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Integrated catalytic pyrolysis and catalytic upgrading of *Leucaena leucocephala* over natural catalysts

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ABSTRACT

Catalysts are used to improve the properties of the bio-oil produced through biomass pyrolysis. In this study, natural minerals, such as natural zeolite and dolomite, were used for the catalytic upgrading of the pyrolysis of *Leucaena leucocephala*. Pyrolysis was performed using an in-house designed fluidized bed system wherein the catalyst beds can be placed in two locations. Dolomite and zeolite catalyst beds were placed in the primary fluidized bed and/or secondary hot filter reactors. Four catalyst configurations were utilized in this study. The effects of the type and location of the catalyst on the catalytic activity and product selectivity were investigated. The highest light bio-oil fraction was obtained using the dolomite catalyst situated at the primary and secondary reactors. On the other hand, the configuration in which the natural zeolite catalyst was placed in the two reactors produced heavy bio-oil that was mainly composed of hydrocarbons. The viscosity, flash point, and fly point of the bio-oil produced using the catalysts was lower by 60% of that of the product obtained through noncatalytic biomass pyrolysis. The pyrolysis system employed in this study is a promising technology for catalytic biomass pyrolysis processes.

1. Introduction

Alternative energy resources have been sought after to limit the dependence on fossil fuels for energy-related applications to minimize the adverse environmental effects, which include air pollution due to carbon emissions, associated with the use of these conventional resources. Biomass is considered as a promising alternative energy resource due its renewability, availability, low cost, and carbon neutrality [1–5]. Moreover, the biomass had low heteroatom content, including sulfur and nitrogen within 50 ppm [6,7]. Therefore, the bio-oil can be a good candidate to reduce the pollution emission. It can be converted into value-added fuels through biochemical and thermochemical processes [8,9]. Among the thermochemical processes, pyrolysis is an efficient and cost-effective method to produce bio-oil.

However, the produced bio-oil is usually of low-quality and is highly corrosive due to the high concentration of oxygenated compounds, such as aldehydes, ketones, and carboxylic acids, in it [10]. Therefore, the pyrolysis process should be refined to improve the properties and composition of the bio-oil.

Catalysts are used for the upgrading of the bio-oil produced through pyrolysis. In particular, they are employed to reduce the oxygen content in the desired product [11,12]. Catalytic pyrolysis can generally be divided into two configurations based on the location of the catalysts [13,14]. In the first configuration, the catalyst–biomass mixture is put in the same reactor in which the volatile product is produced. In this setup, the product has direct contact with the catalyst, which increases the rate of the reaction. However, the formation of pyrolysis coke is also facilitated in this configuration, which can deactivate the catalyst rapidly.

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The improvement in properties of the bio-oil occurs in situ with pyrolysis, meanwhile, the temperature of catalytic reaction was dependent on pyrolysis temperature may have promoted undesired reactions. In the second configuration, the catalyst is placed downstream of the pyrolysis reactor, hence only the pyrolyzed product would contact the catalyst for upgrading. The catalytic reaction conditions are not necessarily the same as those of pyrolysis. Moreover, in the second setup, the catalyst degrades slowly as there are only light compounds present in the feed stream. For both configurations, the catalyst is critical for the control of the reaction activity and the properties of the desired products.

Zeolite is a widely used catalyst to promote the cracking reaction during pyrolysis [15–17]. Faujasite zeolites were used for the catalytic pyrolysis of white pine wood chip. The H⁺ and Na⁺ ions in its framework played key functions to control the product yield and selectivity. In addition, the Faujasite zeolite catalyst exhibited good recyclability [18]. The different types of zeolite promoted different pathways for the catalytic pyrolysis of mannose. Ultra-stable zeolite Y (USY) induced the dehydration reaction, whereas HZSM-5 and $H-\beta$ promoted the deoxygenation reaction. Additionally, the H- β catalyst facilitated the formation of ethylene [18]. In another study, HZSM-5 and activated carbon were used for the pyrolysis of corncob. The HZSM-5/activated carbon cocatalyst was highly selective toward the formation of C₈-C₁₆ aromatics. The bio-oil yield was improved due to the synergism of the properties of the cocatalyst components [19]. HZSM-5, NaOH-modified ZSM-5, and Ni/ZSM-5 catalysts were used for the pyrolysis of lignin. The Ni/ZSM-5 catalyst promoted the demethoxylation and dehydroxylation of the biomass, which are the main reaction pathways for the transformation of the oxygenated compounds into aromatic hydrocarbons [20]. The cost of the catalyst is a particular consideration for the catalytic pyrolysis process. Therefore, catalysts from waste and/or natural materials have received considerable attention [21]. A spent fluidized catalytic cracking (FCC) catalyst was used for the pyrolysis of activated sludge from refinery wastes. The catalyst improved the properties of the pyrolytic bio-oil, such as the H/C molar ratio, saturated chemical structure, and heating values [22]. In another study, a CaO catalyst for the pyrolysis of waste cooking oil was prepared using a waste stream containing egg, crab, and clam shells. CaO promoted the aromatic formation during the pyrolysis [23]. Natural minerals, such as dolomite, have been used for the catalytic pyrolysis of cellulose and chitin. Dolomite significantly reduced the activation energy of pyrolysis, and exhibited good cracking activity to convert the anhydrosugars into small molecules [24].

