Steam Reforming of Residual Oils in Fluidized Bed of Calcined Dolomite

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Steam reforming of eight different residual oils in a fluidized bed of calcined dolomite was investigated under atmospheric pressure to produce fuel gas. Sufficient gasification was achieved at 900 °C under the following reaction conditions: steam to carbon ratio, 2~9 by weight and residence time, 0.4 sec or larger. Almost 45 % of the carbon in the oils was reformed to CO and CO₂. The yield of gaseous hydrocarbons such as CH₄ and C₂H₄ varied from one oil to another, and it correlated well to the content of saturates in the oil. As a result, the maximum gas yield of 91.3 wt% was obtained.

In the production of CO and CO₂ by steam reforming, the carbon deposited on dolomite played an important role. The steam reforming on dolomite was explained in terms of a mechanism, where the carbon accumulated on the active sites served as intermediate.

During gasification, hydrogen sulfide was removed simultaneously by CaO in dolomite. The concentration of H₂S in the product gas was almost independent of the sulfur content of the feed oil and at 900 °C it was nearly 0.1 vol%. It was shown thermodynamically that the concentration of H₂S is reduced to its equilibrium value according to the reaction: $\text{H}_2\text{S} + \text{CaO} \rightleftharpoons \text{H}_2\text{O} + \text{CaS}$.

1. Introduction

Considerable rise in prices, problems of crude oil supply and the more demanding requirements of environmental protection have recently given a new incentive and impetus to further research and technical development for more efficient utilization of residual oils and coal, and various gasification processes have been proposed¹⁾. Among the processes, the use of dolomite or some other basic materials would be environmentally sound because of their capabilities to desulfurize H₂S^{2),3)}. Actually, basic materials have been used as a sorbent to remove sulfur at high temperature in the following two gasification processes. One is the Chemically Active Fluid Bed process (the CAFB process), developed by Esso Research Center of England, to gasify heavy fuel oil in the fluidized bed of lime particles for production of hot, low sulfur fuel gas⁴⁾. The other is the fluidized bed coal gasification process using dolomite, developed by Westinghouse Research Laboratories, which is a system for combined cycle power generation⁵⁾.

Furthermore, dolomite or other basic materials have been suggested^{6),7),15)~17)} as active catalysts for gasification of residual oils at high temperatures.

A typical application of these catalysts for gasification has been made by Toyo Engineering Corporation and Tokyo Gas Company in their THR process, a tubular steam reforming process, using a catalyst based on calcium oxide to produce synthesis gas and fuel gas from crude oil⁸⁾. However, a fundamental understanding of the catalytic behavior of these materials has lagged far behind their industrial applications.

The authors^{9),10)} have studied steam gasification of residual oils using a fluidized bed of dolomite to produce fuel gas with low sulfur content. Dolomite was chosen as an effective catalyst and sulfur acceptor with good resistance to attrition. The steam reforming of pure and lower hydrocarbons has also been investigated to elucidate the nature of the carbon deposited on a basic catalyst such as dolomite¹¹⁾. The present paper reviews our research work on catalytic behavior of dolomite and its desulfurization ability for steam reforming of residual oils.

2. Experimental

2.1 Materials

Table 1 shows the constitution and properties of feed residual oils used in the present work which were supplied by Nippon Oil Co., Ltd. and Toa Nenryo Kogyo K.K.. This table includes the results of fractionation of residual oils carried out according to the scheme¹²⁾ illustrated in **Fig. 1**.

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Table 1 Constitution and Properties of Feed Residual Oils

	Symbol	Sp. Gr. (15/ 4°C)	C	H (wt%)	S	C. C. R. * (wt%)	H/C (atom/ atom)	Physical Composition (wt%)			
								Saturates	Aromatics	Resins	Asphaltenes
Minas topped bottom	Mi. Top.	0.8875	86.4	13.2	0.14	4.52	1.83	72.8	14.7	12.0	0.5
Arabian light topped bottom	A/L Top.	0.9416	83.4	11.8	2.81	7.77	1.70	54.0	30.9	10.9	4.2
Iranian heavy topped bottom	Ir/H Top.	0.9636	85.6	11.4	2.55	10.65	1.60	47.0	31.9	17.6	3.5
Minas vacuum bottom	Mi. Vac.	0.9454	87.2	12.1	0.28	12.88	1.66	53.3	23.0	22.8	0.9
Arabian medium topped bottom	A/M Top.	0.9508	83.4	11.9	4.25	13.00	1.71	48.4	33.7	13.8	4.1
Berri vacuum bottom	Ber. Vac.	0.9921	85.6	11.0	3.26	14.85	1.54	35.3	46.2	14.5	4.0
Arabian light vacuum bottom	A/L Vac.	1.0118	85.3	10.6	3.87	17.67	1.49	29.5	47.6	22.8	0.9
Arabian medium vacuum bottom	A/M Vac.	0.9990	84.2	10.3	5.50	18.30	1.47	30.1	47.5	17.0	5.4

