Kompleksnoe Ispolzovanie Mineralnogo Syra = Complex Use of Mineral Resources

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Metallurgy

To the question of pyrometallurgical technology for processing antimony-gold-bearing ores and concentrates

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ABSTRACT

The increasing demand for non-ferrous, precious, and rare metals necessitates more comprehensive and efficient use of mineral raw materials, such as gold-antimony ores and concentrates. A promising approach is the use of pyrometallurgical processing in a fluidized bed, which offers more efficient heat and mass transfer than conventional technologies. This study aims to investigate the evaporation kinetics of antimony sulfide (Sb_2S_3) from gold-antimony ores and concentrates in a fluidized bed under various conditions. The experiments involved varying temperature (923-1223 K), particle size (0.09-2.0 mm), and layer thickness (5-15 mm) to determine the evaporation rate of Sb_2S_3. The experimental setup consisted of a laboratory-scale fluidized bed reactor equipped with a controlled gas flow of nitrogen mixed with sulfur vapor. The evaporation rates were measured using a gravimetric method and confirmed by X-ray diffraction and microscopic analysis of samples. The results show that the evaporation rate of Sb_2S_3 in a fluidized bed is 7-9 times higher than in a fixed bed. This is due to significantly improved heat and mass transfer in the fluidized system. At 1023 K, the overall evaporation rate increased with decreasing grain size. This is associated with an increase in the total surface area of the material, but the specific evaporation rate normalized to unit surface area was independent of particle size. The process was not significantly affected by bed height in the range of 5-15 mm. Antimony recovery into sublimates improved by 2-3% compared to conventional technology. It reached 98-99% due to suppression of Sb_2O_5 formation. These findings confirm the efficiency of supplying an inert gas with sulfur vapors into the fluidized bed. This reduces harmful gas emissions and minimizes dust entrainment. It also allows for effective distillation of volatile components at lower temperatures.

Keywords: Sulphiding roasting, oxidising roasting, neutral gas, condensation.

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Introduction

Antimony concentrates are supplied to the metallurgical processing stage from beneficiation plants already in a finely ground form, and, if necessary, are subjected to only drying. Antimony ores are supplied to plants, as a rule, in a wet lump form, so it is necessary to subject them to the processes of drying, crushing and grinding. Drying is carried out in tubular rotary kilns with gases from the combustion of fuel oil. After drying, the raw material is crushed in jaw and roller crushers and ground in ball mills. Antimony is present in gold-antimony ores and concentrates in the form of the mineral antimonite, Sb_2S_3 and in small quantities in the form of the following minerals: tetrahedrite

Cu_3SbS_3, valentinite Sb_2O_3, and servantite Sb_2O_4 (over 93%). According to the classical technology of processing gold-antimony polymetallic raw materials in fluidized bed furnaces [[1], [2]], the liquefaction of the charge is carried out with ordinary air, which leads to the formation of volatile antimony compounds and partially non-volatile antimony pentoxide Sb_2O_5, which remains in the cinder.

The authors of this work propose a technology for liquefying the charge with a mixture of neutral gas (N_2) with elemental sulfur vapor, which allows increasing the extraction of antimony into marketable sublimates by 2-3%, due to the exclusion of the formation of non-volatile Sb_2O_5, while the degree of antimony extraction was 98-99%, whereas using standard technology it did not exceed 95-96%.

Research was conducted at various rates and concentrations of the gas mixture supplied to the layer of the processed material to determine the optimal flow rate of the mixture of neutral gas with elemental sulfur vapor.

The materials of the following composition were studied in the work, %: gold - antimony concentrate containing: Sb-58.8; S-21.6; SiO_2-18.1; As-0.27; Pb-0; Fe-0.12; Al_2O_3 -0; MgO-0; C-0; Au-38.0 g/t Ag-0.0 g/t and gold – antimony ore containing: Sb-22.4; S-8.3; SiO_2-53.0; As-0.6; Pb-0.05; Fe-2.2; Al_2O_3 -5.3; MgO-0.3; C-4.5; Au-21.0 g/t Ag-0.001 g/t.

