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Process Estimation for Effective Development of Biomass to Liquids Process

Shinji Fujimoto*, Takashi Yanagida*, Kiyotaka Saga*, and Tomoaki Minowa*¹

Abstract – In order to accelerate the development of the biomass-to-liquids (BTL) process used to produce Fischer-Tropsch oil (liquid hydrocarbons), the economy and environmental effects of the process were simulated and estimated. The effects of the gas composition in syngas and the carbon conversion to syngas on the yield of liquid hydrocarbons were considered. The effects of heat recovery from syngas and compression on the economy of the system and the environment were investigated. The system simulation and estimation indicated that the gas composition of syngas might strongly affect the yield of liquid hydrocarbons compared to the carbon conversion to syngas. The heat recovery from syngas improved the economy of the system because of the lower energy cost of the heat exchanger. A BTL process that incorporates pressurized gasification and does not require compression might have better economy because of its lower electricity consumption compared with a BTL process that requires compression. In this case, the cost of the pressurized gasifier might be an important factor. That is, the cost should be less than 1.676 times the cost of the atmosphere gasifier. It was found that the process that does not involve compression could mitigate CO₂ emission.

Keywords – Biomass to liquids (BTL), Fischer-Tropsch (FT) synthesis, process simulation, process estimation, CO₂ mitigation.

1. INTRODUCTION

Climate change has become a serious global issue and the active utilization of biofuel is an attractive solution for the mitigation of climate change. In particular, there have been significant developments in liquid biofuel production for the mitigation of green house gas emissions from automobiles. Bioethanol (e.g., [1]), biodiesel fuel (BDF) (e.g., [2]), methanol (e.g., [3–5]), dimethyl ether (DME) (e.g., [6]), and Fischer-Tropsch (FT) oil (e.g., [7]) have been investigated as liquid biofuels. FT oil (liquid hydrocarbons) can be synthesized using a Biomass-to-Liquids (BTL) process. The BTL process involves pyrolysis or gasification and liquid-fuel synthesis. That is, biomass is converted to syngas by pyrolysis or gasification and then the converted syngas is synthesized to liquid fuel by catalytic liquid-fuel synthesis. Because various kinds of biomass can be converted to syngas by pyrolysis or gasification, the limitation of feedstock is not a problem in the BTL process as second generation biomass can be used as feedstock. FT oil, which is a BTL fuel, can be directly used as fuel for diesel engines. Moreover, FT oil has a high cetane number and good ignition. Its output power is also high. The use of FT oil mitigates emissions of NO_x and particulate matter, which cause air pollution. Therefore, FT oil has high potential for use as biofuel.

With this background, we studied the BTL process comprising steam gasification and FT synthesis in a previous study [8]. In that study, the heat balance,

carbon balance, and yield of FT oil were estimated by conducting simulated case studies. The effects of the operation condition of CO₂ mitigation were also considered. It was found that the standalone use of biomass without fossil fuel might not effectively reduce CO₂ emission. If an external electricity source has a lower CO₂ emission factor and is more efficient than biomass power generation, then the use of the external electricity source is recommended. Although the recycling of the offgas discharged from the FT reactor into the gasifier could increase the yield of liquid hydrocarbons, it could not reduce CO₂ emission because of the increase in energy consumption in the process. These results can be used as guidelines for the development of the BTL process.

However, further studies are required to realize the BTL process. For example, the theoretical H₂/CO ratio of syngas is approximately 1.0 for steam gasification. However, the stoichiometric H₂/CO ratio for FT synthesis is approximately 2.0. In order to increase the H₂/CO ratio in the BTL process, a water-gas shift reaction must occur. However, in this case, some of the CO that could be converted to FT oil is converted to CO₂ by the shift reaction. The syngas composition might also be an important factor in liquid-fuel synthesis.

Because pyrolysis and gasification are endothermic reactions, heat recovery from syngas and efficient utilization of the recovered heat are important issues. The heat recovery condition might affect the cost of the heat exchanger and thus the economy of the BTL process.

Generally, because liquid-fuel synthesis is carried out under pressurization in the BTL process, compression is required after pyrolysis or gasification under atmospheric pressure. Even if pyrolysis or gasification is carried out under pressurization, the

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pressure of syngas might be reduced by wet gas cleaning such as the use of a scrubber, and compression would be required. The energy consumption in the compression of syngas is relatively high.

Study of the above issues is important to develop guidelines for research and development. Therefore, in this study, the effects of the heat recovery from syngas on the energy balance and the economy of the BTL process were considered by process simulation in order to develop clear guidelines for research and development of the BTL process that comprises steam gasification and FT synthesis. The effect of compression was also studied. Additionally, the effect of syngas composition on the yield of FT oil was considered.

