Thermo-economic evaluation of a banana waste pyrolysis plant for biofuel production

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Abstract: Health and environmental problems are presented in the world due to gases produced during the combustion of fossil fuels. This problem has focused on researching the production of renewable fuels from biomass. One type of biomass are wastes produced during culture of banana which its industry for every tonne of banana harvested there are, on average, four tonnes of wastes which also contaminated environment. This paper aims to evaluate the economic and exergetic viability of a banana waste pyrolysis plant in Colombia. The processes of the plant were simulated on Aspen Plus®, which solved material and energy balance. The yield of products was validated with experimental results from literature, where differences could be attributed to biomass composition and reactor type. The results obtained revealed yields of 40.20%, 24.75% and 35.07% from bio-oil, char and gas respectively. Economic and exergy data were combined for an exergy costing method to carry out the exergoeconomic analysis. The overall exergy efficiency of the plant was 61.76%. The highest exergy destruction is located in the dryer and the pyrolysis reactor, which accounts for 35.84% and 32.96% respectively. The combustion chamber has the highest exergy destruction cost (180.19 $/h) and a low value of the exergoeconomic factor (5.96 %), indicating that an improvement of the equipment is required, even though this means an increment in cost investment. The study reveals, that the analysis made demonstrates the potential of bio-oil production from banana wastes, and also the possibility to use gases products in a combined heat and power plant. In addition, the study evidence in which process it is necessary to achieve better results of efficiencies in order to improve the whole plant.

1. Introduction.

The extraction and use of fossil fuels have generated environmental impacts due to gases and particulate matter (PM) produced in combustion (Munawer, 2018). Among the gases, there are nitrogen oxide (NO_X), carbon dioxide (CO_2), and sulfur dioxide (SO_2) (Gaffney & Marley, 2009). These substances have greatly affected the environment, giving rise to phenomena such as the greenhouse effect, the increase of the temperature of the earth, and the thawing of glaciers (Caballero, Lozano, & Ortega, 2007; Herrán, 2012). In addition to the environmental problems mentioned, the consequences for people's health due to the exposure of these gases cause 4.5 million deaths per year (Farrow, Kathryn, & Lauri, 2020). These problems have led to obtaining fuels from renewable sources such as biomass since these fuels help to reduce greenhouse gas emissions (C. T. Wright, Boardman, Yancey, & Sokhansanj, 2011) because the carbon footprint is lower in these biomass-based fuels (Organización de las Naciones Unidas Para la Aagricultura y la Alimentación, 2008).

The variety of feedstock for the obtaining of biofuels is immense, many studies use tires(Altayeb, 2015), plastic wastes(Veksha, Giannis, & Chang, 2017), algae(Chernova, Kiseleva, Larina, & Sytchev, 2019), wood (Lestinsky & Palit, 2016), leaves or branches (Klinger et al., 2018), due to their economic potential of biomass because of the volumes of agricultural production (Moreno, Samerón, & Perea, 2019). Biofuels production is classified in four-generation based on feedstocks and method of production (Alalwan, Alminshid, & Aljaafari, 2019). Currently, sugar cane and palm oil (first-generation), are the principal feedstocks used to obtain bioethanol and biodiesel around the world, respectively (Eduardo, Young, & Steffen, 2008). However, the use of first-generation biofuels has generated negative environmental and social impacts, such as the increase in food prices (Castro, 2012). Second-generation biofuels have different studies that evidence the potential energy of the lignocellulosic feedstock such as agricultural and forest residues (García, Pizarro, Lavín, & Bueno, 2012), (Cortes, 2014). Third and fourth generation biofuels are promising because microorganisms are used as feedstocks (Leong, Lim, Lam, & Uemura, 2018), the advantages are the high energy content, low emission, high oil content, the ability to reduce CO_2 and being eco-friendly (Yi-Feng & Wu, 2011), even so, it is not economically viable due to the low yield and high production costs (B. Abdullah, Anuar, Muhammad, Shokravi, & Ismail, 2019) which represents the need of more investigations to achieve higher yields. Those problems with third and fourth generations demonstrate that the second generation is ideal since food industry will always generate wastes that must be reduced, which is one of the targets for a circular economy (Morseletto, 2020) with environmental and socioeconomic benefits.