In this study, natural zeolite and dolomite were used as catalysts in a catalytic pyrolysis system. An in-house designed fluidized bed reactor was used. The catalyst beds were placed in the fluidized bed primary reactor and/or hot filter secondary reactor. The effect of the configuration of catalyst location on the bio-oil yield and compositions was studied. Optimum catalytic pyrolysis conditions were identified. Finally, the properties of the bio-oil produced using the optimized conditions were analyzed.

2. Material and methods

2.1. Materials

The used biomass feedstock, *Leucaena leucocephala*, was obtained from Kalasin, Thailand. The biomass was dried, ground, and sieved within a particle size of 0.6–3 mm. Prior to pyrolysis, the biomass stock was dried in a hot-air oven to lower its moisture content to 8.5%. Table 1 summarizes the results of the proximate and ultimate analyses, and heating values of the biomass feedstock.

Natural zeolite and dolomite catalysts were obtained from TPI POLENE Bio-organics Company Limited, Thailand. To remove the impurities, the catalysts were calcined at 500 °C for 4 h. Then, they were sieved within 0.2–0.5 mm and 0.5–3 mm for use in the primary and secondary reactors, respectively.

[23]

Leucaena leucocephala

Table I	
Properties	

operties of L	eucaena leucocephala.	
Analysis	Leucaena leucocephala (This work)	
Drovimate ana	lucis (ut% dry basis)	

Proximate analysis	(wt%, dry basis)	
Volatile matter	80.10	79.96
Fixed carbon ^a	17.67	3.32
Ash	2.2	16.72
Moisture ^b	8.5	3.70
Ultimate analysis (1	wt%, dry, ash-free basis)	
Carbon	45.39	46.20
Hydrogen	5.42	5.80
Nitrogen	0.63	0.90
Sulfur	0.18	N/A
Oxygen ^a	49.19	47.3
HHV [MJ kg ⁻¹]	18.23	18.40
LHV [MJ kg ⁻¹]	17.05	17.13 ^c

^a Calculated by difference.

^b As-received basis.

^c Calculated through *LHV* $\left(\frac{MJ}{kg}\right) = HHV - 21.822 \left(\frac{\%H}{100}\right)$

2.2. Characterization of catalysts

The surface area and porosity of the samples were determined from N₂ adsorption–desorption isotherms using V-Sorb 2800P analyzer (Gold APP Instruments Corporation). The BET specific surface area was calculated through the Brunauer–Emmett–Teller (BET) method, whereas the pore size and average pore diameter distribution were obtained through the Barrett–Joyner–Halenda (BJH) method; the total pore volume was determined from the volume of the adsorbed N₂ gas at a relative pressure ≈ 0.99 . Powder X-ray diffraction (XRD) was performed from 2 $\theta = 5-80^\circ$ using a diffractometer (D8 ADVANCE, Bruker) with a Cu K\alpha radiation ($\lambda = 0.154$ nm) operated at 40 kV and 40 mA. The data was recorded at a scan speed of 0.1 s step⁻¹.