2. PROCESS DESIGN AND SIMULATION

Figure 1 shows the flow of the BTL process. In this study, atmospheric gasification and pressurized gasification were considered. The process involving atmospheric

gasification (AG-FT process) comprises steam gasification, wet gas cleaning, compression, FT synthesis, hydrogenolysis, and distillation, as described in our previous study [1]. The process involving pressurized gasification (PG-FT process) comprises steam gasification, dry gas cleaning, FT synthesis, hydrogenolysis, and distillation. In practice, if the dry gas cleaning cannot achieve sufficient refinement for FT synthesis, subsequent wet gas cleaning is employed. However, in this study, it was assumed that dry gas cleaning could achieve sufficient refinement because extreme cases need to be compared to clarify positive and negative points of the two gasification methods. Therefore, compression was not included in the PG-FT process. The preconditions of the BTL process are almost the same as those in our previous study [8], which details each process. The main preconditions are as follows.

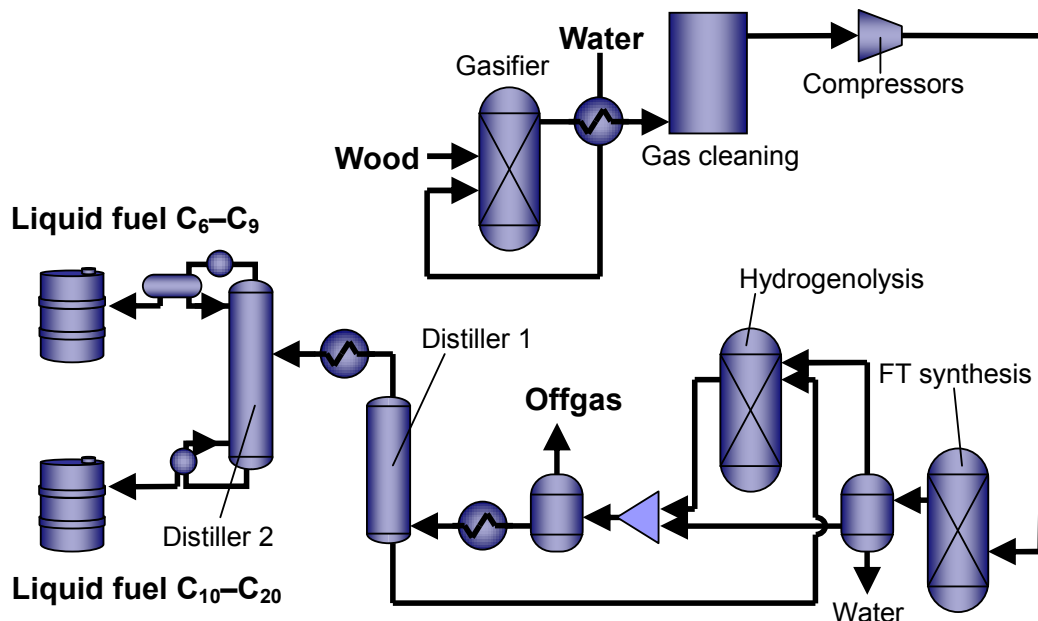


Fig. 1. Schematic representation of BTL process.

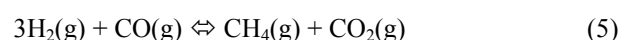
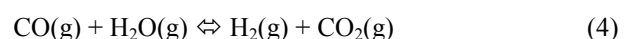
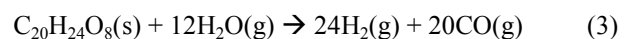
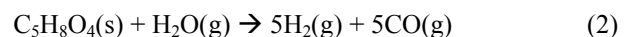
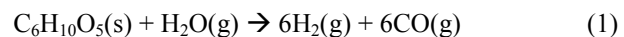
2.1 Properties of Biomass

In this study, woody biomass was considered as the feedstock biomass. Molecular weights of cellulose, hemicellulose, and lignin were assumed to be those of $C_6H_{10}O_5$, $C_5H_8O_4$, and $C_{20}H_{24}O_8$, respectively. The composition of woody biomass was assumed to be 50wt% cellulose, 20wt% hemicellulose, 27wt% lignin, and 3wt% ash. Consequently, the chemical formula of woody biomass was $C_{3.8}H_{5.3}O_{2.2}$. The moisture content of woody biomass was 20wt% and the daily amount of biomass was 100 t on a wet basis.

2.2 Gasifier

Although steam, air (oxygen), and CO_2 can be used as gasifying agents, only steam was used in this study. Atmospheric gasification and pressurized gasification

were considered. Both types of gasification were assumed to progress according to the following reactions.



The carbon conversion to syngas was assumed as 95% in the basic case. It was assumed that syngas contained 5% CH_4 . The composition of H_2 , CO , and CO_2 in syngas was calculated as 80% of the chemical equilibrium composition. It was assumed that 2.5% char and 2.5% tar were generated [8].

2.3 Heat Recovery

Because the syngas has a high temperature, heat can be recovered by a heat exchanger. In this study, the heat was used to preheat the gasifying agent and feedstock. In the AG-FT process, heat recovery was performed at a fixed minimum internal temperature approach of the heat exchanger. On the other hand, in the PG-FT process, heat recovery was performed until the temperature of the syngas reached 260°C.