* C. C. R. means Conradson Carbon Residue.

Fig. 1 Schematic Diagram for Fractionation of Residual Oil

Dolomite from Kuzuo, Tochigi Prefecture, was used as catalyst after it was calcined at 950 °C for 4 hr and crushed (32~65 mesh). The calcination temperature of 950 °C was found most effective in range (850~1,270 °C) tested. The composition of calcined dolomite is as follows: 0.75 wt% SiO₂, 0.30 wt% Al₂O₃, 0.11 wt% Fe₂O₃, 65.43 wt% CaO and 33.41 wt% MgO.

2.2 Apparatus and Procedures

Gasification was carried out over the temperature range of 700~900 °C in a fluidized bed reactor under atmospheric pressure. A schematic diagram of the reactor assembly and details of the reactor tube are shown in **Figs. 2** and **3**, respectively. The reactor tube was of 42 mm ϕ I. D. SUS-27 stainless steel and its bed depth was 400 mm. Stainless beads (3 mm ϕ) were placed at the conical-bottom of the reactor that served as bed support and gas distributor. Feed oil was admitted from the bottom part of the catalyst bed through an atomizer by use of a microfeeder. From the nozzle of the atomizer, the oil was atomized with steam produced in the steam generator. To prevent the feeding pipe plugging by coking, the atomizer was cooled by

1. Steam generator
2. Feed oil atomizer
3. Reactor
4. Electric furnace
5. Steam preheater
6. Ice trap
7. Mist remover
8. H₂S trap
9. H₂S remover
10. Wet gas meter

Fig. 2 Schematic Diagram of Reactor Assembly

Fig. 3 Details of Reactor

circulating a coolant and periodically rotating a piano wire inserted into the pipe. Steam for fluidization was superheated at 500 °C in the pre-heating section and admitted from the bottom of

the reactor to the catalyst bed through the stainless beads gas distributor. The catalyst was heated in a stream of nitrogen up to 500 °C, then further heated in a stream of steam up to the desired temperature. After keeping the catalyst at the temperature for 30 min, oil feed was started.

After condensation and removal of liquid products and excess steam with the ice trap, the product gas passed through a solution of ammine complex of zinc to absorb the hydrogen sulfide produced after which the gas volume was measured with a wet gas meter. The content of H₂S in the solution was determined by the JIS method K 0108. Gas sampling was done every 40 min. The composition of gas was determined by use of gas chromatography.

Initially, the yield of product gas increased gradually with reaction time and attained after 60~80 min, a steady value which was adopted in the following figures and tables. The yield of hydrogen sulfide did not change within 150 min of reaction time, because the amount of dolomite used was very large in comparison with that of sulfur contained in the feed oil fed.

In these experiments, it was difficult to get an accurate material balance because the yield of liquid products was too small to be determined precisely. As this study was designed primarily to produce fuel gas, discussion will be limited to the results of and based on gasification products.

3. Results and Discussion

3.1 Effect of Reaction Conditions on Gasification

To determine reaction conditions for gasification, reaction of Arabian Medium topped bottom residue were investigated under various conditions. **Table 2** summarizes the results obtained at various reaction temperatures. The total gas yield was determined by summing up the yields of gaseous hydrocarbons, CO, CO₂, H₂ and H₂S. Here, the yield of hydrogen would mean the amount of hydrogen yielded from the oil. The amount of hydrogen yielded from steam was calculated using the yields of CO and CO₂.