As mentioned, second-generation biomass has different advantages like the minimal impact on food price, removal of wastes and no additional land is needed (Carriquiry, Du, & Timilsina, 2011). As there are different kinds of crops that their waste represents high potential energy, such as rice straw, wheat straw or wood sawdust (Sahoo, Kumar, & Prasad, 2020). Banana industry, only uses from 20 to 30 % of its mass, leaving 70 to 80 % of waste (Mazzeo M., León Agatón, Mejía Gutierrez, Guerrero Mendieta, & Botero López, 2010). According to Fernandes (Fernandes, Marangoni, Souza, & Sellin, 2013), for every tonne of banana harvested there are on average four tonnes of lignocellulosic wastes.

Furthermore, it is worth mentioning the thermochemical conversion routes, which are the most used to transform biomass into fuels and chemicals; there are direct combustion, liquefaction, pyrolysis, gasification and so on (Balat, Balat, Kırtay, & Balat, 2009). Combustion burned the biomass under an excess of O_2; in gasification, biomass reacts with steam water, low air or O_2 and it is widely recognized for hydrogen production (Beheshti & Ghassemi, 2015). On the other hand, pyrolysis processes transform biomass in the absence of oxygen for the production of solids, liquids and high heating values (Cerdá, 2012), (Lédé & Authier, 2011). Based on the heating rate, pyrolysis can be slow (>100°C/s), fast (1000°C/s) and flash (>1000°C/s) (Basu, 2010a), (Boateng, 2020) . Among these processes, fast pyrolysis maximizes liquid yield as a result of a high heating rate and short residence time (2-3 s) (Tessini, Segura, & Berg, 2013).

Considering this, different studies had determinate the potential usage of banana waste in fast pyrolysis owing to gas yield from 32.5 to 49.6 %, liquid from 27 to 29.65% and solids from 23.3 to 39.5% (Ghosh, Das, & Chowdhury, 2019; Omulo, Banadda, Kabenge, & Seay, 2019; Ozbay, Yargic, Zerrin, Sahin, & Yaman, 2019). Yields may vary depending on the feedstock, reactor, heating rate and pyrolysis temperature (Basu, 2010b). These studies conclude that the products obtained have attractive characteristics to be used as fuels after applying refinery processes (Hussain, Zhao, Ren, Rasool, & Raza, 2019; Sellin, Ricardo, Marangoni, & Souza, 2016). The liquids obtained are also a potential source to obtain valuable chemical products with great commercial opportunities (N. Abdullah, Sulaiman, & Mohd, 2015), (Czernik & Bridgwater, 2004). Thus, Colombia and specifically the department of Antioquia represents a great opportunity due to in this region is concentrated 73% of the cultivated area in the country (FINAGRO, 2018), with 37,838 hectares and an annual banana production of 1,299,526 tons (2019) (Dirección de Cadenas Agricolas y Forestales, 2020). Likewise, there are many studies on the economic part results may vary depending on feedstock quality. But, the fact that pyrolysis products may be considered as a substitution of fossil fuels evidence the high benefit of producing bio-oil, char and syngas (Badger, Badger, Puettmann, Steele, & Cooper, 2011).

According to this, this article aims is to evaluate the economic and exergetic viability of a fast pyrolysis plant to obtain biofuels, using banana wastes as feedstock in Colombia. To assess the performance of the whole industrial-scale fast pyrolysis plant, a simulation is needed. The simulation is carried out in Aspen Plus®, which provided mass and energy balances for a thermo-economic analysis, where methods and data will be provided for the model validation and the efficiency of the products obtained.

2. Materials y Methods.

The study was carried out by defining that the plant would be in Uraba, a region of Antioquia (Colombia); also to have an idea of how much waste could be obtained, an area of 100 ha was established to make reference of the wastes from the banana-producing farms, data was obtained from the annual report of Finagro (FINAGRO, 2018).