2.4 Gas Cleaning

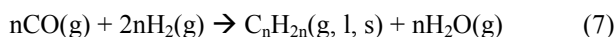
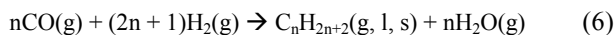
In order to separate tar from syngas, it was assumed that a scrubber was used for wet gas cleaning in the AG-FT process. In the PG-FT process, it was assumed that adsorption by activated carbon was employed for dry gas cleaning. Hanaoka *et al.* studied dry gas cleaning by activated carbon [9]. They found that the performance of refining syngas might not be sufficient for FT synthesis. However, in this study, to make an extreme comparison, it was assumed that sufficient refinement could be achieved by the dry gas cleaning. That is, in the PG-FT process, only dry gas cleaning was employed.

2.5 Compression

In the AG-FT process, compression of the syngas is required. Considering the results of our previous study [8], the syngas pressure was assumed to reach 5 MPa. When a compressor was used to compress the syngas to 5 MPa, the efficiency of compression worsened because of an increase in syngas temperature. Therefore, the syngas was assumed to be gradually compressed using three compressors. Two heat exchangers were placed between the compressors to cool the syngas, whose temperature was increased by compression, to 65°C. The efficiency of the compressors was assumed to be 85% according to reference [12] and our experience [8]. The outlet pressures of the first and second compressors were controlled to minimize the electricity consumption of the three compressors.

2.6 FT Synthesis

FT synthesis reactions progress as follows.



Olefin and paraffin are generated in the actual reaction. However, in this study, considering previous works [8, 10], only paraffin was assumed to be generated. It was assumed that the distribution of products was a Schulz-Flory distribution. The value of the Schulz-Flory distribution function, α , was assumed to be 0.9. The maximum carbon number of paraffin was 36. Eighty percent conversion of CO to hydrocarbon was assumed.

2.7 Distillation and Hydrogenolysis

The generated hydrocarbons (paraffin) were separated from offgas and water by gas-liquid phase separation and two-liquid-phase separation. The separated hydrocarbons were classified into those with carbon numbers 6–9 and those with carbon numbers 10–20 by

two distillation columns. The separated hydrocarbons were heated to 350°C and then fed to the first distillation column. From the bottom of the first column, heavy hydrocarbons with more than 21 carbon atoms were discharged. These were fed to a hydrogenolysis reactor. In the hydrogenolysis reactor, heavy hydrocarbons were cracked by hydrogen in offgas. The cracked hydrocarbons were returned to the first distillation column.

The hydrocarbons with a carbon number less than 20 were discharged from the top of the first distillation column and fed to the second distillation column. The hydrocarbons with carbon numbers 6–9 were discharged from the top of the second distillation column and those with carbon numbers 10–20 were obtained from the bottom.

2.8 Offgas

The offgas separated from the hydrocarbons was used for the hydrogenolysis of heavy hydrocarbons. The offgas was then combusted and used as a heat source for the gasifier.

2.9 Process Simulation

A steady-state process simulator (PRO/II, Invensys Systems Japan Inc.) was used for the process design and simulation. The energy balance and material balance could be obtained from the simulation results, including the degree of heat and electricity required for the process. In this simulation, heat loss was not considered.

First, the heat recovery after the gasifier was considered. Next, the effect of compression on energy consumption in the BTL process was considered. That is, the AG-FT process and PG-FT process were compared. Finally, the effect of the composition of syngas on the yield of liquid hydrocarbons was studied.

3. RESULTS AND DISCUSSION

3.1 Effect of H₂/CO Ratio of Syngas on Yield of Liquid Hydrocarbons

Figure 2 shows the change in the yield of liquid hydrocarbons with a changing H₂/CO ratio. Although the CO conversion to liquid hydrocarbons might change with the changing H₂/CO ratio in an actual experiment, it was fixed at 80% in the simulation. However, in the section where the CO conversion to liquid hydrocarbons could not reach 80% because of the lower H₂/CO ratio, CO conversion to liquid hydrocarbons was assumed to be 80% for existing H₂. In the figure, the yield of liquid hydrocarbons increased with an increasing H₂/CO ratio until the H₂/CO ratio reached 1.74. Progress of a shift reaction is required in order to increase the H₂/CO ratio although this reaction consumes CO that would be used in FT synthesis. That is, the increase in the H₂ content accompanies a decrease in the CO content. When the H₂/CO ratio is less than 1.74, the decrease in the CO content increases the yield of liquid hydrocarbons because of the presence of unreacted CO. On the other hand, for a H₂/CO ratio greater than 1.74, there is a shortage of CO and H₂ is oversupplied. Therefore, the

decrease in CO content decreases the yield of liquid hydrocarbons. A maximum yield of liquid hydrocarbons of 17.93 t/d was obtained for a H₂/CO ratio of 1.74. The yield was 12.18 t/d when the H₂/CO ratio was 2.8; this was 67.95% of the maximum yield. For a H₂/CO ratio of 1.1, the yield was 14.55 t/d; this value was 81.14% of the maximum yield. It is indicated that the H₂/CO ratio is a sensible factor that affects the yield of liquid

hydrocarbons. Although the CO conversion to liquid hydrocarbons was assumed to be a constant value of 80% in the simulation, it might vary with the H₂/CO ratio in actual operation. In the development of a catalyst, a catalyst that has high activity at a H₂/CO ratio of 1.74–2.2 should be studied and developed. Such development is an important part of establishing an efficient BTL process.