The yield of CO plus CO₂ increased with increasing reaction temperature, and the total gas yield reached about 90% at 900 °C. The liquid product obtained in the reaction was found to be light oil which contained small quantities of tar and pitch fractions. In this table, the yield of hydrogen obtained at 700 °C was remarkably high value (8.6 wt%). This high value was due to the fact that, under these conditions, a steady state was attained only after a long period of time; as a result, part of

Table 2 Effect of Reaction Temperature

Catalyst Dolomite			
Steam Ratio (g/g) 2.2			
Cata. Weight (g) 200			
React. Temp. (°C)	700	800	900
Feed Rate (g/hr)	67.8	67.8	71.7
Residence Time (sec)	0.75	0.75	0.46
Gas Vol. (ml/g-feed)	1,485	2,303	2,835
Gas Yield (wt %)			
Total	57.4	74.9	90.3
Hydrocarbon	41.1	39.3	41.1
CO+CO ₂	7.7	33.8	47.9
H ₂	8.6	1.8	1.3
Carbon Recov. *1)	49.5	75.0	91.6
H ₂ S Yield *2)	6.2	8.5	8.2
Calorific Value *3)	5,521	4,049	3,874
Thermal Yield *4)	779	887	1,045
Gas Composition (vol %)			
H ₂	69.0	57.9	54.6
CO	1.6	11.0	18.3
CO ₂	8.1	16.4	13.2
CH ₄	7.5	6.7	8.4
C ₂ H ₆	2.1	0.9	0.6
C ₂ H ₄	6.4	5.9	4.8
C ₃ H ₆	2.5	0.8	0.1
C ₄ +	2.8	0.4	0.1
H ₂ S	0.12	0.11	0.08

*1) as gas (C mol of gas/C 100moles in feed oil)

*2) (H₂S mol/S 100moles in feed oil)

*3) (kcal/Nm³)

*4) (JHU/kcal)

the hydrogen produced from steam was included in the yield of hydrogen since dolomite absorbed CO₂ produced at this temperature. The X-ray examination of dolomite used at 700 °C showed that CaO in dolomite changed to CaCO₃, whereas MgO remained unchanged.

Figure 4 shows the results obtained at 900 °C with various residence times. Here, the residence time (θ) was calculated according to Eq. (1).

$$\theta = \frac{V_R - (W/d)}{(Q_1 + Q_2)(273 + t)/273} \quad (1)$$

V_R : efficient reactor volume (ml)

W : catalyst weight (g)

d : density of catalyst (g/ml)

t : reaction temperature (°C)

Q_1 : volume of steam (Nml/sec)

Q_2 : volume of product gas (Nml/sec)

Efficient reactor volume was obtained by multiplying the inner cross section of reactor tube and the depth of the catalyst layer. The depth of the

Reaction temperature : 900°C

Fig. 4 Effect of Residence Time on Gas Yield

catalyst layer was determined by measuring the region of a uniform temperature along the length of the reactor. The residence time was varied mainly by changing the catalyst volume. The yield of CO plus CO₂ increased with increasing residence time, whereas the yield of hydrocarbon remained unchanged.

Effects of the ratio of steam to feed oil (the steam ratio) were examined at 900 °C, where the feed rate of steam was fixed and that of oil varied. **Figure 5** shows typical results obtained. The yield of CO plus CO₂ was almost independent of the steam ratio from 3.8 to 7.5. The smaller yield at the steam ratio of 1.9 may partly stem from the reaction that could not be carried out in a shorter residence time to obtain good fluidization of dolomite.

3.2 Gasification of Various Residual Oils

The reactivity of eight different residual oils was compared at the reaction temperature of 900 °C with steam ratio of 2.8~3.0 and residence time of

Reaction temperature : 900°C
Residence time : 0.22~0.34 sec

Fig. 5 Effect of Steam Ratio on Gas Yield

Fig. 6 Relation between Gas Yield and Saturate Content of Feed Oil

Fig. 7 Relation between Content of Asphalics and Percent of Feed Oil not Converted into Gaseous Products

0.4~0.5 sec. These results are shown in **Figs. 6** and **7**. Although various factors characterizing residual oils have been used, the content of saturates was used here. **Figure 6** shows the relation between gas yields and saturate contents in the residual oils. The yield of CO plus CO₂ was almost unchanged from one oil to another, being approximately 45%. On the other hand, the yield of hydrocarbons proportionally increased from 26.5% to 51.2% when the saturate content increased from 29.5% to 72.8%. It seems that gaseous hydrocarbons such as C₂H₄ and CH₄ were produced mainly through cracking of saturate fraction of the feed oil. The percent of feed oil which was not converted into gaseous products (100-total gas yield) decreased with increase in the content of saturates. As shown in **Fig. 7** it was correlated linearly to the content of asphalics (asphaltenes plus resins). This linear relationship suggests that heavier constituents such as asphalics hardly were converted into gaseous products during gasification.