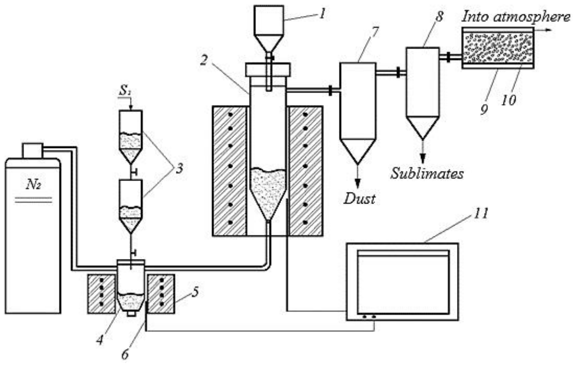
Experimental part

Laboratory installation: design and operating principle

Studies on the processing of polymetallic raw materials in a fluidized bed were carried out using a laboratory installation schematically shown in Figure 1. To avoid disproportionate consumption of elemental sulfur, the system was equipped with an evaporator, which supplied sulfur vapors into a separate reaction vessel containing the charge throughout the experiment. This solution enabled precise control of sulfur consumption and ensured a uniform supply of sulfur vapors mixed with nitrogen into the bed, providing stable fluidization and effective sulfidation of the material.

The experiment started when the charge reached the specified temperature. At the end of the run, the retort was removed from the furnace, cooled, and the obtained products were weighed and analyzed.

The procedure was as follows. The feed material was loaded from hopper (1) into the reaction vessel (2). The system was sealed and heated to the target temperature. Sulfur powder was charged into the evaporator (4) through hoppers (3). After the furnaces (5) were heated, the gas flow was adjusted with a rotameter to achieve stable fluidization in the vessel. The electric furnace (5) was then lowered so that the material bed was positioned within the isothermal zone.



1 – feed hopper; 2 – evaporative reaction vessel;
3 – sulfur powder feed hoppers;
4 – sulfur evaporator; 5 – electric furnaces for the initial charge and sulfur evaporators; 6 – thermocouples;
7 – cyclone; 8 – condenser; 9 – filter; 10 – filter packing;
11 – electric potentiometer

Figure 1 - Schematic diagram of the laboratory fluidized-bed installation

The temperatures in both the sulfur evaporator and the charge were measured with platinum–platinum-rhodium thermocouples and recorded using a potentiometer. Dust and sublimates generated during the process were collected in a cyclone (7) and a condenser (8), with the remaining volatiles captured in a filter (9).

The moment when the charge reached the target temperature was defined as the start of the experiment. After completion, the furnace was raised, the charge cooled, and the system disassembled.

Research methods

Features of the transition of a layer of granular material into a fluidized state. In order to create a stable boiling bed regime when blowing gas through a granular material layer placed in a vessel with a porous bottom, the layer can be in two qualitatively different stationary states.

At flow rates W below a certain critical value W_0, solid particles are motionless, the "penetrating" of the layer ε (the volume fraction of free space between particles) remains unchanged, and its hydraulic resistance P_{re} increases with the speed W. Upon reaching the speed W_0, the hydraulic resistance of the layer becomes equal to its mass, the layer is suspended, gas bubbles are observed jumping through it, and waves and splashes are observed on its free surface. In this state, the layer resembles a boiling liquid, due to which it is called pseudo-liquefied or boiling. The gas speed at which these phenomena begin is called the critical liquefaction speed - W_0 [[3], [4], [5], [6]].

There are a large number of theoretical and empirical formulas that describe with varying degrees of accuracy the moment of transition of a layer to a pseudo-fluidized state [[7], [8], [9], [10], [11], [12], [13], [14], [15]]. In this case, there are two different approaches to determining the rate of onset of liquefaction. One of them is based on the calculation of the hydrodynamic resistance of the pores and channels of a stationary layer; the second is based on the calculation of the speed of movement of individual particles of the layer with subsequent transition through certain pseudo-fluidized states.