2.3. Catalytic pyrolysis of Leucaena leucocephala

The catalytic pyrolysis of Leucaena leucocephala was performed using a fluidized bed reactor shown in Fig. 1 The reactor system consisted of six main units including the fluidized bed reactor (primary reactor), two cyclone separators, hot filter (secondary reactor), and water-cooled, electrostatic precipitator (ESP), and ice-salt condensers (Fig. 1). The biomass was stored in a glass hopper and then fed to the fluidized bed reactor using a N₂ pneumatic conveyor. The fluidized bed reactor with an internal width and a height of 100 and 800 mm, respectively, was made up of stainless steel. The cyclones were used to remove the fine particles in the pyrolysis vapor, which were then collected in a char pot. The secondary reactor was a fixed-bed reactor, which was used to upgrade the pyrolysis vapor. The bio-oil condenser was composed of three consecutive units of condensers. The heavy oils were collected using a water-cooled condenser, which is a double tube heat exchanger, and an ESP condenser operated at 15 kV. The light oils were collected using an ice-salt unit, which is a glass tube condenser. The non-condensed products were sampled and analyzed by a gas chromatograph (Shimadzu GC-8A) equipped with Porapak N (80/100 SS 2.3 mm I.D. \times 1 m) and Unibeads C (60/80 SS Col. 3 mm. I.D. \times 2 m), with a thermal conductivity detector (TCD).

For the pyrolysis experiments, 500 g of the catalyst was placed in both the primary and secondary reactors. The pyrolysis temperature in both reactors was 500 °C. The biomass in the hopper was fed into the fluidized bed reactor using a conveyor at a rate of 5 L min⁻¹ with N₂ as the fluidized gas purged at a flow rate of 20 L min⁻¹. Considering the two abovementioned rates, the biomass feed rate was 1 kg h⁻¹. In the experiment, four catalyst combinations were designed. In the first two configurations, either natural zeolite (NZ/NZ) or dolomite (DM/DM) was introduced in both reactors. The remaining catalyst design



Fig. 1. (a) Schematic, and (b) actual setup of the integrated pyrolysis and catalytic upgrading system.

combinations were labelled as NZ/DM and DM/NZ, where the catalyst used in the primary (secondary) reactor was NZ (DM) and DM (NZ), respectively. All experiments were replicated three times.

2.4. Analysis of the properties of the bio-oil products

The compositions of the bio-oil products were determined using a Shimadzu GCMS-QP2010 gas chromatograph mass spectrometer (GCMS). Prior to its injection, the bio-oil sample was diluted using a 1:1 methanol-dichloromethane solution and then filtered using a Filtrex nylon filter with a pore size of 0.2 µm. The products were separated using a 30 m \times 0.25 mm (internal diameter) Restex Rtx-5MS (Restex, USA) column with a film thickness of 0.25 µm. The temperature was held at 60 °C for 2 min and then increased to 270 °C at a rate of 5 °C min⁻¹. The final temperature was maintained for 5 min. The injector temperature was 270 $^{\circ}$ C with a split ratio of 100. He gas was used as the carrier introduced at a linear velocity of 40 cm s⁻¹. The mass spectrometer was operated in an electron impact (EI) mode at ion source and interface temperatures of 250 and 230 °C, respectively. The m/z range selected was 20-650 with a rate of 0.5 s. The data acquisition and processing were performed using SHIMADZU LabSolutions. The degree of deoxygenation was calculated based on the GC-MS data using Eq. (1):

$$degree of \ deoxygenation = \frac{T_{no-catalyst} - T_{catalyst}}{T_{no-catalyst}} \times 100$$
(1)

Here, $T_{no-catalyst}$ and $T_{catalyst}$ denote the total mole percentages of oxygen for all the oxygenated compounds, as determined from the GC–MS data of the pyrolysis oil, without catalyst and with catalyst, respectively [25].

The high heating values (HHV) of the bio-oil were determined according to the DIN 51900 method using an SMD Torino bomb calorimeter. Its kinematic viscosity was measured according to the ASTM D445 standard at 40 °C using a Cannon–Fenske opaque viscometer (size 350). The density of the bio-oil was calculated using a 5 mL density bottle following the ASTM D4052 standard. Lastly, the flash and fire points were determined according to the ASTM D93 standard. The water content in bio-oil was determined by volumetric Karl Fischer titration following the ASTM E203 method.