3.3 Carbon Deposition on the Catalyst

When residual oils were gasified, the product gas yield increased initially and reached a steady level

Fig. 8 Variation of Gas Yield with Time

within about 1 hr at 900 °C. A typical time dependent variation of gas yields is shown in **Fig. 8**. The period required to attain a steady activity varied with the amount of catalyst or feed rate of residual oil. These results suggest that reaction intermediate is accumulated on the catalyst surface in the initial period of reaction and the amount of accumulated intermediate reaches a steady level. This intermediate seems to be the carbon deposited on the active sites of the catalyst. Owing to the deposition of carbon, the color of the catalyst used for 3 hr at 900 °C turned grayish

The amount of carbon deposited at 900 °C was not sufficient for its exact determination. The accumulation of carbon at 700 °C was, however, much more extensive. To determine the relation between the amount of deposited carbon and gasification activity, reaction was carried out at 700 °C. **Figure 9** shows the time dependent variation of the amount of deposited carbon and activity in the gasification of vacuum bottom of Berri. The amount of carbon deposited on the catalyst was determined by with drawing the catalyst from the reactor after a certain reaction period and burning off the carbon. The amount of deposited carbon initially increased with reaction period until a maximum is reached then it decreased. The variation in the yield of hydrogen corresponded to the variation in the amount of deposited carbon, indicating that the deposited carbon played an important role probably as an intermediate in steam reforming.

The yield of CO₂ increased remarkably after about 60 min of reaction. At the initial stage of reaction,

Fig. 9 Variation of Gasification Activity and Carbon Deposition with Reaction Time

CaO absorbed CO₂ produced because the temperature was below the decomposition temperature of CaCO₃ (825 °C). Therefore, the increase in the yield of CO₂ seems to be due to the saturation of absorption of CO₂. After saturation, both carbon deposition and hydrogen production decreased. Thus, CaO loses its ability not only to deposit carbon but also to catalyze gasification reactions when it absorbs CO₂.

3.4 Nature of Deposited Carbon

To elucidate the nature of deposited carbon, this study was supplemented with the investigation of steam reforming of pure and lower hydrocarbons on some basic catalysts such as calcium oxide in a conventional flow system.

Table 3 shows the activity of various catalysts and the reactivity of carbon deposited on them. The catalytic activity was expressed by the space

Table 3 Comparison between Rate of Deposited Carbon-Steam Reaction and Catalytic Activity of Various Catalysts for Steam Reforming at 900°C

Catalyst	r^* (mmol/g·hr)	Space Time Yield (mmol/g·hr)
Dolomite	14.2	15.3
Calcium Oxide	16.9	14.8
Magnesium Oxide	13.4	12.8
Active Alumina	5.3	5.0
α -Alumina	2.6	2.8
Silica Gel	0.9	0.6

* The rate of reaction between steam and carbon deposited from *n*-hexane in the absence of steam.

time yield which was calculated by multiplying the fractional conversion of *n*-hexane to CO plus CO₂ and space velocity. The rate of reaction between deposited carbon and steam(*r*) was determined from the rates of evolution of CO, CO₂ and CH₄ when only steam was admitted to carbon deposited from *n*-hexane onto the catalyst. Moreover, the following relationship was found among the quantities of H₂, CO, CO₂ and CH₄ produced in the deposited carbon-steam reaction.

$$[H_2] = [CO] + 2[CO_2] - 2[CH_4]$$

This relationship means that hydrogen is produced according to the reactions (1) and (2).

Therefore, it is suggested that the deposited carbon contains hydrogen but slightly.

Among the catalysts tested, basic materials such as dolomite and calcium oxide were found to be more effective for steam reforming than alumina and silica gel. Furthermore, it was found that catalytic activity was closely related to the rate of reaction between steam and carbon deposited in the absence of steam. The results suggest that steam reforming of *n*-hexane proceeds via highly reactive carbon deposited on the catalyst as an intermediate. Saito et al.^(13,14) have proposed the following reaction scheme for steam reforming of *n*-hexane on silica-supported nickel catalysts and they have concluded that the steam reforming proceeds mainly via the reaction between steam and the adsorbed hydrocarbon species on the catalyst (step I).