2.1. Feedstock specification

The banana wastes chosen for the study were leaves and pseudo-stem which are the main wastes that are left in the plantation without treatment (Mazzeo M. et al., 2010). Banana peels are excluded because the most part is disposed of at the place of consumption (Mazzeo M. et al., 2010). Biomass is modeled as a non-conventional solid component that cannot be represented by a molecular structure (Aspen Technology Inc, 2000). For that reason, a set of ultimate and proximate analyses are required. Table 1 compiles data of leaves and pseudo-stem from authors; with that data, an average of each value is obtained for the simulation. Aspen Plus® requires that proximate and ultimate analysis sums up to 100%, except for moisture. Moisture content presented by authors was obtained after drying and size reduction, meaning that is the value that is sought to be obtained for pyrolysis.

Table 1. Summary of Proximate and Ultimate Analysis of banana waste from different authors.

	Leaves (Fernandes et al., 2013)	Pseudostem (Ghosh et al., 2019)	Leaves (Sellin et al., 2016)	Pseudostem (N. Abdullah et al., 2015)	Leaves (Kabenge et al., 2018)	Pseudostem (Kabenge et al., 2018)	Selected data
Proximate analysis wt% (dry basis)							
Moisture	8.30	15	7.80	10.20	6.67	7.98	**9.33**
VM	78.80	76.67	78.20	80.60	83.35	89.43	**82.68**
Ash	8.70	6.80	6.20	12.50	9.05	9.36	**8.13**
FC	12.5	1.53	15.6	6.90	7.60	1.21	**9.20**
Ultimate analysis wt%							
C	43.5	40.4	43.5	21.95	38.57	33.46	**35.76**
H	6.30	6.02	6.20	2.86	6.44	6.44	**4.71**
N	1.30	0.11	0.86	0.24	2.45	0.80	**0.96**
O	48.70	53.47	42.30	74.95	43.49	49.94	**50.14**
S	0.20	-	0.95	-	0.0032	0.04	**0.31**

Note: wt%- Percentage weight, VM-Volatile matter, FX- Fixed Carbon

2.2. General description of the plant

A brief description of the general gets presented in Table 2, the specifications presented for each process were established from articles about pyrolysis. Further, figure 1 shows a schematic representation of all processes.

Table 2. Description of pyrolysis plant

Process	*Description*	*Specification*
Feedstock reception	Reception of leaves and pseudo-stems from the banana harvest, previously cleaned and cut.	Feedstock must be without dirtiness.
Drying	The raw material is dried to reduce its moisture content in a range of 7-15%.	Biomass temperature: 25°C Air at 25°C and humidity of 80%
Grinding	The raw material gets ground to obtain a fine powder.	Grinding and sieving must ensure a maximum size of 2 mm
Sieving	The powder is sieved to discard diameters greater than 2 mm.	
Reactor	The powder enters fluidized bed reactor to start the pyrolysis process, nitrogen is the carrier gas.	Reactor at 500°C and 1 atm, residence time of 2 seconds (Bridgwater & Peacocke, 2000) 2,75 kg of N_2 per kg of biomass (M. M. Wright et al., 2010)
Combustion chamber	Char and a part of non-condensable gases are burnt to provide energy for the reactor.	Atmospheric air is compressed at 9 atm of pressure and 342.57°C. Air burns with char and gas.
Heat exchanger	To pre-heat the carrier gar, the exhausted gases from the combustion chamber are employed.	Nitrogen inlet at room temperature. The outlet is at pyrolysis temperature.
Cyclone	From the combustion of char the ash are separated from gases.	Approximately 90% particle removal (Ward, Rasul, & Bhuiya, 2014).
Cyclone 2	The cyclone removes solid particles by the centrifugal force of the gas vortex. The pyrolytic gases and char get separated.	