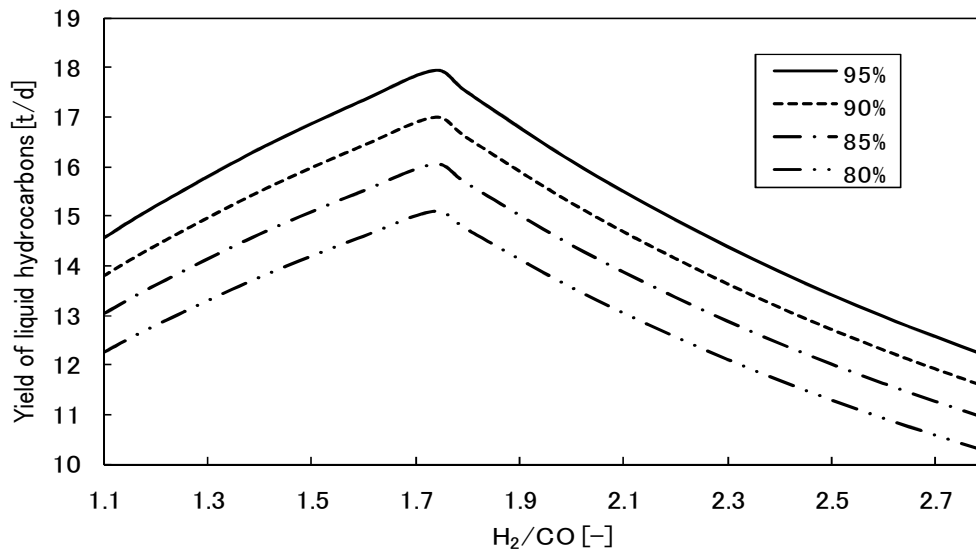


Fig. 2. Effects of H₂/CO ratio and carbon conversion to syngas on yield of liquid hydrocarbons.

3.2 Effect of Carbon Conversion to Syngas on Yield of Liquid Hydrocarbons

Figure 2 shows the effect of the carbon conversion to syngas on the yield of liquid hydrocarbons. When the rate of carbon conversion to syngas was 80%, the yield of liquid hydrocarbons was 15.10 t/d. In the case that the carbon conversion was 95%, the yield of liquid hydrocarbons excluding the yield for H₂/CO ratios of 1.185–2.171 was less than the maximum yield obtained in the case of 80% carbon conversion. In the cases of carbon conversions of 90% and 85%, the yields of liquid hydrocarbons were less than the maximum yield obtained for 80% carbon conversion excluding yields for H₂/CO ratios of 1.329–2.029 and 1.508–1.889, respectively. The results indicate that the yield of liquid hydrocarbons did not increase if the gas composition of syngas was inappropriate in spite of high carbon conversion. Although investigations into methods to increase the carbon conversion to syngas are naturally essential, technologies to optimize the gas composition for FT synthesis are also important.

3.3 Effect of Heat Recovery from Syngas on Economy of BTL Process

In the simulation, a heat exchanger was placed after the gasifier, and heat was recovered by using H₂O as a gasifying agent and syngas. The amount of H₂O supplied to the gasifier as the gasifying agent was

controlled in order to adjust the H₂/CO ratio to 1.74. The properties of heat exchange were examined for different minimum internal temperature approaches of the heat exchanger. Here, the minimum internal temperature is the temperature most similar to the syngas temperature and H₂O temperature in the heat exchanger. The relationship between the minimum internal temperature and the outlet temperature of the heat exchanger is shown in Figure 3. In the figure, it is seen that the outlet temperature of H₂O increased to 831.8–737.8°C. On the other hand, the outlet temperature of syngas decreased to 78.23–60.79°C. The slope of the outlet temperature of syngas changed at approximately 35°C, indicating the condensation of steam in the syngas. That is, steam in the syngas condensed below approximately 62.9°C at atmospheric pressure according to the relationship between the partial pressure of steam and saturated steam pressure and the production of condensation heat, and thus, the gradient in the figure is low below 62.9°C. Here, the condensation of steam in the syngas depends on the relationship between the partial pressure of steam in the syngas and saturation steam pressure.

Figure 4 shows the relationship between the minimum internal temperature approach and amount of heat recovered. Heat recovery increased with decreasing minimum internal temperature approach. Heat recovery can reduce the gasification reaction heat. If the reaction heat is provided by heavy oil, the recovered heat can decrease the cost of purchasing heavy oil (the energy cost) included in the running cost. The decrease in cost

varies with the heavy oil cost and heat efficiency. On the other hand, heat recovery increases the equipment cost included in the fixed cost. The decrease in energy cost and the increase in equipment cost are trade-offs.