The geometry of a layer consisting of spherical particles of the same size is determined by their diameter d and packing density. In practice, however, one often has to work with mixtures composed of particles of different sizes and shapes.

In this case, the conventional particle size d_0 is calculated using the formula:

$$d_0 = 1/\, \Sigma\, (\Delta_i/d_i) \tag{1}$$

where Δ_i – mass fraction of particles with diameter d_i (geometric mean of the cell sizes of adjacent sieves).

The surface area of a non-spherical particle S_P always exceeds that of a sphere S_S having the same volume. To take this difference into account, the concept of the shape factor is introduced:

$$F_S = \sqrt{\frac{S_s}{S_p}} < 1 \tag{2}$$

The numerical values of F_S are determined experimentally.

The concentration of solid material in the fluidized bed γ_{fl} can be calculated from the specific mass of solid particles γ_s and the porosity of the bed ε:

$$\gamma_{fl} = \gamma_s\, (1 - \varepsilon) \tag{3}$$

Then the bulk density of the stationary layer γ_0 with porosity ε_0 will be:

$$\gamma_0 = \gamma_s\, (1 - \varepsilon_0) \tag{4}$$

The transition of a stationary layer to a fluidized state occurs under the condition that the force of hydrodynamic resistance of the layer and the effective mass of suspended solid particles G_e are equal:

$$P_{fl}\, f_s = G_e, \tag{5}$$

where f_s – cross-sectional area of the layer.

Taking into account the layer penetrability ε and the buoyancy (Archimedes) force proportional to the difference in the specific masses of particles γ_s and the fluidizing agent γ, we obtain:

$$G_e = (\gamma_s - \gamma)\, (1 - \varepsilon) \cdot f_c \cdot H \tag{6}$$

where H – fluidized bed height.

Then

$$P_{fl} = (\gamma_s - \gamma)\, (1 - \varepsilon) \cdot H \tag{7}$$

Obviously, if the fluidizing agents are gases at low pressures, then $\gamma_s – \gamma \rightarrow \gamma_s$, and

$$P_{fl} = \gamma_T\, (1 - \varepsilon)\, H \tag{8}$$

To calculate the liquefaction rate at which the required pressure drop in the layer is ensured, the Todes formula [16] is widely used, with ε_0 = 0.4 having the form,

$$Re_0 = -\frac{Ar}{1400 + 5{,}22\sqrt{Ar}}, \tag{9}$$

where $Re_0 = W_0\, d/\mu$ – Reynolds criterion (μ – viscosity of the fluidizing agent);

$Ar = \frac{gd^3}{\mu^2} \cdot \frac{\gamma_T - \gamma}{\gamma}$ Archimedes criterion (g – acceleration of gravity).

Formula (9) has been tested in many studies over a wide range of variable changes and the results have shown good agreement with experimental data.

Effect of fluidization on gas flow rate and fluidized bed height. The intensity of the mutual movement of solid particles in a fluidized bed increases with the increase of the fluidization number. The speed of the pulsating movement of particles is usually at the level of tens of centimeters per second, and the free path of a particle between two collisions is measured in millimeters or centimeters. In this case, the movement of particles in the vertical direction is more intense than in the horizontal direction.

When gas bubbles appear in the layer, a significant role is played by the movements of not individual particles, but their aggregates. The chaotic movement of particles and aggregates leads to the equalization of temperature and other properties in the volume of the fluidized bed. Along with the chaotic (pulsating) movement of solid particles, depending on the configuration and geometric dimensions of the layer, a directed circulation of granular material may also occur in it, largely due to the bubbling of bubbles. Thus, with a ratio of the layer height to its diameter close to one, in devices of small diameters, predominantly an ascending movement of solid particles and a descending movement along the periphery of the layer is observed. In a layer of significant height, several such zones are formed along the height, which is quite clearly seen in Figure 2, which shows a typical scheme of circulation flows of solid material (dotted lines) and gas (solid lines) during fluidization in a small diameter device.

The intensity of mixing of solid particles determines the nature of mixing of the fluidizing agent in the fluidized bed, which also depends significantly on the properties of the fluidizing agent and the solid material.

