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Research Article

A KINETIC MODEL IN PYROLYSIS OF PALM EMPTY FRUIT BUNCH AND BIO-OIL UPGRADING FOR POWER GENERATION

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ABSTRACT:

In this work, a comprehensive process model for pyrolysis of palm empty fruit bunch, bio-oil upgrading, and electric power generation can predict the production under various operating conditions. This work proposed a kinetic reaction model of 149 individual reactions. Palm empty fruit bunch was used as a feedstock for pyrolysis at 500 °C. The upgraded bio-oil to gasoline and diesel via hydrotreating process was implemented by a pseudo-first order reaction of lumped bio-oil species followed by the hydrocracking. Dry Empty Fruit Bunch can produce 67.25% of bio-oil, 21.37% of gas, and 11.38% of char. The bio-oil upgraded at 1 kg of wet empty fruit bunch can produce 0.053 kg of gasoline and 0.069 kg of diesel. Electricity generation in the pyrolysis and the upgrading can save electricity up to 72%. The energy conversion efficiency of the pyrolysis combined with the upgrading step was 57.38%. Therefore, this work is a suitable option for biofuel production from empty fruit bunch.

Keywords: *Biofuel, Electric Power Generation, Empty Fruit Bunch, Kinetic model, Pyrolysis*

1. INTRODUCTION

Renewable energy is becoming interesting due to the reduction trend of fossil fuel usage and also the mitigation of the greenhouse effect [1]. Pyrolysis; the absent oxygen thermal decomposition, is the well-known technic to produce fuel from biomass. Lignocellulosic biomass (i.e., empty fruit bunch, pinewood, and eucalyptus) as a feedstock can yield a gas fuel, a carbonaceous residue (char), and a liquid fraction (bio-oil) [2]. Palm empty fruit bunch (EFB) is the primary residue from harvesting palm fruits. The government policy is about to promote the cultivation of palm oil and increase palm oil production capacity. The plan is to increase palm oil production up to 26% of total world consumption in 2020 [3]. Hence, it is an excellent opportunity to leverage this residue as the energy source.

The bio-oil can be maximized by fast pyrolysis which is operating at 500 °C and short residence times, while slow pyrolysis achieves high char yields at slightly lower temperatures around 450 °C and very long vapor residence times [4]. Biomass pyrolysis is mainly in the research stage, and almost no commercial pyrolysis installation exists to-date [5]. Scale up protocol from the pilot plant into the industrial plant can principally be conducted by process simulation, which is the tool used for the design, development, analysis, and optimization of processes [6]. Moreover, there is bio-oil upgrading process used for quality improvement of bio-oil because bio-oil produced from pyrolysis process still is an unacceptable fuel, i.e., viscous, low volatility, corrosive, exhibits high oxygen content [7].

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The bio-oil production process mainly consists of pretreatment (drying and size reduction), fast pyrolysis (fluidized-bed), gas-solid separation, quenching (indirect heat exchanger) and combustor [8]. The bio-oil upgrading process mainly consists of hydrotreater, flash drum, decanter, hydrocracker, gasoline and diesel column, pressure swing column, and steam reformer [9].

The kinetic reaction models for biomass pyrolysis processes have been researching over a year. Blasi and Branca [10] modeled the one component mechanism, which considered the decomposition of biomass into gas, bio-oil, and char directly. Miller and Bellan [11] modeled the multi-component mechanism covered the biomass as the mixture of bio-component (cellulose, hemicellulose, and lignin). For each bio-component are decomposed into active components before the composition into bio-oil, char, and gas. The detailed composition was not determined. Ranzi *et al.* [12] modeled the multi-component with detailed composition. However, the effect of secondary reaction that occurs by the catalytic effect of alkali metal in the biomass had not been considered. Anca-Couce *et al.* [13] modeled kinetic scheme of biomass pyrolysis considering secondary charring reactions. They created the pathway considering the volatile component and char obtained from secondary reactions. The recent work, Peters *et al.* [2] provided the detailed kinetic model with consideration of secondary reaction that can be applied for any lignocellulosic biomass with the satisfying result. The model included 149 reactions to predict the volatilization, decomposition, and recomposition processes of biomass pyrolysis with a linear regression algorithm accounted for the secondary pyrolysis reaction. Their work was validated by the experimental data at various type of lignocellulosic biomass and a wide range of reaction temperature and vapor residence times. Hence, the kinetic model [2] is proposed in this work.

In this study, the bio-oil production process from palm empty fruit bunches and bio-oil upgrading process was simulated and designed. The energy conversion efficiency of the fast pyrolysis process combined with the bio-oil upgrading process was evaluated.