Generally, the equipment cost can be estimated using the following equation with the known base equipment cost, the known base equipment size, and the known base power scaling factor.

$$C_1 = C_0 \times (S_1/S_0)^R \tag{8}$$

In the estimation of the cost of the heat exchanger, a heat transfer area or amount of heat transfer is used as the base size. Hamelinck *et al.* estimated the heat exchanger cost on the basis of an amount of heat

transfer [11]. In their study, the base cost, the base size, and the base power scaling factor were 8.1 MEuro2002, 138.1 MWth, and 0.6, respectively. The values were used in this study for the estimation of the heat exchanger cost. Figure 4 also shows the relationship between the minimum internal temperature approach and the estimated heat exchanger cost. The exchange rates between the Euro, United States dollar (\$), and Japanese yen (JPY) are listed in Table 1. The rates are average Japanese foreign exchange rates in 2009 (excluding the weekends, national holidays, and the beginning and end of the year).

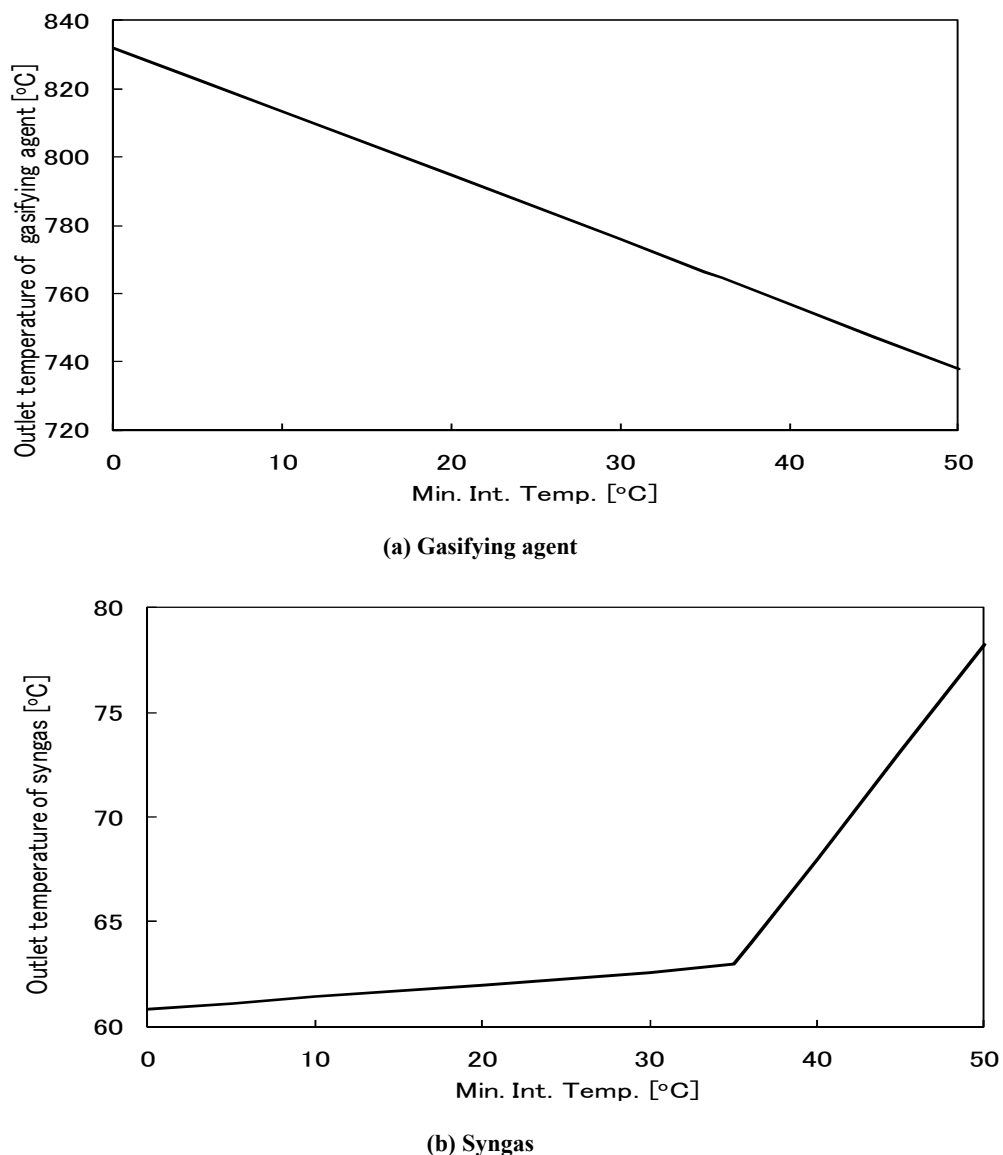


Fig. 3. Relationship between minimum internal temperature approach and outlet temperatures in heat exchanger.