According to this reaction scheme, steam reforming on dolomite proceeds mainly through step II and step III.

Based on the assumption that deposited carbon is the intermediate of steam reforming, the reactivity of carbon deposited from various pure hydrocarbons was examined by use of the calcium oxide catalyst at 900 °C. The results are shown in **Figs. 10 and 11**. In these figures, the rate of reaction between carbon and steam is plotted against the amount of carbon deposited. Cracking of various aromatic hydrocarbons and aliphatic hydrocarbons gave rise to deposited carbon of almost the same reactivity. Furthermore, the reactivity of carbon deposited

Fig. 10 Reactivity of Carbon with Steam Deposited from Hydrocarbons (mainly Aliphatic Hydrocarbons)

Fig. 11 Reactivity of Carbon with Steam Deposited from Aromatic Hydrocarbons and Heavy Oil

from vacuum bottoms of Berri was almost the same as that of the carbon deposited from various pure hydrocarbons as shown in **Fig. 11**. Accordingly, these results suggested that the nature of carbon deposited on calcium oxide was not affected by the species of hydrocarbons cracked.

3.5 Desulfurization Capability of Dolomite

Figure 12 shows the effect of residence time on H₂S yield in the product gas in gasification of topped bottoms of Arabian Medium. Here, H₂S yield was defined as the percent conversion of sulfur contained in the oil to hydrogen sulfide. Therefore, the smaller the yield of H₂S is the higher the ability of dolomite for desulfurization at the same gas yield. H₂S yield slightly decreased with increasing residence time up to about 0.4 sec. It seems that this is a rapid desulfurization reaction and is thermodynamically controlled. X-ray examinations of dolomite used gasification showed CaO in dolomite partly changed to CaS, whereas MgO was unchanged. This result shows that MgO plays no part in fixing sulfur and only CaO is capable of accepting sulfur from hydrogen sulfide according to the reaction: $CaO + H_2S \rightleftharpoons CaS + H_2O$, because the

Fig. 12 Effect of Residence Time on H_2S Yield³

Fig. 13 Effect of Steam Ratio on H_2S Yield

equilibrium for conversion of MgO to MgS is unfavorable at all reaction conditions that are of interest to us as suggested by Squires et al.¹⁸⁾ **Figure 13** shows the effect of steam ratio on desulfurization during gasification of topped bottom of Arabian Medium. In this figure, a dotted line represents the equilibrium yield of H_2S on the assumption that all the sulfur in the feed oil is converted to hydrogen sulfide, whose yield decreased with decrease in the steam ratio, always being very close to the equilibrium yield of H_2S .

Percent desulfurization and H_2S concentration in the product gas obtained in gasification of various residual oils are plotted in **Fig. 14** against the sulfur content in the feed oil. Percent desulfurization increased as the sulfur content increased, while the concentration of H_2S was almost unchanged. **Table 4** shows the concentration of H_2S in the product gas and the ratio of partial pressure of steam to

Fig. 14 Percent Desulfurization in Gasification of Various Oils

Table 4 H_2S Concentration in Gaseous Products

	S Content (wt %)	H_2S Concentration in Product Gas (vol %)	P_{H_2O}/P_{H_2S}
Mi. Top.	0.14	0.04	2,656
Mi. Vac.	0.28	0.07	1,326
Ir/H Top.	2.55	0.21	446
A/L Top.	2.81	0.09	642
Ber. Vac.	3.26	0.17	429
A/L Vac.	3.87	0.15	485
A/M Top.	4.25	0.09	551
A/M Vac.	5.50	0.14	649

that of hydrogen sulfide (P_{H_2O}/P_{H_2S}) under the reaction conditions. Here, the partial pressure of steam was calculated from the amount of steam obtained by subtracting the amount of steam consumed in the formation of CO and CO_2 from that of steam fed. The concentration of H_2S in the product gas was always about 0.1 vol. % regardless of the sulfur content in the feed oil. The P_{H_2O}/P_{H_2S} ratio was almost constant except when residual oils of Minas of extremely low sulfur content were used. These results imply that CaO in dolomite reduces the concentration of H_2S in the product gas to its equilibrium value according to the reaction: $CaO + H_2S \rightleftharpoons CaS + H_2O$.