2. MATERIAL AND METHODS

2.1 Process description

The bio-oil production plant is basically integrated into seven operations; pretreatment, pyrolysis, solid removal, bio-oil recovery, bio-oil upgrading, combustion, and power generation as depicted in Fig. 1. EFB is dried and ground prior to being fed into the fluidized-bed pyrolysis reactor where the EFB is converted into bio-oil under an oxygen-free environment at around 500 °C and ambient pressure. Then, solids (ash and char) are removed from the pyrolysis gas stream before condensed to a liquid. At the bio-oil recovery, bio-oil is condensed and stored to preserve the oil compounds and to obtain a high yield of good quality. Non-condensable pyrolysis gases and part of the char generated from pyrolysis are combusted in combustion to provide heat for pyrolysis and power generation. The bio-oil upgrading section produces standard gasoline, and diesel fuel in Fig. 2. This section performs both hydrotreating and hydrocracking processes with the supplemented hydrogen required by steam reforming. Finally, product separation gives diesel and gasoline products.

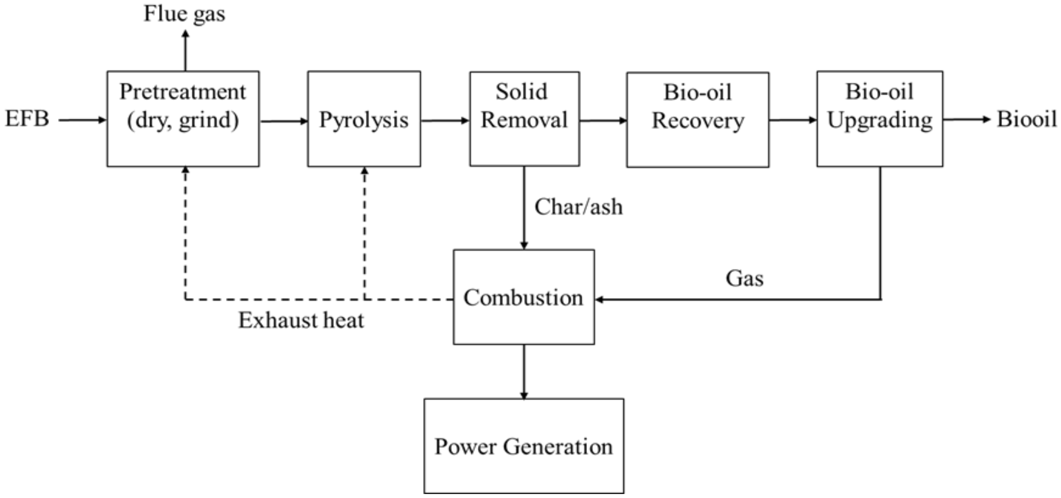

Fig. 1. Block diagram for pyrolysis of palm empty fruit bunch.

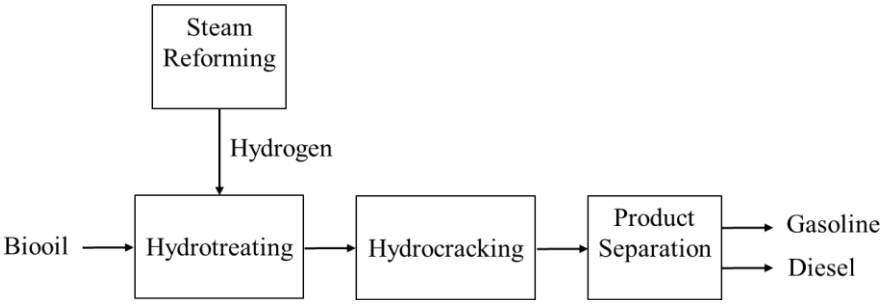

Fig. 2. Block diagram for bio-oil upgrading.

2.2 Process simulation

The pyrolysis model of EFB is implemented in Aspen Plus V8.8. The properties of EFB are listed in Table 1. The ultimate and proximate analysis for dry EFB containing moisture of 7.95 wt% and ash of 5.36 wt%. The comprehensive process flow diagrams for bio-oil production and electric power generation is shown in Fig. 3. EFB is dried at 500 °C and finely ground to particles 3 mm in diameter with 6.04% moisture. Next, EFB is sent to a fluidized bed reactor with three model blocks at a pyrolysis temperature of 500 °C and the residence time of 1 second [2]. Where the EFB is converted into bio-oil, char, and gas. After leaving the pyrolysis reactor, the gas passes a cyclone where the char is separated. After the cyclone, the hot gas is quenched down to 100 °C by direct contact with a bio-oil stream at 50 °C to avoid further pyrolysis reaction. The gas then goes into the flash at 50 °C, and the condensed vapor is recycled to mix with bio-oil. The quenched bio-oil is sent for further upgrading process. Char and gas are subsequently fed into a combustor. The heat required by the pyrolysis reactor and dryer is provided by the combustor, which combusts gas and char. The combustor is modeled as a combination of two reactors, a first decomposition reactor (RYield reactor) which decomposes the char fractions into its elements and a second RGibbs reactor that calculates heat balance and combustion products based on Gibbs energy minimization. The combustion is operating at 1269 °C, and the pressure of 10 bar [9]. The combustion products leave the combustor after passing a cyclone, which separates the ash from the flue gas. The flue gas is recycled to exchange heat for dry air. After that, the flue gas is converted into the work by a gas turbine and finally transformed into electricity.