The first of them is characterized by complete equalization of the properties of the working bodies, and the residence time of different portions of gas (or different solid particles in a layer with continuous input and output of the solid phase) is not the same.

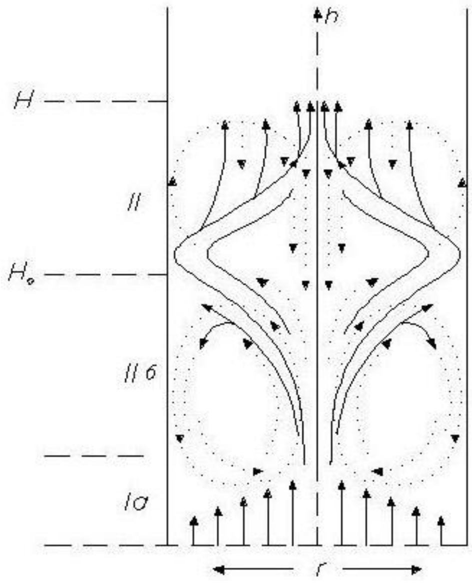


Figure 2 - Typical diagram of the circulation flows of solid material (dashed lines) and gas (solid lines) during fluidization in a small diameter apparatus

In contrast, ideal displacement systems are characterized by a flat front of movement of particles of the solid phase and the fluidizing agent (without longitudinal mixing) and, consequently, the same residence time, $z_{0,}$ in the working volume. The value of z_0 is determined by the flow of each phase and the working volume of the apparatus. For example, for a solid material it will be:

$$z_0 = G_w / G_c, \tag{10}$$

where G_w – weight of material in a layer;
G_c – weight consumption of material.

A complete mixing system is characterized by the probability of the time an object stays in the system x. For example, for solid particles:

$$x = e^{-\tau/z_o} \tag{11}$$

or

$$1 – x = 1- e^{-\tau/z_o}, \tag{12}$$

where x – is the weight fraction of particles whose residence time τ exceeds (or the probability that athe rticle will remain in the layer for a period of time τor longer);

(1-x) – is the probability that the particle's stay time, τ will not exceed.

The uniformity of the residence time of particles in the system can be greatly increased by dividing the fluidized bed into several sections. In this case, complete mixing of particles occurs in each section, but their transfer from section to section is possible only in one direction (from the previous section to the next) and in quantities equal to the flow rate.

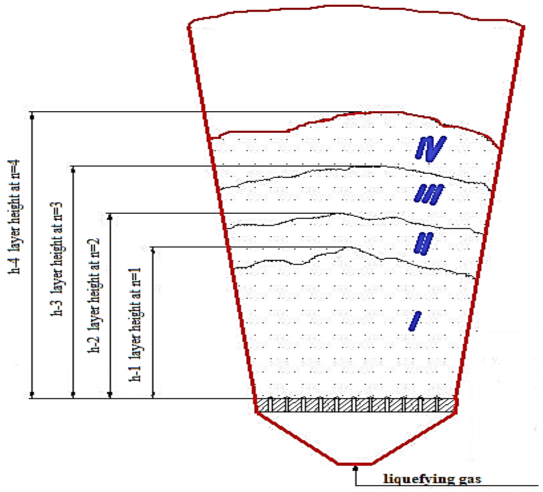
The residence time of particles in the N-section apparatus is determined by the formula:

$$1\text{-}x\text{=}1\text{-}e - \frac{\tau}{zo\backslash N}[\left[1+\frac{\tau}{zo\backslash N}\right]+$$

$$+\frac{1}{21}\left(\frac{\tau}{zo\backslash N}\right)^{2+.....+}\frac{1}{(N-1)}\left(\frac{\tau}{zo\backslash N}\right)^{N-} \tag{13}$$

Studies have shown that in a single-section apparatus, pseudo-fluidized systems are close to systems with complete mixing of particles. The exception is high layers, pseudo-fluidized in small-diameter apparatuses, where the process occurs in a piston mode. When using apparatuses with 8-10 sections, the movement of particles in the pseudo-fluidized layer approaches ideal displacement by its nature. Note also that the placement of nets, partitions, packing, etc., in the apparatus also brings the pseudo-fluidized layer closer to an ideal displacement system [[17], [18]].