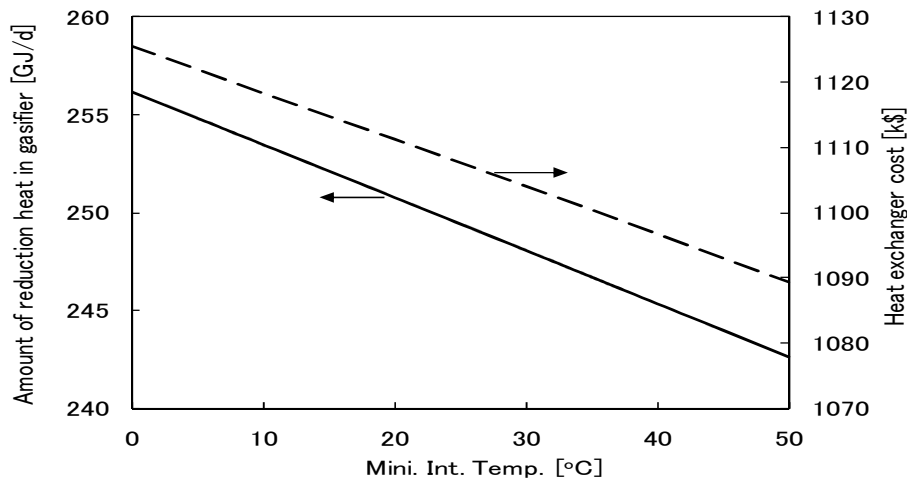


Fig. 4. Relationship between minimum internal temperature approach and amount of heat recovery and between minimum internal temperature approach and estimated cost of heat exchanger.

Table 1. Exchange rates between Euro, US dollar (\$), and Japanese yen (JPY).

	/Euro	/\$	/JPY
Euro	1	0.7182	0.007671
\$	1.392	1	0.01068
JPY	130.36	93.63	1

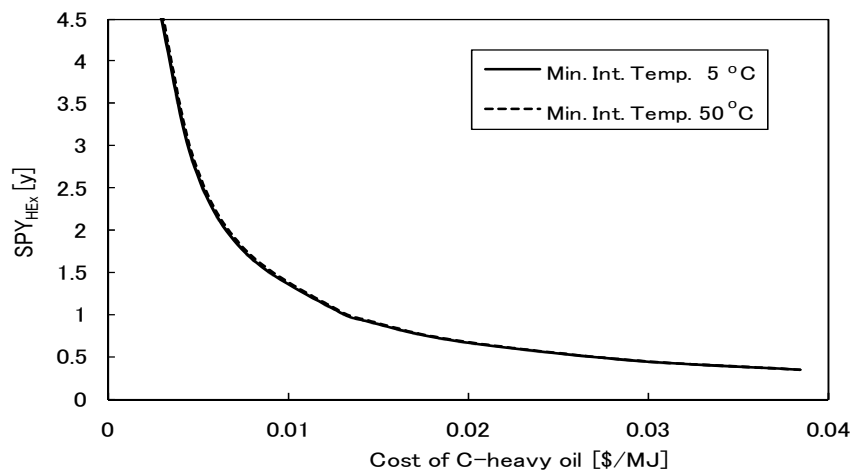


Fig. 5. Relationship between cost of C-heavy oil and relative simple payback year of heat exchanger.

The relative simple payback year of the heat exchanger (SPY_{HEX}) is defined by the following equation, which includes the decrease in energy cost due to heat recovery ($\Delta C_{Energy-HEX}$) and the heat exchanger cost (C_{HEX}).

$$SPY_{HEX} = C_{HEX} / \Delta C_{Energy-HEX} \quad (9)$$

Here, the decrease in energy cost varies with the heat efficiency and the heavy oil cost. In this study, C-heavy oil was assumed as the heat source of the gasifier. The cost of C-heavy oil (C_{Oil}) was a variable parameter. The heat efficiency was fixed at 90% with 10% heat loss. The heating value of C-heavy oil (HV_{Oil}) was 41.7 MJ/L-oil. The decrease in energy cost is defined as the

following equation.

$$\Delta C_{Energy-HEX} = H_{Ex} \times C_{Oil} / HV_{Oil} / E_{Heat} \times n_{day} \quad (10)$$

Here, H_{Ex} is the amount of heat transfer, E_{Heat} is heat efficiency, and n_{day} is the number of operation days per year. The plant was assumed to operate 24 hours a day for 300 days a year. Figure 5 shows the relationship between the cost of C-heavy oil and the relative simple payback year of the heat exchanger. The cases of minimum internal temperature approaches of 5°C and 50°C are shown in the figure. It is seen that the effect of the minimum internal temperature approach on the simple payback year was practically small. This result indicates that the increase in heat transfer

decreased the energy cost although the heat exchanger cost increased in the case of a minimum internal temperature approach of 5°C. That is, the decrease in minimum internal temperature approach does not decrease the economy of the BTL process. Additionally, because the lower minimum internal temperature approach decreased energy consumption in the gasifier, CO₂ mitigation increased.