Furthermore, it is predicted that SO_2 concentration in the combustion off-gas becomes about 200 ppm when the fuel gas obtained in this study is burnt with excess air (105% of stoichiometric amount required for combustion). This calculation is made

on the assumption that hydrogen sulfide is completely converted into SO_2 during combustion. This value of SO_2 concentration is almost 2/3 of the SO_2 concentration resulted when Minas topped bottom is burnt with 105% of stoichiometric amount of air required for combustion.

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要 旨

焼成ドロマイト流動層による残油の水蒸気改質

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現在重質燃料のガス化はエネルギー資源の低公害有効利用法として注目されている。この重質燃料のガス化プロセスは種々提案¹⁾されているが、ドロマイトや石灰を用いるプロセスはこれらの物質が脱硫能力を有することから最も経済的でしかも環境問題に対して有効なプロセスであり、第3のガス化法として最も期待されているものである。実際、残油のガス化法として石灰の流動層を用いるイギリスの Esso Research Center の開発による CAFB 法⁴⁾や Westinghouse Research Laboratories によって開発されたドロマイトを用いる石炭の流動層によるガス化プロセス⁵⁾等がある。さらにこれらの塩基性物質が高温での残油のガス化に対し活性な触媒であることも従来の研究^{6), 7) 15)~17)}から明らかにされている。この塩基性物質を触媒としてガス化プロセスに用いたものとして東洋エンジニアリングと東京瓦斯が開発した THR 法⁸⁾や三井造船と三井鉱山コークスで開発した M-Gas 法がある。

しかし工業的な開発研究に比較してこれらの塩基性物質の触媒作用や反応に関する基礎的研究は少ない。そこで本論文では燃料ガス製造を目的として焼成ドロマイト流動装置を用い常圧下での8種の異なる残油の水蒸気改質の研究を行った。

まず反応温度 (Table 2), 滞留時間 (Fig. 4), 水蒸気比 (Fig. 5) 等の諸条件を検討したがその最適条件は反応温度 900°C, 滞留時間 0.4 秒以上, 水蒸気比 2~9 である。この条件下では 90 wt % 以上のガス収率が得られた。

いずれの残油を原料油に用いてもほぼ 45% の炭素が CO ,

CO_2 に改質された (Fig. 6)。一方 CH_4 , C_2H_4 のような気体炭化水素の収率は使用した残油によって変化した。そしてその収率は残油の飽和成分の含有量と良い相関性を示した (Fig. 6)。この相関関係から気体炭化水素は主として残油中の飽和成分の分解により生成することが示唆された。気体生成物 ($\text{CO} + \text{CO}_2$, 炭化水素, H_2 , H_2S) へ転化されない原料残油の割合とアスファルテックス成分 (アスファルテン分+レジン分) の含有量とが良好な対応を示した (Fig. 7)。このことからアスファルテックス成分のように重質な成分はガス化されにくいことが推定された。

水蒸気改質により生成する水素の経時変化と析出炭素量のそれとが対応すること (Fig. 9), 析出炭素と水蒸気との反応速度と水蒸気改質活性 (CO , CO_2 の生成速度) とが良く対応すること (Table 3) から、ドロマイト上での水蒸気改質は中間体が活性点上に析出した炭素であるという機構によって説明された。

ガス化中同時に硫化水素はドロマイト中の CaO によって除去された。滞留時間の影響 (Fig. 12) および水蒸気比の影響 (Fig. 13) の検討から脱硫反応速度は大きく、次に示される反応の熱力学的平衡近くで抑えられていることが示された ($\text{CaO} + \text{H}_2\text{S} \rightleftharpoons \text{CaS} + \text{H}_2\text{O}$)。生成ガス中の H_2S 濃度は残油中の硫黄含有量にほとんど無関係であり 900°C では約 0.1 vol % であった (Fig. 14, Table 4)。この結果からもドロマイトは生成ガス中の H_2S 濃度を上式の平衡値まで減少させることが明らかとなった。このガス化で得られた燃料ガスを燃焼に必要な化学量論量の 105% の空気で燃焼すると、燃焼廃ガス中の SO_2 濃度はほぼ 200 ppm になる。この SO_2 濃度はミナスの常圧残油を同じ空気量 (燃焼に必要な化学量論量の 105%) で燃焼したとき生成する SO_2 濃度の 2/3 である。

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Keyword

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