In a fixed granular bed, the nature of the gas movement is close to ideal displacement, and in a fluidized bed, it occupies an intermediate position between ideal displacement and complete mixing. Mixing of the fluidizing agent is also prevented by the sectioning of the apparatus. Often, 4-6 sections (and sometimes fewer) are sufficient to bring the nature of the fluidizing agent movement closer to ideal displacement. Figure 3 shows a cross-section of a fluidized bed of solid particles at different fluidization numbers.



where: N = n * Pi, where N is the gas consumption for liquefaction; n is the fluidization number (0.5; 1; 2, etc.); P is the gas consumption at a fluidization number equal to 1

Figure 3 - Cross-section of a fluidized bed of solid particles at different fluidization numbers

A fluidization number equal to 1 corresponds to a gas flow rate at which solid particles pass into a state of fluidization over the entire surface of the fluidized bed mirror, i.e., there are no stagnant zones on the entire surface of the fluidized bed. When the fluidization number increases to 2, 3, etc., the gas flow rate for fluidization increases accordingly, as does the height of the fluidized bed of the material.

Results and Discussion

The main criterion for assessing the behavior of a component when heated in a fluidized bed can be the saturated vapor pressure of metal sulfides.

Figure 4 shows the dependence of the vapor pressure not only of the sulfides considered, but also of other substances that may be present in ores and concentrates.

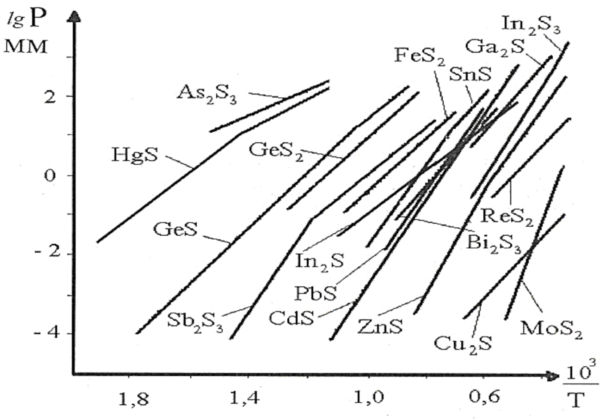


Figure 4 - Dependence of vapor pressure of non-ferrous metal sulfides on temperature

From the data on the dependence of the vapor pressure of non-ferrous metal sulfides on temperature, it follows that one of the promising methods for processing or enriching complex polymetallic raw materials, mainly sulfide, is pyroselection in a fluidized bed with the supply of a mixture of neutral gas with vapors of elemental sulfur into the layer of the material being processed, based on the difference in vapor pressure, dissociation of sulfides and their decomposition products.

In comparison with the processing of materials in conventional fluidized bed furnaces, where the liquefaction of the material is carried out by ordinary air, according to the proposed technology, the supply of a mixture of neutral gas with elemental sulfur into the bed is characterized by the formation of a much smaller amount of reaction gases, and therefore a sharp reduction in the emitted gases and better conditions for the condensation of vapors, reduced dust entrainment and the possibility of distilling off volatile components at lower temperatures [[19], [20]].

Microscopic examination of the residues revealed that the size of the particles located deep in the layer remained unchanged.

The influence of the layer height, duration and temperature of the initial sample processing on the evaporation rate of Sb_2S_3 was investigated. The experiments were conducted with antimony sulfide of a size of - 0.5 + 0.1 mm at a temperature of 923 – 1023 K, respectively. The layer height varied from 5 to 15 mm. The experimental data are presented in Table 1.