The relative simple payback year decreased with increasing C-heavy oil cost. That is, the importance of heat recovery increased with soaring oil cost. In the case of the minimum internal temperature approach of 5°C, the relative simple payback year was less than 1 year when the C-heavy oil cost was 0.1324 \$/MJ (0.6135 \$/L).

3.4 Comparison of AG-FT and PG-FT Processes

The economies and environmental impacts of the AG-FT and PG-FT processes were considered. Table 2 lists the energy consumptions of the gasifier and compressors. In the AG-FT process, the energy consumption of the gasifier increased with increasing minimum internal temperature approach. The energy consumption of the compressors was the same for each minimum internal temperature approach because the syngas composition was the same and because the syngas was cooled to 25°C by the scrubber. On the other hand, in the PG-FT process, 3 to 4 times the energy of the AG-FT process was required by the gasifier because heat was recovered only until 260°C, which was the temperature of the FT synthesis reaction. However, the energy of compression was zero because no compressor was required. Consequently, the total energy consumptions of the AG-FT process and PG-FT process were almost the same.

Table 3 lists the energy costs for the gasifier and compressors. The cost of C-heavy oil and the electricity cost were assumed to be 50 JPY/L (0.5340 \$/L) [13] and 15 JPY/kWh (0.1602 \$/kWh) [14], respectively. The yearly energy cost of the AG-FT process using the gasifier and compressors was estimated at 1.430–1.482 1000\$/y. In the PG-FT process, the value was 0.5262 1000\$/y; *i.e.*, less than that of the AG-FT process. The advantage of the PG-FT process is indicated by the economy. Additionally, because a small heat exchanger and no compressor were required, the equipment cost was less. The relative simple payback year of introducing the PG-FT process (SPY_{PG-FT}) is defined as the following equation.

$$\text{SPY}_{\text{PG-FT}} = (C_{\text{Eq-PGFT}} - C_{\text{Eq-AGFT}}) / \Delta C_{\text{Energy-PGFT}} \quad (11)$$

Here, $C_{\text{Eq-PGFT}}$ and $C_{\text{Eq-AGFT}}$ are the equipment costs of the PG-FT and AG-FT processes, respectively. The equipment cost comprises the costs of the gasifier, heat exchanger, gas cleaning, and compressors. $\Delta C_{\text{Energy-PGFT}}$ is the decrease in cost due to the introduction of the PG-FT process. The base values of each piece of equipment in estimating the cost in Equation 8 are listed

in Table 4 [10].

When the cost of the pressurized gasifier was the same as that of the atmospheric gasifier, the relative simple payback year was estimated at –5.773 years. However, the cost of increasing the pressure resistance in the case of the pressurized gasifier would be great. The cost of the pressurized gasifier (C_{PG}) was defined as the following equation with a factor F_{R} .

$$C_{\text{PG}} = C_{\text{AG}} \times F_{\text{R}} \quad (12)$$

The cost of the pressurized gasifier was a variable parameter and the simple payback year was examined. Figure 6 shows the relationship between F_{R} and the relative simple payback year of the PG-FT process defined by Equation 11. In the figure, as a comparative case, the AG-FT process with a minimum internal temperature approach of 5°C is shown. In the figure, it is seen that for F_{R} less than 1.676 (*i.e.*, when the cost of the pressurized gasifier is less than 1.676 times the cost of the atmosphere gasifier), the relative simple payback year was negative. That is, introduction of the PG-FT process could increase the economy of the BTL process. However, when the cost of the pressurized gasifier exceeded 1.676, the simple payback year had a positive value. When the cost of the pressurized gasifier was twice the cost of the atmosphere gasifier, the relative simple payback year was less than 3 years. However, when it was three times the cost of the atmosphere gasifier, the relative simple payback year was 11.31 years. In the last case, there is not much to be gained by the introduction of the pressurized gasifier.

The effect on the environment was also considered. Here, CO₂ was used as an indicator of environmental impact. The CO₂ emission factors of C-heavy oil and electricity were taken as 2.939 kg-CO₂/L and 0.3821 kg-CO₂/kWh, respectively. The consumptions of C-heavy oil and electricity were estimated from Table 2 and then the amount of CO₂ emission was calculated using the factors of CO₂ emission. Table 5 lists the amounts of CO₂ emitted in the AG-FT process and PG-FT process. In the AG-FT process, the CO₂ emission was large in the compressors although that in the gasifier was small. The PG-FT process emitted 3 to 4 times the CO₂ amount emitted in the AG-FT process in the gasifier. However, because it had no compressor, there was no CO₂ emission from electricity. It is seen that the PG-FT process could mitigate 30% of CO₂ emitted in the AG-FT process. In a country that has a larger factor for CO₂ emission from electricity, the percentage would be greater. The PG-FT process might have potential in terms of the environment and the economy and it is expected that further study will result in effective development.