Table 1: Ultimate and biochemical composition of empty fruit bunch [14].

Ultimate analysis	(wt%)	Proximate analysis	(wt%)	Biochemical composition	(wt%)
C	49.07	Moisture	7.95	Cellulose	59.7
H	6.48	Volatile	83.86	Hemicellulose	22.1
N	0.7	Ash	5.36	Lignin	18.1
S	<0.1	Fixed carbon	10.78		
O	38.29				

A process flow diagrams of the upgrading plant is given in Fig. 4. The hydrotreating stage converts to reduce the oxygen content of the bio-oil in a two-stage hydrotreating process over Pt/Al_2O_3 catalyst [9]. The hydrotreater is used as a pseudo-first-order kinetic model of lumped species. The lumped bio-oil is contained five pseudo-components, namely, light non-volatile, heavy non-volatile, phenolics, aromatics + alkanes, coke + water + gas. The yields are adjusted to give an oxygen content of around 2%. In the subsequent distillation process, the organic fraction is distilled into gasoline, diesel, and heavy residue. The latter is sent to a hydrocracker, which crack s the heavy residue to smaller chain hydrocarbons, increasing thereby the gasoline and diesel yields. The hydrogen required by the process is produced by steam reforming. A high-pressure steam cycle is generated heat by receiving from hydrotreater and hydrocracker in the supply of heat for the turbine, gasoline, and diesel column.