Table 1 - Effect of grain size on the evaporation rate of Sb_2S_3

Experience number	**T, K**	**Grain size, mm**	**Evaporation rate of Sb_2S_3, g/cm^2·sec**
1	923	- 2.0 + 1,0	$0.070 \cdot 10^{-5}$
2	923	- 0.5 + 0.25	$0.072 \cdot 10^{-5}$
3	923	- 0.1 + 0.09	$0.072 \cdot 10^{-5}$
4	923	- 0.1 + 0.09	$0.072 \cdot 10^{-5}$
5	1023	- 2.0 + 1.0	$5.45 \cdot 10^{-5}$
6	1023	- 0.5 + 0.25	$5.62 \cdot 10^{-5}$
7	1023	- 0.25+ 0.1	$5.66 \cdot 10^{-5}$
8	1023	- 0.1 + 0.09	$5.70 \cdot 10^{-5}$

For comparison, the results of the study of the kinetics of antimony sulfide Sb_2S_3 evaporation from a fixed bed, we conducted experiments on the evaporation of antimony sulfide from a fluidized bed. The results of the experiments on the kinetics of antimony sulfide evaporation from fixed and fluidized beds are given in Table 2.

As can be seen from the data provided, X-ray structural analysis established that in both cases (for freshly ground and calcined), the preparation consisted of Sb_2S_3 with a rhombic structure.

Microscopic examination revealed that the sample was crystals of perfect shape with smooth edges. Apparently, the decrease in the evaporation rate when the sample was kept in its vapor occurred due to a decrease in crystal defects, a noticeable elimination of microscopic irregularities and surface roughness.

When studying the effect of foreign impurities on the kinetics of Sb_2S_3 evaporation, it was found that the rate of evaporation of a volatile component from its mixture with an inert substance can be affected by the physical properties of the surface of the non-volatile, its adsorption capacity, gas permeability and the nature of the bonds. The hydrodynamic resistance of the layer of non-volatile granular material formed in the process of incongruent evaporation of the mixture with the non-volatile component should have a great influence on the rate of evaporation from the mixture. Thus, it was established that the rate of antimonite evaporation does not depend on the height of the layer and the size of the grains, but depends on the preliminary treatment of the preparation.

Figure 5 shows the kinetic dependences of the Sb_2S_3 evaporation rate on the duration and temperature of the treatment.

Table 2 - Effect of sample layer height and experimental temperature on the evaporation rate of Sb_2S_3

Experience number	Temperature, K	Layer height, mm	Sb_2S_3 evaporation rate g/cm^2·sec	Lg K g/cm^2·sec Sb_2S_3	evaporation rate g/cm^2·sec	Lg K g/cm^2·sec
			Fixed layer		Fluidized bed	
1	923	5	$0.070 \cdot 10^{-5}$	- 5.154	$0.56 \cdot 10^{-5}$	- 4.251
2	923	15	$0.066 \cdot 10^{-5}$	- 5.180	$0.50 \cdot 10^{-5}$	- 4.301
3	1023	5	$5.450 \cdot 10^{-5}$	- 3.263	$3.28 \cdot 10^{-4}$	- 3.484
4	1023	15	$5.420 \cdot 10^{-5}$	- 3.266	$3.19 \cdot 10^{-4}$	-3.496
5	1073	5	$0.9 \cdot 10^{-4}$	- 4.045	$7.80 \cdot 10^{-4}$	- 3.107
6	1073	15	$0.84 \cdot 10^{-4}$	- 4.075	$7.56 \cdot 10^{-4}$	- 3.121
7	1123	5	$2.4 \cdot 10^{-4}$	- 3.619	$1.2 \cdot 10^{-3}$	- 3.920
8	1123	15	$2.32 \cdot 10^{-4}$	- 3.634	$1.16 \cdot 10^{-3}$	- 3.935
9	1173	5	$4.2 \cdot 10^{-4}$	- 3.376	$2.8 \cdot 10^{-3}$	- 3.552
10	1173	15	$4.15 \cdot 10^{-4}$	- 3.381	$2.74 \cdot 10^{-3}$	- 3.562
11	1223	5	$8.7 \cdot 10^{-4}$	- 3.060	$4.75 \cdot 10^{-3}$	- 2.323
12	1223	15	$8.6 \cdot 10^{-4}$	- 3.065	$4.6 \cdot 10^{-3}$	- 2.337