Table 2. Energy consumptions of gasifier and compressors

	AG-FT						PT-FT
	5	10	20	30	40	50	260*
Min. Int. Temp.	5	10	20	30	40	50	260*
Reactor, GJ/d	32.8	34.2	36.9	39.6	42.3	45.1	131.8
Compressor, GJ/d	96.6						-
Total, GJ/d	129.4	130.8	133.5	136.2	138.9	141.7	131.8

*: Syngas temperature at outlet of heat exchanger

Table 3. Energy costs of gasifier and compressors

	AG-FT						PT-FT
	5	10	20	30	40	50	260*
Min. Int. Temp.	5	10	20	30	40	50	260*
Reactor, k\$/y	0.140	0.146	0.157	0.169	0.181	0.192	0.563
Compressor, k\$/y	1.290						-
Total, k\$/y	1.430	1.436	1.447	1.459	1.471	1.482	0.563

*: Syngas temperature at outlet of heat exchanger

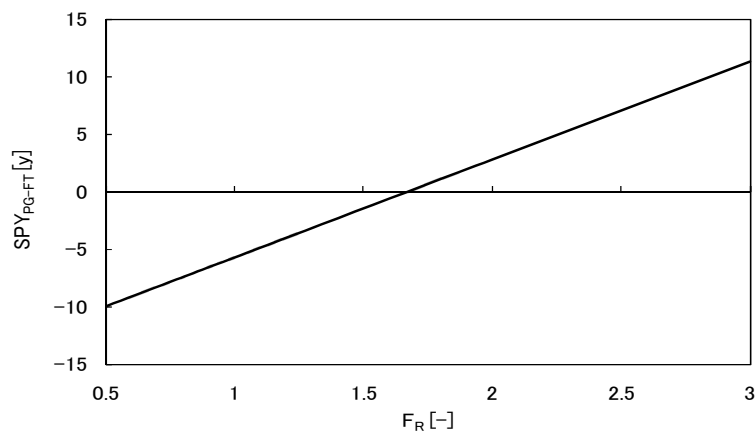
Table 4. Costs of system components in M Euro₂₀₀₂ [11]

Equipment	Base cost	Scale factor	Base scale	Maximum size
CFB gasifier	44.3	0.7	68.8 t-dry/h	75 t-dry/h
Tar cracker	3.6	0.7	34.2 m ³ /s gas	52 m ³ /s gas
Oil scrubber	1.64	0.7	14.7 m ³ _{NTP} /s gas	- m ³ _{NTP} /s gas
High-temperature heat exchanger	8.1	0.6	138.1 MW _{th}	
Scrubbers	3.0	0.7	12.1 m ³ /s gas	64 m ³ /s gas
Dry gas cleaning	35.8	1.0	74.1 m ³ /s gas	-
Guard deds	0.024	1.0	8.0 m ³ _{NTP} /s gas	-

Table 5. Amount of CO₂ emission in AG-FT and PG-FT processes

	AG-FT						PT-FT
	5	10	20	30	40	50	260*
Min. Int. Temp.	5	10	20	30	40	50	260*
CO ₂ emission in reactor, t/d	2.57	2.68	2.89	3.10	3.31	3.53	10.32
CO ₂ emission in compressor, t/d	10.25						
Total, t/d	12.82	12.93	13.14	13.36	13.57	13.78	10.32

*: Syngas temperature at outlet of heat exchanger

**Fig. 6. Relationship between FR and relative simple payback year of PG-FT process.**

4. CONCLUSIONS

In order to accelerate the development of the BTL process used to produce FT oil, the economy and environmental effects of the process was simulated and estimated. The effects of the gas composition of syngas and the carbon conversion to syngas on the yield of liquid hydrocarbons were considered. The results indicated that the optimization of the gas composition for FT synthesis is important, although an increase in carbon conversion to syngas is also essential.

The heat recovery from the syngas was considered. The equipment cost increased when a heat exchanger was included to recover the heat of syngas. However, the economy of the system improved because the energy cost of the gasifier decreased. In particular, when C-heavy oil was used as a heat source for the gasifier, it was found that the economy improved with an increase in the cost of C-heavy oil. If the minimum internal temperature approach in the heat exchanger decreased (*i.e.*, if the heat transfer in the heat exchanger increased), the size and cost of the heat exchanger increased. However, the increase of the heat exchanger size decreased the energy consumption in the gasifier. According to the trade-off, a decrease in minimum internal temperature approach does not reduce the economy. Additionally, because the lower minimum internal temperature approach reduced the energy consumption in the gasifier, CO₂ mitigation increased.

The PG-FT process was compared with the AG-FT process from the viewpoints of economy and the environment. In the PG-FT process, the energy consumption in the gasifier was greater because less heat was recovered from syngas. However, the total energy consumption was less because no compression process was required. When the cost of the pressurized gasifier was the variable parameter, the relative simple payback year decreased with decreasing cost of the pressurized gasifier. Because a compressor was not used in the PG-FT process, CO₂ attributable to the use of electricity was not emitted. Therefore, the PG-FT process could mitigate 30% of CO₂ emitted in the AG-FT process. The PG-FT process (*i.e.*, the process requiring no compression), might have environmental and economic benefits.

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