Condenser	Pyrolytic gases enter the condenser to be cooled with water and obtain bio-oil and the non-condensable gases are sent to the next section.	Water inlet at 20°C and 1 atm (Westerhof et al., 2011)
Cooling tower	The outlet water for the condensation is sent to a cooling tower	Inlet air at 25°C to cool water from condenser

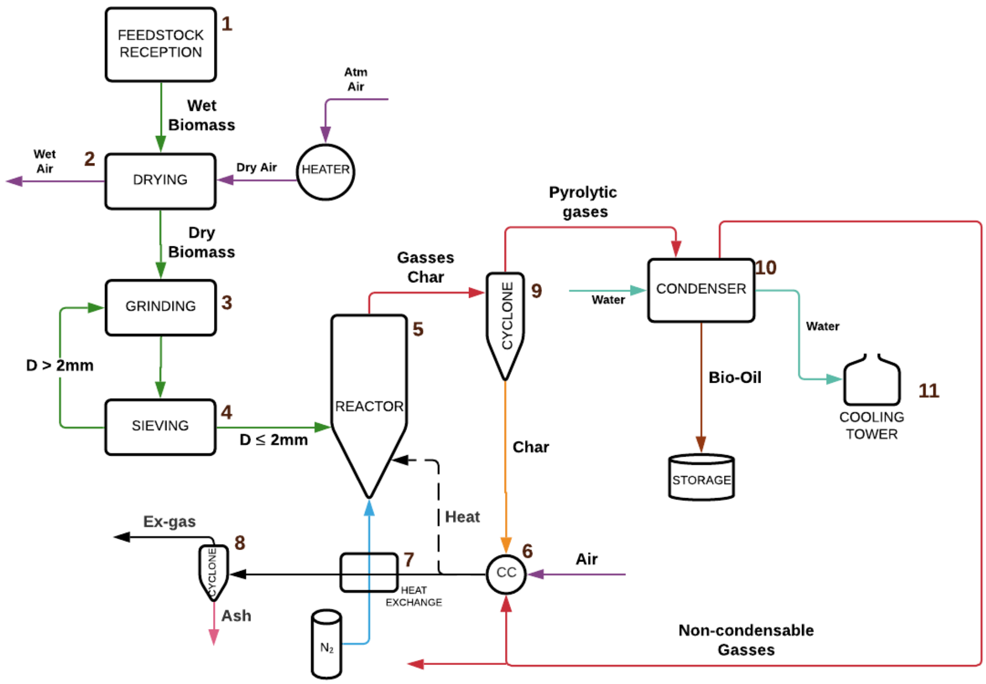

Fig. 1. Schematic diagram of the plant

2.3. Aspen Plus® model description

The description of blocks used in the simulation are shown in Table 3, although the process flow-sheet of the simulation is presented in figure 2. Each process is presented below with the principal operating characteristics. The machines for transporting and supplying raw materials are ignored in the simulation.