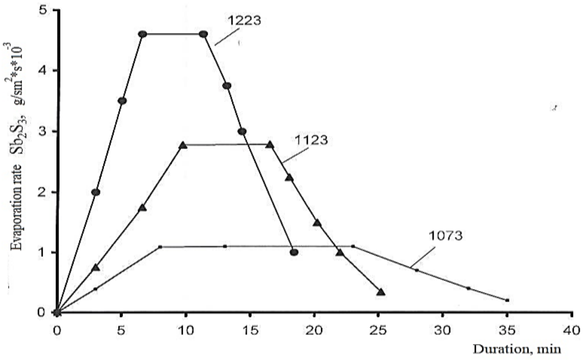


Figure 5 - Kinetic dependences of the evaporation rate of Sb_2S_3 on the duration of the experiment at temperatures of 1073, 1123 and 1223 °K

The data obtained show that the evaporation rate of pure antimony sulfide reaches its maximum value in fairly short periods of treatment time, this is especially noticeable for high temperatures. Thus, if at a temperature of 1073 K the evaporation rate during a treatment time of 7 minutes was 0.98 $g/cm^2 \cdot sec \cdot 10^{-3}$, then at a temperature of 1223 K it reached 4.7 $g/cm^2 \cdot sec \cdot 10^{-3}$, i.e. it exceeded the evaporation rate of Sb_2S_3 for the same treatment time by almost 5 times.

Conclusions

Hydrodynamic and heat transfer processes in a fluidized bed were investigated, and the feasibility of forming a stable fluidized bed through throttling of an inert gas or a mixture of inert gas and sulfur vapor was experimentally confirmed.

The optimal process parameters were determined as follows: particle size between 0.25 and 0.5 mm, bed height between 5 and 15 mm (minimal influence was observed), and operating temperature between 1023 and 1073 K with a gas mixture consisting of nitrogen and sulfur vapor.

It was established that the specific evaporation rate of Sb_2S_3, calculated per unit surface area, is nearly identical for particles of different sizes. At the same time, a decrease in particle size increases the total surface area, which leads to a higher overall evaporation rate of the sample. This effect was experimentally confirmed (Table 1, temperature 1023 K).

Evaporation rates of Sb_2S_3 in a fluidized bed are 7-9 times higher than those in a fixed bed due to enhanced heat and mass transfer. Vapor removal rates from the fluidized bed significantly exceed those from a fixed bed, improving process efficiency.

The evaporation rate does not depend on the bed height within the range of 5-15 mm and occurs mainly from the surface of the particles.

Antimony recovery in sublimates has increased by 2-3% compared to conventional technology, reaching 98-99%, due to the suppression of Sb_2O_5 formation.

The proposed technology of introducing an inert gas with sulfur vapor into the fluidized bed offers ecological and technological benefits: reduced formation of secondary reaction gases, lower dust entrainment, improved vapor condensation, and efficient distillation of volatile components at lower temperatures.

Conflicts of interest. On behalf of all authors, the corresponding author states that there is no conflict of interest.

CRediT author statement: **Sh.Akilbekova**: Conceptualization, Methodology, Supervision; **G.Moldabayeva**: Data curation, Writing-Original draft preparation; **S.Myrzalieva**: Visualization, Investigation; **N.Seidakhmetova**: Software, Validation

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Қабылданды: *23 қыркүйек 2025*