3. Results and discussion

3.1. Characterization of the catalysts

Fig. 2 shows the XRD patterns of the catalysts. The peaks observed at the XRD pattern of dolomite at $2\theta = 23^{\circ}$, 29° , 35° , 39° , 43° , 47° , and 57° can be attributed to calcite (CaCO₃). The peaks at $2\theta = 26^{\circ}$ and 31° can be indexed to anhydrite (CaSO₄) (Fig. 2a). From Fig. 2b, the well-defined peaks at $2\theta = 25^{\circ}$ and 31° observed at the XRD pattern of zeolite, as well as the minute peaks at $2\theta = 38^{\circ}$, 49° , and 52° , can be indexed to clinoptilolite. Trace amounts of mullite was also found as confirmed by the



Fig. 2. XRD patterns of the (a) dolomite, and (b) natural zeolite catalysts.

peaks at $2\theta = 36^{\circ}$, 41° , and 56° . The calcite phase in dolomite can serve as the active site during the catalytic pyrolysis. The calcite (CaCO₃) promoted the cracking reaction through the dehydration pathway as reported in several previous studies [26,27]. The clinoptilolite and mullite are natural aluminosilicate compounds that can potentially serve as the acid sites to promote the cracking reaction during catalytic pyrolysis. Apart from the crystallographic properties, other physical properties such as surface area, pore diameter, and pore volume are also important descriptors for the catalytic activity of the catalyst (Table 2).

The BET specific surface areas of natural zeolite and dolomite were 14.0 and 32.6 m² g⁻¹, respectively. The pore diameter of natural zeolite was twice that of dolomite. On the other hand, the pore volume of natural zeolite was less than that of dolomite. However, the Si/Al ratio of natural zeolite was close to commercial HZSM-5 zeolite, and natural zeolite which could perform a catalytic pathway similar to HZSM-5 zeolite [28,29]. Although the trend observed in the physical properties of the catalysts did not have a direct effect on their catalytic activities. Therefore, the interaction between the physical and crystallographic

Table 2

Physical properties of the natural zeolite and dolomite catalysts.

Catalyst	Calcination Temperature/ Time	Bulk density [kg m ⁻³]	Particle density [kg m ⁻³]	BET surface area [m ² g ⁻¹]	Mean pore diameter [nm]	Total pore volume [cm ³ g ⁻¹]	Si/Al ratio
Natural zeolite	500 °C/4 h	700	1165	14.0	25.4	$\textbf{6.83}\times 10^{-2}$	12.34
Dolomite	500 °C/4 h	1128	1728	32.6	13.7	0.20	N/A

properties should be considered.

3.2. Effect of the catalyst combination design on product yield

In this study, natural zeolite and dolomite were used as catalysts for pyrolysis using primary fluidized bed and secondary hot filter reactors. Four different catalyst combination designs were utilized, namely NZ/NZ, DM/DM, NZ/DM, and DM/NZ. The gas yield increased by 10% in the presence of the catalysts with decreasing the bio-oil yield by 10% (Fig. 3). The gas yield increased due to the promotion of the deoxy-genation step through the decarboxylation and decarbonylation reaction pathways using the catalyst [30,31]. The catalyst combination design did not have a substantial effect on the overall product yield. However, this factor affected the heavy and light oil fractions in the bio-oil.

The heavy fraction of the bio-oil was obtained using the water-cooled and ESP condensers meanwhile, the light oil was collected using the ice-salt condensing unit. The catalyst promoted the cracking reaction, which increased the fraction of the light bio-oil. The fraction of the heavy bio-oil decreased by approximately 13% after the introduction of the catalysts (Fig. 4). The DM/DM catalyst configuration exhibited high cracking activity, which resulted in considerable improvement in the light bio-oil fraction. The highest light bio-oil fraction (21.6 wt%) was achieved using this catalyst combination design. The CaCO₃ phase in the dolomite catalyst exhibits good cracking activity [32,33]. Therefore, the dolomite in the primary reactor facilitated the in situ cracking of the pyrolysis vapor which was further improved by the dolomite catalyst in the secondary reactor. The yield of the light oil using the DM/NZ catalyst design was higher than those using the NZ/DM and NZ/NZ configurations. These results further confirm that dolomite promoted the cracking reaction during the pyrolysis, and its presence in the primary reactor synergized the reaction activity. Fig. 5 presents the distribution of the heavy bio-oil fractions obtained using the four catalyst combinations. The heavy bio-oil fraction obtained from the water-cooled condenser is more viscous than that collected from the ESP unit. As previously discussed, the total heavy bio-oil fraction obtained using the DM/DM configuration was low due to the good cracking activity of the catalyst.