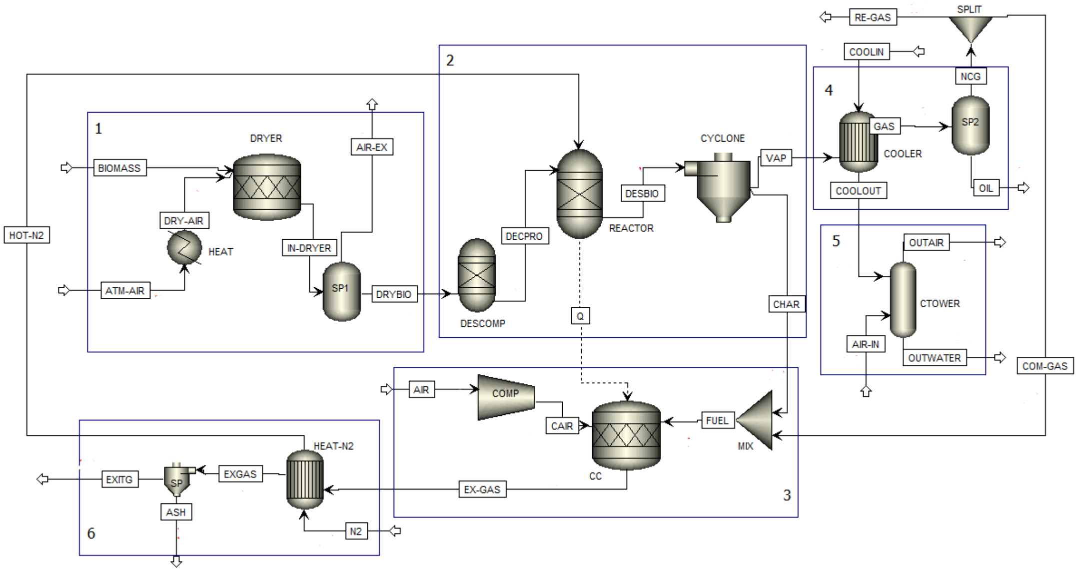

Fig. 2. Flowsheet of the simulation. 1- drying of biomass, 2- pyrolysis of biomass, 3-combustion, 4-condensation of bio-oil, 5-cooling tower, 6-pre-heat carrier gas.

Table 3. Description of Aspen Plus® unit operation

BLOCK ID	*ID ASPEN PLUS®*	*DESCRIPTION*
Heat	Heater	Relative humidity of air is reduced and the temperature increase.
Dryer	RStoc	Moisture content of biomass reduction.
SP1	Flash2	Separation water from dry biomass.
Descomp	RYield	Conversion non-conventional biomass into conventional components.
Reactor	RYield	Definition of products based on experimental data
Comp	Compressor	Compression of atmospheric air
CC	RStoc	Combustion of natural gas
Cyclone	SSplit	Separation of solids and gaseous products.
Condenser	Heatx	Condensation of gases to produce bio-oil.
SP2	Flash2	Separation of the bio-oil and the non-condensable gases.
CTower	RadFrac	Cooling outlet water from condensation with air
Heat-N_2	Heatx	Heating of nitrogen before enter the reactor
SP	SSplit	Separation of ashes from combustion gases

2.3.1. Feedstock pre-treatment (drying, grinding and sieving)

Drying the biomass is an important initial process to reduce the moisture content considering that high moisture will affect product yield (García et al., 2012). Ambient air is used for the drying process. The location of the pyrolysis plant is Antioquia (Colombia), therefore dry-bulb temperature (25°C) and relative humidity (80%) (IDEAM, 2020) are known. A convective dryer is used to dry the biomass, as this technique is frequently used for bio-oil production (Chandramohan, 2020). To ensure final moisture of about 10 wt% of the biomass, the inlet air is pre-heated to reduce the relative humidity and enhances the drying operation. The biomass passes through the hot air and gives up its moisture to the hot air. The heating of air was simulated with a HEATER block. The wet biomass and dehumidified air were fed into the RSTOIC block, where the nonconventional moisture of the biomass is converted into conventional water (Zheng, Kaliyan, & Morey, 2013). The operation was controlled by a calculator block to determine the final moisture of the biomass. Then, a FLASH2 block is added to separate the dried biomass and the wet air with the removed moisture of the biomass.

Particle size is a meaningful parameter for fast pyrolysis, owing to the size may affect product yields (Taylor et al., 2014), because particle size increment causes temperature gradient inside the particle, originating a lower temperature in the core than in the surface (Encinar, Beltran, & Bernalte, 1996). However, grinding process was not modeled in Aspen Plus®. The power consumption was determined from a hammer mill which is accurate for the target particle size (<2 mm) (Gil & Arauzo, 2014). The model given by Mani et al. (Mani, Tabil, & Sokhansanj, 2004), establishes a correlation between the size of the grinding and the energy requirements of a hammer mill. The energy consumption for grinding is approximately 14.57 kWh/ton.

2.3.2. Combustion chamber

The heat for the pyrolysis is provided externally. The heat need of the reactor to treat the feedstock is calculated as the difference in the reactor feed stream enthalpy and product stream enthalpy provided by the simulator. Char separated from the cyclone is directly burnt at 1000°C (Yang, Wang, Chong, & Bridgwater, 2018), due to the heat requirement of the pyrolysis reactor is not reached with the produced char, a part of the non-condensable off-gases are burnt too. The flue gas stream is pumped through the heating jacket inside the reactor skin to maintain the pyrolysis temperature. The fuel combusts in excess air to generate CO_2, H_2O and heat (Hammer, Boateng, Mullen, & Wheeler, 2013). The reactor RStoic was used in the simulation to represent the combustion of the char and non-condensable gas with air. A compressor was used to compress the air at the specifications indicated in table 2. The reactions of gases were introduced in the reactor according to the stoichiometry balance. The exhausted gases from the combustion are used also to pre-heat the carrier gas and then a cyclone separate the ashes from the gas.