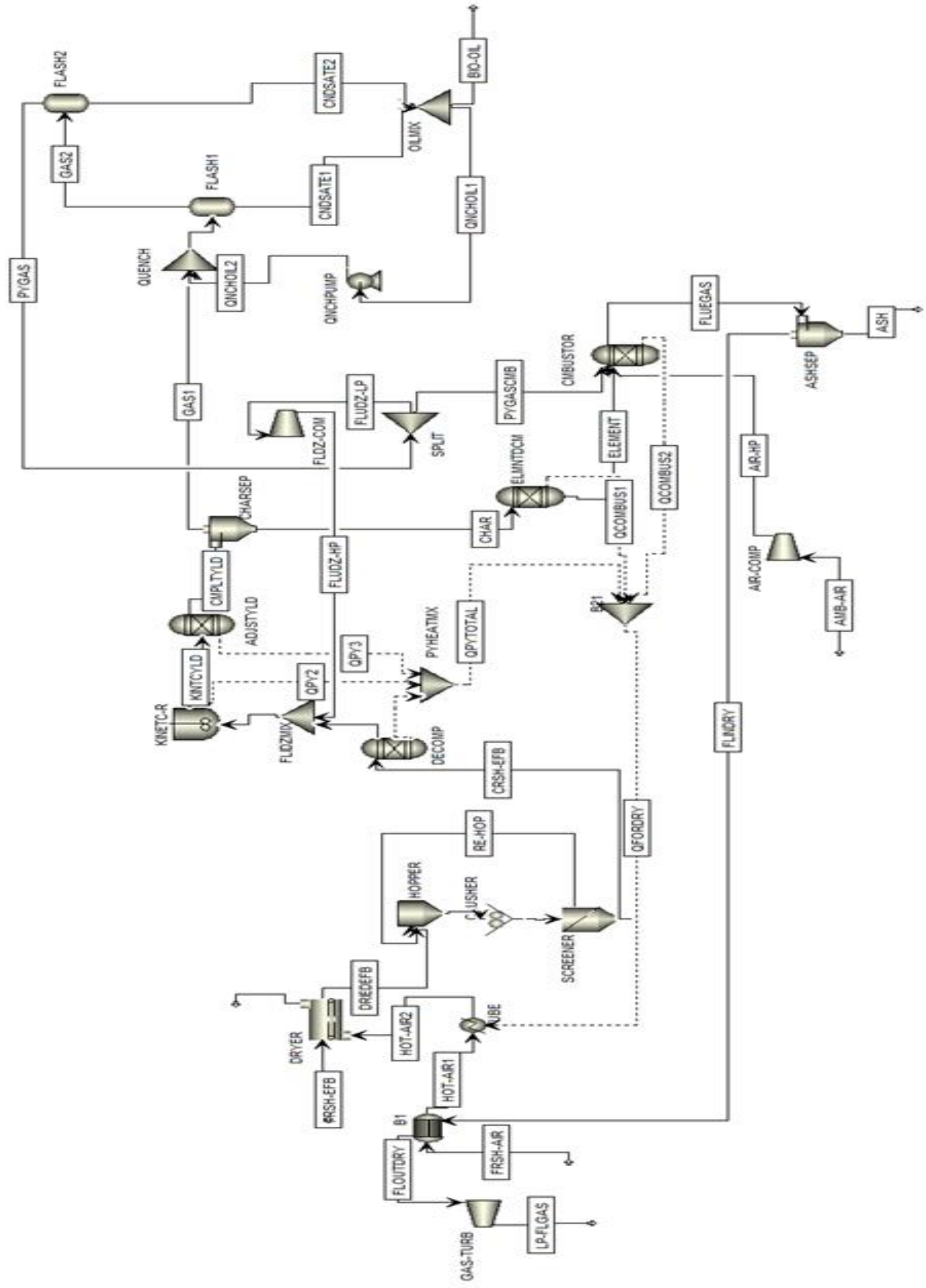

Fig. 3. Process flow diagrams for bio-oil production and electric power generation.

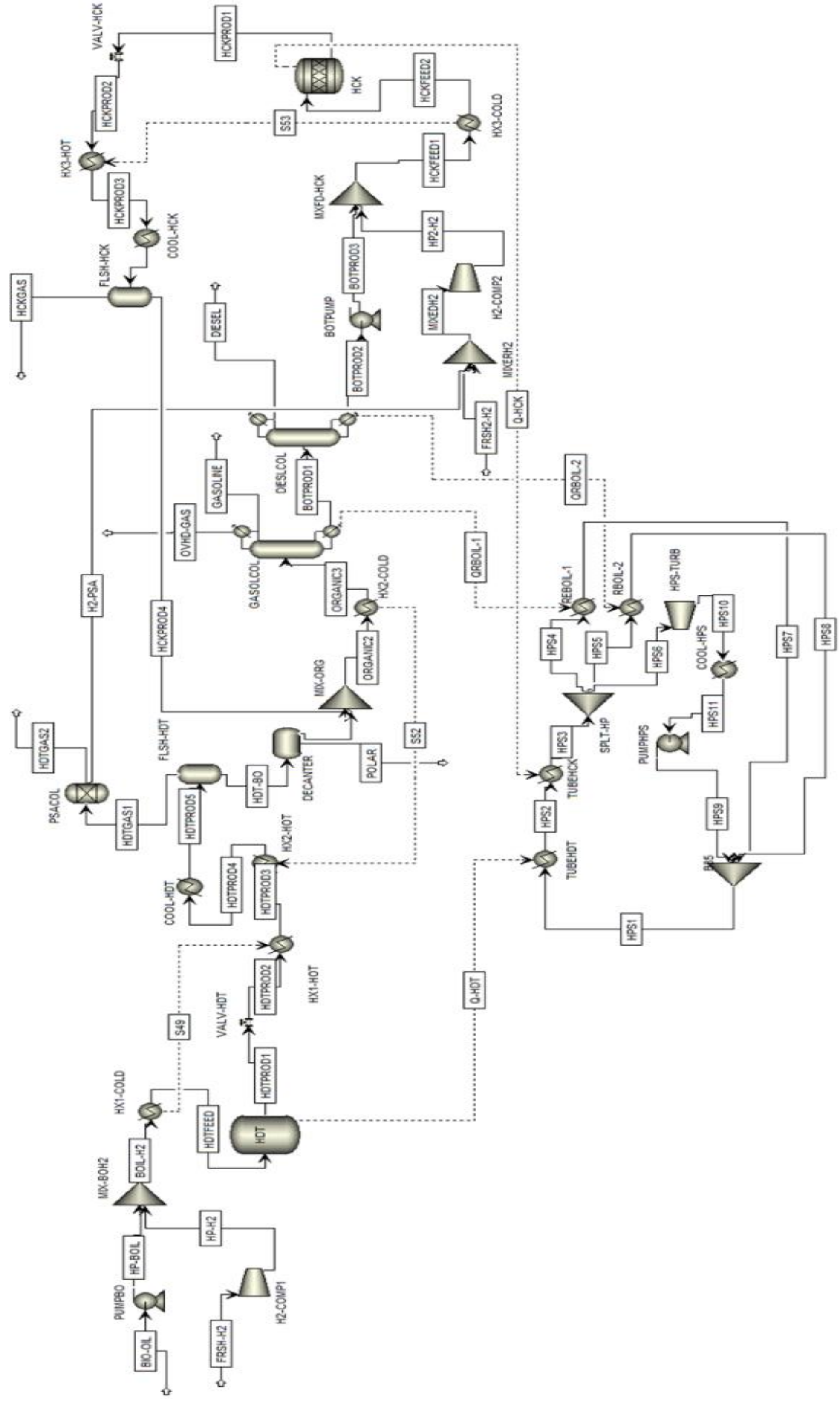

Fig. 4. Process flow diagrams for bio-oil upgrading.

