Yuan, P., Guo, T., Pan, X., Hu, X., Ma, J., Xu, D., Zhou, Z., & Guo, Q. (2023). Process optimization and thermodynamic analysis of autothermal coal chemical looping gasification industrial demonstration system. *Fuel*, 334, 126667. <https://doi.org/10.1016/j.fuel.2022.126667>
- [41] Li, G., Wang, L., Wang, C., Wang, C., Wu, P., & Che, D. (2020). Experimental study on coal gasification in a full-scale two-stage entrained-flow gasifier. *Energies*, 13(18), 4937. <https://doi.org/10.3390/en13184937>
- [42] Duan, W., Li, R., Yang, S., Han, J., Lv, X., Wang, Z., & Yu, Q. (2024). Theoretical study on coal gasification behavior in CO₂ atmosphere driven by slag waste heat. *Energy*, 305, 132269. <https://doi.org/10.1016/j.energy.2024.132269>
- [43] Baraj, E., Ciałhotný, K., & Hlinčík, T. (2022). Advanced catalysts for the water gas shift reaction. *Crystals*, 12(4), 509. <https://doi.org/10.3390/cryst12040509>
- [44] International Energy Agency. (2024). *Global hydrogen review 2024*. Retrieved from: <https://www.iea.org/reports/global-hydrogen-review-2024>
- [45] Tahir, M.H., & Chen, D. (2024). Integrated approach for H₂-rich syngas production from wastes using carbon-based catalysts and subsequent CO₂ adsorption by carbon-based adsorbents: A review. *International Journal of Hydrogen Energy*, 59, 679-696. <https://doi.org/10.1016/j.ijhydene.2024.02.045>
- [46] Monteiro, E., Bourguig, O., & Rouboa, A. (2024). Hydrogen-rich syngas production based on a co-gasification process coupled to a water-gas shift reactor without steam injection. *Hydrogen*, 5(4), 919-928. <https://doi.org/10.3390/hydrogen5040048>

Енергетичне обґрунтування режимів газифікації вуглецю у вуглецевмісних відходах

А. Гелеш, П. Саїк, Н. Лисий, В. Лозинський, М. Петльованій, К. Сай

Мета. Теоретично обґрунтувати енергетично доцільні режими газифікації вуглецевої складової вуглецевмісних відходів за уніфікованими показниками теплової самозабезпеченості, використовуючи відходи вуглевидобутку як базовий приклад застосування моделі.

Методика. Порівняльні матеріальні та теплові розрахунки чотирьох ідеалізованих маршрутів конверсії вуглецю в інтервалі 660-1000°C виконувалися на основі раніше визначених констант рівноваги. Усі питомі показники віднесено до 1 кг вихідного С. Базове порівняння проводилося за абсолютного тиску близько 0.2533 МПа. Вплив тиску досліджувався для парової та СО₂-газифікації в межах 0.1013-1.0133 МПа. Зіставлення потреби теплоти q_{req} , ресурсу q_{res} , різниці $q_{net} = q_{res} - q_{req}$, коефіцієнта потенційного теплового покриття K_{TSS} та нижчої теплоти згоряння газу виконувалося з урахуванням залишкової пари.

Результати. Враховуючи, що $p_{abs} = 0.2533$ МПа для $C + H_2O \leftrightarrow CO + H_2$, коефіцієнт потенційної теплової самозабезпеченості ендотермічних процесів досягає 2.17 при температурі 860°C, при 960°C становить 2.19 і дещо знижується до 2.187 при 1000°C. Для $C + CO_2 \leftrightarrow 2CO$ він зростає до 2.35 при 1000°C. Неповне окиснення повітрям забезпечує найбільшу q_{net} , однак нижча теплота згоряння продуктового газу (LHV_g) залишається стабільно низькою в усьому досліджуваному температурному діапазоні 4.39 МДж/Нм³ через розбавлення горючих газів азотом повітря (N₂). Підвищення p_{abs} від 0.1013 до 1.0133 МПа погіршує рівноважні показники парової та вуглекислотної конверсій.

Наукова новизна. На спільній базі 1 кг вихідного С кількісно зіставлено температурні залежності енергетичного потенціалу та потреби теплоти ідеалізованих маршрутів газифікації. Запропоновано єдину систему $q_{req} - q_{res} - q_{net} - K_{TSS}$ для коректного порівняння екзо- та ендотермічних маршрутів газифікації вуглецю. Показано, що перевага маршруту залежить від обраного критерію: виходу Н₂, утворення СО, об'ємної теплоти згоряння або потенційної теплової самозабезпеченості. Уточнено зв'язок між впливом тиску, рівноважною конверсією та розбавленням продуктового газу.

Практична значимість. Запропоновано критерії попереднього вибору маршруту відповідно до призначення газу та доступного теплового ресурсу. Для парової газифікації при базовому тиску рекомендовано початкову область випробувань 860-900°C; для цільового одержання СО за наявності високотемпературної теплоти – СО₂-конверсію в діапазоні 900-1000°C. Для водневого продукту доцільно поєднувати парову газифікацію з водяним газовим зсувом, а для автотермічної схеми – визначати частку введення окисника в систему.

Ключові слова: відходи вуглевидобутку; газифікація; ко-газифікація; конверсія вуглецю; синтез-газ; питома потреба теплоти; тепловий ресурс; температура; абсолютний тиск

Publisher's note

All claims expressed in this manuscript are solely those of the authors and do not necessarily represent those of their affiliated organizations, or those of the publisher, the editors and the reviewers.