Fig. 3. Yield and composition of the product from the integrated pyrolysis and catalytic upgrading system.



Fig. 4. Composition of the total bio-oil yield from the integrated pyrolysis and catalytic upgrading system.



Fig. 5. Heavy bio-oil fractions obtained from the water-cooled and ESP condensing units.

As such, the lowest heavy bio-oil fraction from the water-cooled condenser was also obtained using the DM/DM catalyst design, implying that the bio-oil produced using this configuration was the least viscous. This result further confirms the excellent cracking activity of the DM/DM catalyst.

3.3. Effect of catalyst combination design on the composition of heavy bio-oil and gas product

Fig. 6 presents the compositions of the heavy bio-oil fractions. Without the catalyst, the heavy bio-oil products from both the watercooled condenser and ESP unit are largely ketone. In addition, the product from the ESP unit, depicted in Fig. 6b, showed 39% ketone content, which were possibly formed through a ketonization reaction



Fig. 6. Compositions of the heavy bio-oil fractions obtained from the (a) watercooled, and (b) ESP condensing units.

during the pyrolysis of the biomass [34]. In the presence of the catalyst, the amounts of ketones and carboxylic acids in the heavy bio-oil decreased by 30-50%. These functional groups were converted to hydrocarbons through decarbonylation and decarboxylation. The heavy bio-oil fraction obtained using the ESP condensing unit was primarily composed of hydrocarbons (Fig. 6b). In the presence of the catalysts, the hydrocarbon fraction increased ten fold. In particular, the highest hydrocarbon fraction in the heavy bio-oils from the water-cooled and ESP condensers was achieved using the NZ/NZ catalyst. The acid sites on the surface of the natural zeolite possibly accelerated the oligomerization of the fragmented molecules into larger hydrocarbons. The dolomite possibly neutralized the acid because of the presence of $CaCO_3$ [35]. The carboxylic acid content of the bio-oil significantly decreased using the NZ/DM and DM/DM catalyst configurations. The dolomite catalyst exhibited good cracking activity, which resulted to high bio-oil yield and low acid content. However, the catalytic pyrolysis using dolomite was not selective toward the formation of the desired product.

Fig. 7 shows the degree of the deoxygenation of the heavy bio-oil fractions, which reflect the oxygen removal efficiency of the catalysts. The heavy bio-oil from the water-cooled and ESP condensing units produced using the natural zeolite catalyst exhibited the highest deoxygenation degree. In particular, the deoxygenation degree of the product from the ESP unit was as high as 60%. The oxygen content of the heavy bio-oil from the water-cooled condenser produced using dolomite was higher than that obtained without a catalyst. Therefore, the utilization of the natural zeolite catalyst produced high-quality bio-oil.

The composition of gas products is presented in Fig. 8. The carbon

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Heavy bio-oil from water-cooled condenser

Fig. 7. Degree of deoxygenation of the heavy bio-oil fraction from the watercooled and ESP condensing units.



Fig. 8. composition of the gas product from the integrated pyrolysis and catalytic upgrading system.

dioxide and carbonmonoxide were the main components of the gas product from pyrolysis without catalyst due to thermal decarboxylation and decarbonylation. The presence of the catalyst caused the decreasing carbon dioxide and carbonmonoxide fraction with increasing hydrocarbon because the catalyst preferably promoted the C–C bond cracking pathway.

Fig. 9 shows the proposed pyrolysis mechanism. The biomass feedstock was mainly composed of sugar-based polymers, such as cellulose and hemicellulose, and aromatics, such as lignin. The sugar-based polymers were cracked to the monosaccharides through the acid hydrolysis at the β-1,4-glycosidic bond. The glucose saccharide was produced through cellulose hydrolysis and then formed levoglucosan during the pyrolysis reaction (step C1). The acid hydrolysis step could occur under thermal or acid-catalyzed cracking conditions. Then, the levoglucosan molecule was deoxygenated through dehydration, decarboxylation, and, decarbonylation, thereby forming furan compounds (step C2). In step C3, the C-C and C-O bonds of the furan compounds were cleaved, forming light oxygenated compounds. Finally, the oxygenated compounds were transformed through various reactions, such as hydrogenation (ketones), aldol condensation (aldehyde), dimerization, and hydrodeoxygenation (step C4) to obtain hydrocarbons. From each step, several small gaseous molecules were released including CO, CO₂, H₂O and light hydrocarbons. The lignin fragmented after the breaking of the β -O-4 ether linkage and C-C bond through the Bronsted acid protonation, forming monophenol compounds. Monolignols, such as syringyl propane, guaiacyl propane, and 4-hydroxyphenyl propane, were produced after step L1. In the presence of the acid catalyst, these monolignols possibly underwent selective demethoxylation, dealkylation, and dehydration to form phenol, alkyl phenol,