2.3 Pretreatment

Fresh palm empty fruit bunch stream (FRESH-EFB) with the initial moisture content of 52% is fed into a rotary dryer (DRYER) and the air outlet temperature of 100 °C to reduce moisture content. The dried EFB is fed into a multiple roll crusher (CRUSHER) in which the particle size is reduced to 3 mm and followed by a screen (SCREENER) for particle separation. If the particle size is bigger than 3 mm, it is sent back into a hopper (HOPPER). The energy required for drying is supplied by hot air (HOT-AIR2) from combustor (DRY-COMP) at 1.2 bar, which dry air is sent to a heat exchanger (HXDRYAIR) to obtain heat from flue gas, while the dried EFB exits (DRIEDEFB) the dryer with a 6.04% moisture content. The dried biomass feed then goes into the fluidized bed reactor.

2.4 Pyrolysis

Three model blocks (DECOMP, KINETIC-R, and ADJSTYLD) are used to model a bubbling fluidized bed pyrolysis reactor. In the yield reactor (DECOMP), biomass is fragmented into its subcomponents (cellulose, hemicellulose, and lignin). The fluidized bed (KINETC-R) is used to model the reactor's fluid dynamics with a specified bed pressure drop of 0.01 mbar at a reactor temperature of 500 °C and the residence time is 1 second [2]. The gas-to-EFB ratio for bed fluidization is 1:1 [15]. The reaction rate implements 149 individual reactions are modeled by the CSTR (KINETC-R) using multistep reactions kinetics of biomass pyrolysis developed by the work [2]. The NRTL-RK thermodynamic model is selected to calculate the mass and energy balances in the quenching unit.

2.5 Combustion

The combustion unit is modeled by a yield reactor (ELMNTDCM) and a Gibbs reactor (CMBUSTOR). Unreacted biomass separated from the cyclone goes into the yield reactor (ELMNTDCM) where it is decomposed into its constituent elements before it is fed into the Gibbs reactor (CMBUSTOR) along with char (assumed to be 100% carbon in the elemental constitution) and gas. The Gibbs reactor calculates the multi-phase chemical equilibrium by minimizing Gibbs free energy, and it was modeled using the Peng-Robinson (PENG-ROB) equation of state. Although a maximum temperature of 1700 °C can be achieved at complete combustion, the fuel mixture of solids and gas are combusted in 60% theoretical air at a combustion temperature of 1269 °C in order to mitigate ash melting and prevent material failure at severe temperatures [9]. The design specification is used to determine the air flow rate for combustor to achieve 3.15% O_2 excess, which is the typical condition in industries. Ash is separated from the combustion gases by a hot cyclone (ASHSEP). The flue gas (FLINDRY) is sent into the heat exchanger (B1) to supply heat for the feed air, which goes to the dryer (DRYER).

2.6 Power generation

Flue gas (FLOUTDRY) after heating drying air**,** it is sent to a turbine (GAS-TURB) to obtain the shaft work which will be used to drive the electricity generator later. The outlet pressure of the turbine is specified equal to atmospheric pressure, and the isentropic efficiency of the turbine is set to be 0.85 following the efficiency of a typical turbine.

2.7 Hydrotreating

Bio-oil from the pyrolysis process is pressurized by a pump (PUMPBO) to 87 bar. In a mixer (MIX-BOH2), bio-oil is mixed with hydrogen (FRSH-H2) coming from the steam reforming. Next, it is heated to 360 °C by heat recovery from the hot effluent of hydrotreater (HDTPROD2) in the heat exchanger (HX1-COLD). Then, it is sent to hydrotreater (HDT) operated at 383 °C, a design specification adjusts 87 bars [9], WHSV of 0.135 h-1 to deoxygenate the stabilized bio-oil deeply up to an oxygen content less than 2% [15]. The hydrotreated bio-oil is sent to the valve to decrease pressure to 20 bar and cooled to 50 °C by heat exchangers (HX1-HOT1, HX2-HOT2, COOL-HDT), which transfer heat into feedstream of hydrotreater (BOIL-H2), inlet stream of gasoline column (ORGANIC2) and cooling water. After cooling, it is sent to the adiabatic flash drum (FLSH-HDT) at 20 bars to separate hydrogen and light hydrocarbon (HDTGAS1) out of hydrotreated bio-oil. Then, the hydrotreated bio-oil (HDT-BO) is sent to an adiabatic decanter (DECANTER) at 20 bar to separate into oil phase (ORGANIC) out of the polar phase (POLAR). In oil-phase, it is preheated to 144.5 °C by recovering heat from hydrotreated bio-oil in the heat exchanger (HX2-COLD) and sent to gasoline column (GASOLCOL). The vapor stream (HDT leaving from the flash drum (FLSH-HDT), it is sent to pressure swing adsorption (PSACOL) to recover 90% of hydrogen with purity 99.99% and recovered hydrogen (H2-PSA) for hydrocracking section.