2.3.3. Pyrolysis reactor

Biomass must be converted into its constituting components carbon, hydrogen, oxygen, sulfur, nitrogen and ash by specifying the yield distribution according to ultimate analysis from table 1. The split fractions were specified by the calculator block. The reactor RYIELD was used for the decomposition (Wang, Chen, Ma, Zhu, & Wang, 2012), (Nikoo & Mahinpey, 2008) this reactor is used when stoichiometry and reaction kinetics are unknown. Since biomass was defined with data from Table 1, which are on dry basis, it is necessary to convert the ultimate analysis to wet basis (Aspen Technology Inc, 2000). The calculator block is used to convert the components and provide the yield values.

The liquid product is the product of interest because of the range of attractive qualities (Pearson & Turner, 2014) and fast pyrolysis is the best option due to the bio-oil is obtained in yields up to 75 wt% (Bridgwater, 2012; Jahirul, Rasul, Ahmed, & Ashwath, 2012). However, this value may vary depending on the feedstock, reactor type, temperature, residence time and cooling process (Montoya Arbeláes, 2014). Reaction temperature is between 450 and 500°C (Ighalo & George, 2019), short residence time <2 s (Basu, 2010a). The data reported above was the data for the simulation. The reactor is a bubbling fluidized bed reactor which is ideal as it exhibits good heat transfer characteristics (Nor, Wan, Hisham, Ambar, & Hin, 2012). The bed material is heated with the gases of the combustion chamber. Nitrogen and the rest of the non-condensable gases are used as carrier gas that suspends the particles, the flow rate of carrier gas is 2.7 times (Lv, Yue, Xu, & Zhang, 2018) of dry biomass stream. Carrier gas is pre-heated in order to diminish the required heat of the reactor.

Within the simulation in Aspen Plus®, the following assumptions got implemented. Model is a steady-state, heating rate and residence time are not explicitly studied, kinetic free model, the particle size distribution of biomass was not modeled and char get assumed as a mix of pure carbon and ashes. The pyrolysis product yield can be simulated by defining the distribution of pyrolysis products based on experimental data. Pyrolysis reaction takes place at 500°C (Balat et al., 2009), due to maximum liquid yields are obtained at this temperature (Bridgwater, Meier, & Radlein, 1999), 1 atm pressure value with the assumption that no pressure drop occurs. To handle the problem of the lack of information about reactions to estimate the composition of the final product, another reactor RYIELD was used to define the distribution of pyrolysis products based on experimental data (Lv et al., 2018)

2.3.4. Cyclones

The process uses two cyclones. Products obtained from pyrolysis are fed into the first cyclone to separate solids from the gas mixture. Clean the gas is an essential step, so the solid generated during pyrolysis (char) does not interfere with bio-oil yield (Innanen, 2013). Char was sent to the combustion chamber. At the end of the process a second cyclone is used to remove ashes of exhaust gases.

2.3.5. Condenser

Clean gases contain condensable and non-condensable gases, which must be rapidly cooled to do not interfere with bio-oil yield (Klug, 2012). The block Heater is used to simulate the condensation of gases. The condenser was kept at 20°C (Westerhof et al., 2011). Water for the cooling was not shown in the simulation, but it was supposed to return to a cooling tower to reduce the temperature and be used again in the heat exchanger. The amount of water needed for condensation is calculated in Aspen Plus® according to the inlet temperature of the gases and the target outlet temperature.

2.3.6. Cooling tower

A cooling tower is used to remove the heat of the hot water from the condenser by contacting it with air at a lower temperature. Part of the water evaporates in the air that passes through the tower and escapes through the summit of the tower. The tower is modeled by the Aspen Plus® RadFrac block which allows the calculation of the liquid and vapor equilibrium on each equilibrium stage (Queiroz, Rodrigues, Matos, & Martins, 2012).