Fig. 9. Proposed reaction pathways for the catalytic upgrading of leucaena leucocephala over natural zeolite and dolomite catalysts.

and hydrocarbon in step L2. Without the catalyst, the bio-oil produced in this study through the biomass pyrolysis exhibited variations due to the uncontrollable reaction pathways of the radical compounds. The catalysts controlled the reaction pathway toward the formation of mono-functional compounds and hydrocarbons from the sugar-based and lignin-based polymers [36].

3.4. Effect of the catalyst combination design on the fuel properties of the bio-oil

Aside from the product yield and composition, the fuel properties and water content of the bio-oil should also be considered for its practical applications and compared to bio-oil standards [37]. Table 3 summarizes the HHV, viscosity, density, flash point, and fly point of the produced bio-oil.

The viscosity of the bio-oil produced using the catalysts decreased by > 60% relative to that of the bio-oil product obtained without dolomite and/or natural zeolite. The deoxygenation and cracking reactions decreased the molecular weight and intermolecular interactions in the bio-oil, which in turn decreased the viscosity. Furthermore, the reductions in the molecular weight and intermolecular interactions also decreased the flash and fly points of the product. Nevertheless, the flash and fly points of the bio-oil produced in this study are still within the acceptable values based on the fuel oil standards. The water content was in the range of 10–20 wt%, which is within the bio-oil standard. The heating value is also an important parameter for fuel characterization. The HHV of the bio-oil produced using the catalysts was higher by 25% of that of the product obtained without the dolomite and zeolite. The

HHV increased due to the reduction of the oxygen content and molecular weight of the bio-oil products. Table 4 presents the benchmarking of the catalytic performance of various catalyst types and configurations in a fluidized bed reactor. The natural zeolite and dolomite showed 57.3% selectivity to hydrocarbon which was the highest in comparison with previous studies. However, the HHV of the bio-oil produced in this work was below the standard set for fuel oils therefore, bio-oil could be considered as a blend stock.

4. Conclusions

The catalytic pyrolysis of Leucaena leucocephala was carried out in a fluidized bed reactor using natural catalysts such as zeolite and dolomite. The catalyst beds were placed in two different locations in the reactor system, which utilized fluidized bed and hot filter reactors. The gas yield after the pyrolysis increased by 10% in the presence of the catalysts due to the promotion of the deoxygenation reaction. Dolomite facilitated the cracking reaction, which increased the light bio-oil yield. On the other hand, the natural zeolite catalyst exhibited good catalytic activity toward the deoxygenation reactions, which promoted the production of hydrocarbons. The catalyst design combination did not have a significant synergistic effect on the catalytic activity. The viscosities of the bio-oil products obtained using the catalysts were lower by 60% of that of the bio-oil produced without dolomite or zeolite. The flash and fly points of the product were within the fuel oil standards. However, its HHV was below the acceptable value. The bio-oil sample can be used as a blend stock instead to lower the sulfur content of commercial fuel oils. The catalytic pyrolysis process design employed in this study can be