2.8 Distillation

The stream (ORGANIC3) enters the first distillation column (GASOLCOL), which separates the gasoline fraction from the heavier fraction (BOTPROD1). The gasoline column has nine stages of partial condenser column with a reflux ratio of 1.2 and distillate to feed ratio of 0.454. The heavy fraction (BOTPROD1) enters the second distillation column (DIESLCOL). The diesel column has seven stages with a reflux ratio of 1.2 and distillate to feed ratio of 0.884. The heavy residue from the diesel column (BOTPROD2) is pressurized to 90 bar by the pump (BOTPUMP) and mixed with hydrogen (HP2-H2) at 90 bar coming from the compressor (H2-COMP2). After that, the stream (HCKFEED1) is preheated from 308 °C for reaction condition of hydrocracker (550 °C) by receiving heat from the hot effluent of hydrocracker (HCKPROD2) in the heat exchanger (HX3-COLD/HOT).

2.9 Hydrocracking

The heavy stream coming from the diesel column is pressurized to 90 bar, the operating pressure of the hydrocracker (HCK), then mixed with H_2 at 90 bar. The hydrocracker is operated at 677 °C and 90 bar [2] and modeled by Rstoic reactor, which contains eight reactions [16]. After hydrocracking, the product (HCKPROD1) is sent to the valve (VALV-HCK) to decrease pressure to 20 bar. Then, it is cooled to 300 °C by the heat exchanger (HX3-COLD/HOT), which preheat the inlet stream of hydrocracker. After that, it is cooled to 50 °C by cooling water in the heat exchanger (COOL-HCK). Before this cooled hydrocracked product (HCKPROD4) is recycled to distillation, it is sent to flash drum (FLSH-HCK) for separating light gas to save the duty in the heat exchanger (HX2-COLD). The liquid stream (HCKPROD4) coming from the flash drum is sent to combine with oil phase stream (ORGANIC) coming from the decanter to redistill in gasoline and diesel column again.

2.10 High-pressure steam

The high-pressure steam (HPS3) at 285 °C, 70 bar and 8601.38 kg/hr is generated by receiving heat from hydrotreater and hydrocracker. The generated high-pressure steam is supplied heat to turbine, gasoline, and diesel column.

3. RESULTS AND DISCUSSION

The kinetic reaction model of fast pyrolysis is verified with simulation work [2] on a fluidized bed pyrolysis reactor using Beechwood feedstock. Peters *et al.* [2] simulate the pyrolysis model of Beechwood at different reactor temperature (470, 520, and 550 °C). The result of pyrolysis yields (char, gas, and oil) as a function of the reaction temperature is shown in Fig. 5. It observed that the pyrolysis reactor model result is a good agreement with simulation result [2], which is the highest bio-oil yield is obtained at 520 °C.

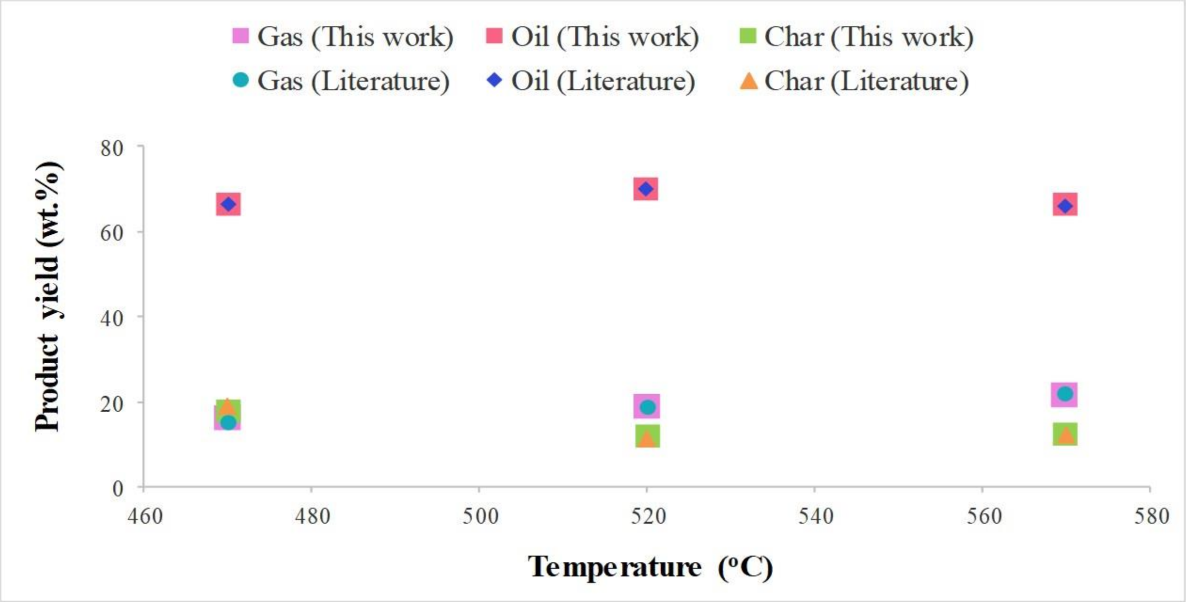

Fig. 5. Verification of simulation results with experiment [2].