2.4. Thermo-economic analysis

2.4.1. Exergy analysis

The principles of mass and energy conservation (first law of thermodynamics) were obtained from the Aspen Plus® model. Unlike energy, the exergy shows the quality of energy in terms of workability when a system is put into the thermodynamic balance with its environment (J. F. Peters, Petrakopoulou, & Dufour, 2014). The exergy can be divided into kinetic, potential, chemical and physical (Ahmadi & Dincer, 2018). The common ones are physical and chemical exergy. The other two, kinetic and potential, are considered negligible here because of low altitude changes and relatively low airspeeds (Kotas, 1986). The exergy balance of a steady-state system is described in equation 1; where Ex_{in} and Ex_{out} represents de fuels and products respectively, Ex_{des} is the exergy destruction rate. The exergy efficiency is given by equation 2.

$$\sum Ex_{in} - \sum Ex_{out} = Ex_{des} \quad (1)$$

$$\eta_E = \frac{\sum Ex_{out}}{\sum Ex_{in}} \quad (2)$$

For each stream, exergy were calculated in terms of physical and chemical exergy as follows:

- *Biomass*

Physical exergy was not taken into account because biomass was fed at room temperature. Chemical exergy of biomass can be computed from the correlation suggested by Szargut et al. (Szargut, Morris, & Steward, 1988), according to equation 3. Where LHV_{dry} is the Low Heating Value given by equation 4 and 5; λ is latent heat of vaporization of water at 25°C; x_m is mass fraction of moisture; β_{dry} is the ratio of standard specific chemical exergy to the LHV on a dry basis given by equation 6, where x_i is the mass fraction of a certain element on dry basis; xs_{dry} is the mass fraction of sulfur on dry basis.

$$E^{ch} = \left(LHV_{dry} + \lambda \cdot x_m\right) \cdot \beta_{dry} + 9417 \cdot xs_{dry} \quad (3)$$

$$LHV_{dry} = HHV_{dry} - 2.442 * (8.936H/100) \quad (4)$$

$$HHV_{dry} = 0.3578 * \%C + 1.1356 * \%H + 0.0594 * \%N - 0.0854 * \%O - 0.974 \quad (5)$$

$$\beta_{dry} = 1.0437 + 0.1882 \cdot (x_H/x_C) + 0.0610 \cdot (x_O/x_C) + 0.0404 \cdot (x_N/x_C) \quad (6)$$

- *Pyrolysis gases*

Physical exergy is given by equation 7, where enthalpy (h) and entropy (s) are calculated by Aspen Plus®, and at a reference state(p_0, T_0) of 25°C and 101.13 kPa, respectively. For chemical exergy was considered that the gaseous streams behaved as ideal gas (Atienza-martínez, Ábrego, Mastral, Ceamanos, & Gea, 2018), expressed by equation 8; where R is the gas constant; $x_{gas,i}$ is the mole fraction of a certain gas component; $e_{ch,gas,i}$ is the specific chemical exergy of a certain gas component.

$$E^{ph} = h(p,T) - h(p_0,T_0) - T_0[s(p,T) - s(p_0,T_0)] \quad (7)$$

$$E^{ch} = \sum_i x_{gas,i} \cdot e_{ch,gas,i} + R \cdot T_0 \cdot \sum_i x_{gas,i} \cdot \ln x_{gas,i} \quad (8)$$

- *Air*

The exergy of air is calculated according to equation 9 and 10 where is physical and chemical, respectively. According to Dincer et al.(Dincer & Sahin, 2004), the equations take into account the moisture of the air, and made the calculus is more accurate. c_p is the specific heat, R is the gas constant, ω_0 and ω are humidity ratio at environmental conditions and operational conditions, respectively. This ω value was obtained according to the psychometric chart. For reference state 0.016 kg/kg. The humidity ratio at the inlet and outlet of the dryer 0.0319 and 0.0484, respectively.