ТҮЙІНДЕМЕ

Түсті, асыл және сирек металдарға сұраныстың артуы минералдық шикізатты, соның ішінде алтын-сурьма кендері мен концентраттарын барынша кешенді әрі тиімді пайдалануды қажет етеді. Перспективалы бағыттардың бірі – дәстүрлі технологиялармен салыстырғанда жылу- және масса алмасу үдерістерін анағұрлым қарқынды жүргізетін жалған сұйықтанған қабаттағы пирометаллургиялық қайта өңдеу болып табылады. Осы зерттеудің мақсаты – алтын-сурьма кендері мен концентраттарынан сурьма сульфидінің (Sb_2S_3) булану кинетикасын жалған сұйықтанған қабат жағдайында зерттеу. Эксперименттер температураны (923–1223 К), бөлшектер өлшемін (0,09–2,0 мм) және қабат қалыңдығын (5–15 мм) өзгерту арқылы жүргізілді. Зертханалық қондырғы құрамында азот пен күкірт буларынан тұратын газ қоспасының реттелетін ағынымен жабдықталған жалған сұйықтанған қабатты реактор болды. Булану жылдамдықтары гравиметриялық әдіспен анықталып, рентгенқұрылымдық және микроскопиялық талдаулармен расталды. Зерттеу нәтижесінде Sb_2S_3 булану жылдамдығы жалған сұйықтанған қабатта қозғалмайтын қабатқа қарағанда 7–9 есе жоғары екендігі анықталды, бұл жылу- және масса алмасу үдерістерінің қарқындануымен түсіндіріледі. 1023 К температурасында бөлшектер өлшемі кішірейген сайын олардың меншікті бетінің ұлғаюына байланысты жалпы булану жылдамдығы артты, ал ауданның бірлігіне шаққандағы жылдамдық өзгеріссіз қалды. 5-15 мм шегінде қабат биіктігі айтарлықтай әсер етпейді. Сурьманы ұшырындыларға шығару дәрежесі 98–99% жетіп, дәстүрлі технологиямен салыстырғанда 2–3% жоғары болды, бұл Sb_2O_5 түзілуін болдырмау есебінен мүмкін болды. Алынған нәтижелер инертті газ бен күкірт буларын жалған сұйықтанған қабатқа беру тиімділігін дәлелдейді, бұл зиянды газдардың бөлінуін азайтуға, шаңның таралуын төмендетуге және ұшқыш компоненттерді төменірек температурада тиімді айдауға мүмкіндік береді.

Түйін сөздер: Күкірттендіре күйдіру, тотықтырып күйдіру, бейтарап газ, конденсация.

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АННОТАЦИЯ

Растущий спрос на цветные, благородные и редкие металлы обусловливает необходимость более комплексного и эффективного использования минерального сырья, в том числе золото-сурьмяных руд и концентратов. Перспективным направлением является пирометаллургическая переработка в псевдоожиженном слое, обеспечивающая более интенсивный тепло- и массообмен по сравнению с традиционными технологиями. Целью данного исследования являлось изучение кинетики испарения сульфида сурьмы (Sb_2S_3) из золото-сурьмяных руд и концентратов в условиях псевдоожиженного слоя. Эксперименты проводились при изменении температуры (923–1223 К), размера частиц (0,09–2,0 мм) и толщины слоя (5–15 мм). Лабораторная установка включала реактор с псевдоожиженным слоем, снабжённый регулируемым потоком газовой смеси на основе азота и паров серы. Скорости испарения определялись гравиметрическим методом и подтверждались рентгеноструктурным и микроскопическим анализом образцов. Установлено, что скорость испарения Sb_2S_3 в псевдоожиженном слое в 7–9 раз выше, чем в неподвижном, что связано с интенсификацией процессов тепло- и массообмена. При температуре 1023 К общая скорость испарения возрастает с уменьшением размера частиц вследствие увеличения их удельной поверхности, тогда как скорость, отнесённая к единице площади, остаётся постоянной. Высота слоя в пределах 5–15 мм существенного влияния не оказывает. Степень извлечения сурьмы в возгоны достигала 98–99%, что на 2–3% выше по сравнению с традиционной технологией, благодаря подавлению образования Sb_2O_5. Полученные результаты подтверждают эффективность подачи инертного газа с парами серы в псевдоожиженный слой, что обеспечивает снижение газовых выбросов, уменьшение пылеуноса и эффективную дистилляцию летучих компонентов при более низких температурах.

Ключевые слова: Сульфидирующий обжиг, окислительный обжиг, нейтральный газ, конденсация.

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