The fast pyrolysis model developed in this study is validated with experimental work [14] on a fluidized bed pyrolysis reactor using EFB feedstock. The comparison between fast pyrolysis reactor model results and experimental measurements of pyrolysis products (Bio-oil, gas, and char) are presented in Table 2. They operated at 500 °C and vapor residence time equal to 1 second. This work has gas, oil, and char were 14.70, 72.93, and 11.73%, respectively. It is observed that the pyrolysis reaction model results agree considerably with experimental data [14].

Table 2: Validation (Fast pyrolysis, 500 °C) of simulation with experiment [14].

%Yield	**EFB**	
	This work	**Experiment [14]**
Gas	14.70	14.70
Bio-oil	72.93	72.36
Char	11.73	10.76

The temperature in the bio-oil separation unit is compared with experimental work [14], pyrolysis reactor operated at 500 °C, and vapor residence time equal to 1 second. Abdullah *et al.* [14] are operated temperature of the flash drum at 25 °C, and this work is used 50 °C. The percentage of gas, bio-oil, and char at the temperature difference of flash drum in a separation unit is shown in Table 3. The result has a significant effect on product yields. This work has gas, oil, and char are 21.37, 67.25, and 11.38, respectively, which obtains lower bio-oil and higher gas than experimental work [14]. However, flash drum operating at 50 °C because cooling water is the most commonly used cold utility in the temperature range 120 °C to 40 °C [17].

Table 3: The effect of temperature separation unit on product yields.

%Yield	**EFB**	
	This work	**Experiment [14]**
Gas	21.37	14.70
Bio-oil	67.25	72.36
Char	11.38	10.76

The electricity generation and consumption in the pyrolysis process are reported in Table 4. The electricity generation of the turbine is 9.85 MW. The electricity consumption of crusher, pump, and compressors are 1.21, 0.002, and 11.67 MW, respectively. The net electricity consumption and generation are 9.85 and 12.90 MW, respectively. Thus, this pyrolysis process is needed external electricity for the operation around 3.05 MW.

Table 4: Electricity generation and consumption in the pyrolysis process.

Electricity generation	**MW**	**Electricity consumption**	**MW**
Turbine	9.85	Crusher	1.21
		Pump (Quench oil)	0.002
		Compressor (Fluidized gas)	0.24
		Compressor (Air + Pyrolysis)	9.84
		Compressor (Dry air)	1.59
Net	9.85		12.90

Lumped fractions of hydrotreated oil from pinewood in Table 5 are validated with the experiment [7] at the temperature of 400 °C, the pressure of 87.2 bar and WHSV of 2 h^{-1} over Pt/Al_2O_3 catalyst. The hydrotreated oil in this work has heavy nonvolatile, light nonvolatile, phenolics, aromatic/alkane, coke + water + gas are 23.92, 29.54, 10.26, 19.79, and 15.62, respectively. The hydrotreating model results are an inadequate agreement with experimental data [7].

Table 5: The simulation results of hydrotreating model validated with the experiment [7].

Lumped components	This work	Experiment [7]	Error (%)
Heavy nonvolatile	23.92	24.57	2.64
Light nonvolatile	29.54	29.41	0.44
Phenolics	10.26	10.63	3.48
Aromatic/Alkane	19.79	19.52	1.38
Coke + Water + Gas	15.62	15.87	1.57

The hydrotreating reactor model of pyrolysis oil from EFB is operating at the temperature of 383 °C, the pressure of 87 bar [1] over Pt/Al_2O_3 catalyst [9]. A design specification adjusts WHSV of 0.135 h^{-1} to the oxygen content less than 2%. [15]. A stoichiometric reactor simulates the hydrocracker model with eight stoichiometric cracking reactions [16]. The yields of the hydrotreating reactor are shown in Table 6, which have heavy nonvolatile, light nonvolatile, phenolics, aromatic/alkane, coke + water + gas, and hydrogen conversion are 0.0001, 1.11, 1.19, 36.92, 60.77, and 38.46, respectively. The hydrotreated bio-oil has the oxygen content less than 2% and hydrogen conversion of 38.46 %.

Table 6: Lumped fractions of hydrotreated oil from EFB

Lumped fractions	%wt
Heavy nonvolatile	0.0001
Light nonvolatile	1.11
Phenolics	1.19
Aromatic/Alkane	36.92
Coke + Water + Gas	60.77
Hydrogen conversion = 38.46 %	

The electricity consumption and generation of bio-oil upgrading process are shown in Table 7. The electricity generation of the turbine is 0.27 MW. The electricity consumption of the compressor and pump are 0.25 and 0.8354, respectively. The net electricity consumption and generation are 0.27 and 1.09 MW, respectively. Thus, the bio-oil upgrading process is needed external electricity for the operation around 0.82 MW.