Table 3

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Properties	Bio-oil source	No catalyst	Catalyst configu	iration			Bio-oil standard [37]
			NZ/NZ	DM/DM	NZ/DM	DM/NZ	
HHV [MJ kg ⁻¹]	Water-cooled condenser ESP condenser	$\begin{array}{c} 28.5\pm1.2\\ 30.7\pm1.4 \end{array}$	$\begin{array}{c} 35.8 \pm 1.5 \\ 37.3 \pm 1.1 \end{array}$	$\begin{array}{c} 32.0\pm1.2\\ 34.6\pm0.9\end{array}$	$\begin{array}{c} 33.4\pm1.1\\ 35.8\pm1.2 \end{array}$	$\begin{array}{c} 34.4\pm1.3\\ 36.8\pm1.2 \end{array}$	13–18
Viscosity [cSt]	Water-cooled condenser ESP condenser	$\frac{184.9 \pm 2.4}{211.3 \pm 1.4}$	$52.5 \pm 1.4 \\ 73.5 \pm 1.4$	$\begin{array}{c} 41.4 \pm 1.4 \\ 62.2 \pm 1.4 \end{array}$	$\begin{array}{c} 48.2 \pm 1.4 \\ 65.8 \pm 1.4 \end{array}$	$51.9 \pm 1.4 \\ 69.7 \pm 2.4$	15–35
Density [kg m ⁻³]	Water-cooled condenser ESP condenser	$\begin{array}{c} 1145\pm3.2\\ 1216\pm2.4 \end{array}$	$\frac{1055 \pm 2.5}{1149 \pm 3.1}$	$\frac{1291 \pm 2.6}{1390 \pm 2.8}$	$\begin{array}{c} 1247\pm2.1\\ 1310\pm2.7 \end{array}$	$\begin{array}{c} 1139 \pm 2.2 \\ 1215 \pm 2.5 \end{array}$	1100–1300
Flash point [°C]	Water-cooled condenser ESP condenser	$245.8 \pm 2.7 \\ 238.7 \pm 3.4$	$\frac{122.7 \pm 1.7}{118.8 \pm 2.5}$	$\frac{133.4 \pm 2.6}{131.7 \pm 1.5}$	$\frac{134.2 \pm 1.4}{127.3 \pm 2.5}$	$\frac{130.8 \pm 1.1}{122.7 \pm 2.7}$	40–110
Fire point [°C]	Water-cooled condenser ESP condenser	$\begin{array}{c} 264.3 \pm 2.2 \\ 251.8 \pm 1.8 \end{array}$	$\frac{132.7 \pm 3.2}{125.4 \pm 1.8}$	$\frac{153.4 \pm 1.9}{146.4 \pm 2.4}$	$\frac{147.8\pm 3.1}{139.5\pm 2.4}$	$\frac{139.8 \pm 3.1}{132.3 \pm 1.8}$	_
Water content [wt.%]	Water-cooled condenser ESP condenser	$\begin{array}{c} 14.6\pm0.8\\ 12.6\pm0.9\end{array}$	$\begin{array}{c} 16.2\pm0.9\\ 15.4\pm0.8\end{array}$	$\begin{array}{c} 18.2\pm0.1\\ 13.2\pm0.1\end{array}$	$\begin{array}{c} 19.8\pm0.2\\ 14.3\pm0.8 \end{array}$	$\begin{array}{c} 19.4\pm0.2\\ 16.1\pm0.6\end{array}$	20–30

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comparison of cata	alytic perforr	mance of various	catalyst types and o	configurations in a	a fluidized bed rea	ictor.						
Condition				Catalyst				Fluidized	Reactor	Results		Ref.
Biomass feedstock	Temp. (° C)	Flow rate (kg h ⁻¹)	Atmosphere	Reactor	Hot filter	Temp. (°C)	Catalyst to biomass	ID (mm)	Height (mm)	Oil yield (wt.%)	Main composition of oil	
Switchgrass/red oak sawdust	500	0.04-0.2	58% H ₂ in N ₂ (total 1 atm)	Non- catalyst	Ru-TiO ₂	400	0.1-0.15	23.62	304.8	49.90	Mostly acetic acid, acetol (28.22%), furan and pyran (20.25%), and phenolic (30.76%)	[38]
Corncob	400-700	0.036	N_2 (1atm)	HZSM-5	Non-catalyst	400-700	0.2	30	400	13.70	74.22% hydrocarbon	[39]
loblolly pine	501	0.156 - 0.222	N_2 (1atm)	γ-Al2O3	Non-catalyst	350-500	2.25 - 3.21	63	254	42.30	45% phenol	[40]
White Oak Wood	500	2.5	N_2 (1atm)	β zeolite	Non-catalyst	450	0.32	76	203	29.00	Oxygenated compounds	[41]
Pine	500	0.42	N_2 (1atm)	Silica	ZSM-5	500-600	0.55 - 1	52	430	Up to 40	10–30% hydrocarbon	[42]
Leucaena	500	1	N_2 (1atm)	Natural	Dolomite/	500	0.5	100	800	34.60	57.3% hydrocarbon	This
leucocephala				zeolite/ dolomite	Natural zeolite							work

employed as an alternative route for the production of bio-oils.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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