Table 7: Electricity consumption and generation in the bio-oil upgrading process.

Electricity generation	MW	Electricity consumption	MW
Turbine	0.27	Comp1	0.13
		Comp2	0.12
		PumpBO	0.83
		PumpBOT	0.0016
		PumpST	0.0038
Net	0.27	Net	1.09

The energy balance of the overall process is shown in Table 8. The overall process presents 0.053 kg of gasoline and 0.069 kg of diesel can be produced from 1 kg of wet empty fruit bunch. The energy efficiency of EFB and hydrogen are 189.63, and 5.41 MW, respectively. The electricity consumption is about 3.87 MW. The net electricity consumption of gasoline and diesel are 50.95, and 63.18 MW, respectively. The energy efficiency of the whole process is 57.38%. This is a value higher than gasification and bio-ethanol processes [18], Fischer-Tropsch-synthesis process [19], which have an energy conversion efficiency of 52.64, 47.32, and 50.50% respectively. Therefore, this whole process is a suitable option for biofuel production.

Table 8: Energy balance of overall process (Higher heating value basis).

Inputs	MW	Products	MW
Empty fruit bunch	189.63	Gasoline	50.95
Hydrogen	5.41	Diesel	63.18
Electricity	3.87		
Net	198.91	Net	114.13

The higher heating value of hydrogen, gasoline, and diesel are obtained from Aspen Plus. The higher heating value of empty fruit bunch as solid nonconventional compound can be calculated based on its atomic composition according to equation 1 [20].

$$\mathrm{HHV} = [152.19 \times \mathrm{H} + 98.767] \times [(\mathrm{C}/3) + \mathrm{H} - (\mathrm{O} - \mathrm{S})/8] \tag{1}$$

HHV is the higher heating value. The C, H, O, and S are the mass fractions of carbon, hydrogen, oxygen, and sulfur in free ash empty fruit bunch.

The chemical properties of gasoline and diesel from simulation results can be shown in Table 9. The gasoline and diesel densities are 749.05 and 806.46 kg/m^3, respectively. The average molecular weight of gasoline and diesel are 81.92 and 165.36, respectively. The HHV of gasoline and diesel are 44.61 and 42.07 MJ/kg, respectively. The distillation curve of gasoline has distilled liquid 25 vol% at 70 °C, 46-71 vol% at 100 °C and the final boiling point at 250 °C. The design specification adjusts the distillate to feed ratio as 0.454 to achieve the gasoline requirements. The distillation curve of diesel has distilled liquid <65 vol% at 250 °C, >85 vol% at 350 °C. The distillate to feed ratio is adjusted to 0.884 by design spec tool to achieve diesel specification.

Table 9: Chemical properties of gasoline and diesel from the simulation

Fuel	Gasoline (C_5-C_{12})	Diesel (C_8-C_{18})
Paraffines and Naphthenes	**78.16**	**76.20**
Alkanes (wt%)	23.30	48.93
Naphthenes (wt%)	54.86	11.90
Polynapthenes (wt%)	0.00	15.37
Aromatics	**10.46**	**20.65**
Monoaromatics (wt%)	10.46	4.82
Diaromatics (wt%)	0.00	2.70
Polyaromatics (wt%)	0.00	13.13
Oxygenates	**7.87**	**3.16**
Furans (wt%)	6.00	0.30
Alcohols (wt%)	1.81	0.97
Phenol (wt%)	0.06	1.89
Water	1.26	0.00
Impurities (Mainly CO_2)	2.25	0.00
Density (kg/m^3)	749.05	806.46
Average molecular weight	81.92	165.36
HHV (MJ/kg)	44.61	42.07
Distillation	**True boiling point**	**True boiling point**
5 vol%	33 °C	150 °C
10 vol%	44 °C	152 °C
30 vol%	76 °C	169 °C
50 vol%	81 °C	231 °C
70 vol%	98 °C	265 °C
90 vol%	125 °C	353 °C
95 vol%	138 °C	434 °C
100 vol%	157 °C	495 °C

4. CONCLUSION

The biofuel production from empty fruit bunch via fast pyrolysis and upgrading process is implemented in Aspen Plus. It is observed from simulation results that dry, empty fruit bunch generates 67.25% of bio-oil, 21.37% of gas, and 11.38% of char by pyrolysis condition with residence time 1 second at 500 °C. The bio-oil from fast pyrolysis process is upgraded into gasoline and diesel via hydrotreating and hydrocracking process. The upgraded oil still remains less than two wt% of oxygen. In the bio-oil upgrading process, the simulation results indicated that 0.053 kg of gasoline and 0.069 kg of diesel could be produced from 1 kg of wet empty fruit bunch. The gasoline and diesel fuel both are products that have a higher heating value of 44.61 and 42.07 MJ/kg, respectively. The whole process achieves the energy conversion efficiency of 57.38%.

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