$$E^{ph} = \left[(c_{p,a} + \omega \cdot c_{p,s}) \cdot (T - T_o)\right] - T_o \cdot \left\{\left[(c_{p,a} + \omega \cdot c_{p,s}) \cdot \ln\left(\frac{T}{T_o}\right)\right] - (R_a + \omega \cdot R_s) \cdot \ln\left(\frac{P_o}{P}\right)\right\} \quad (9)$$

$$E^{ch} = T_o \cdot \left\{(R_a + \omega \cdot R_{st}) \cdot \ln\left(\frac{1+1.6078 \cdot \omega_o}{1+1.6078 \cdot \omega}\right) + 1.6078 \cdot \omega \cdot R_a \cdot \ln\left(\frac{\omega}{\omega_o}\right)\right\} \quad (10)$$

- *Steam and water*

Physical exergy is given by equation 7. The chemical exergy of water is based on a reference state (p_0, T_0) of 25°C and 101.13 kPa; $9.5\ kJ/mol$ for gas and $0.9\ kJ/mol$ for liquid state.

- *Heat and power*

The outgoing and incoming non-material streams leaving and entering the processes are also calculated in the exergy balance. Heat is computed according to equation 11, where T represents the average temperature in which the heat transfer occurs which depends on the temperature of each process, T_0 is 25°C of reference state. Work is equivalent to exergy, thus power consumption is included in the exergy balances.

$$E^{th} = Q_{process} * (1 - T_0/T\ ) \quad (11)$$

2.4.2. Investment cost

The total capital investment (TCI) consists of fixed capital and operating capital. The fixed capital consist of equipment investment, installation cost and contingency. The equipment investment were carried out by first sizing the equipment, without oversize. The Purchases Equipment Cost (PEC) was extracted from references. If the PEC for the required capacity was unavailable, the cost was adjusted by applying equations 12 and 13. These equations allow the estimation of the equipment cost with the capacity required. With equation 12 the capacity of the equipment is scaled, where EC_{new} is the cost of the required equipment, EC_{old} is the known cost, j is the scaling factor and (S_{new}/S_{old}) is the ratio of capacity equipment taken from Kohl et al. (Kohl et al., 2015). Then, knowing the cost of the equipment, according to equation 13, the cost of each piece of equipment can be adjusted to the prices of the present year using CEPCI (Chemical Engineering Plan Cost Index). The CEPCI of the reference year 2020 is 723.4. All costs were projected in 2020 U.S. dollars.

$$EC_{new} = EC_{old} \cdot \left(\frac{S_{new}}{S_{old}}\right)^j \quad (12)$$

$$EC_{year1} = EC_{year2} \cdot \frac{CEPCI_{year1}}{CEPCI_{year2}} \tag{13}$$

The installation cost was estimated as 25% of the equipment investment. To accommodate any miscellaneous equipment excluded from the analysis was selected a contingency factor of 35% of the equipment investment, a typical value for chemical plant design (M. S. Peters & Timmerhouse, 2003). The operating capital consists of variable costs such as feedstock, maintenance, transportation, personnel and others. Cost estimation of operating capital is 15% of fixed capital (Mobolaji, Sai, & Panneerselvam, 2015).

2.4.3. Exergoeconomic analysis

Once fuels and products of each process component are defined, an exergoeconomic analysis allows calculating the cost of production for multiproduct processes. This analysis divides process total cost in costs of the process material stream and equipment cost. The Specific Exergy Costing (SPECO) method was applied (Lazzaretto & Tsatsaronis, 2006). The method consists of identify the exergy streams, define fuel and product and define cost streams. For all equipment units, fuels and products must be defined. Each exergy stream which undergoes a certain modification in a specific equipment unit the specific equipment cost is charged to the product flow. The cost balance considering the exergy form can be written as equation 14; it describes the sum of all exergy cost streams entering the unit *k* and the cost for the unit operation $\dot{Z}_k$(capital investment, operation and maintenance of each component) is equal to the sum of all cost streams leaving the unit *k*; $c, \dot{E}_{in,out}, \dot{W}, \dot{E}_q$ represent cost, in and out exergy streams, work and heat transfer respectively. Using the SPECO method, a system of linear equations is calculated from the cost balance of each component and solves with auxiliary equations.

$$\sum_{out}(c_{out} \cdot \dot{E}_{out})_k + c_{w,k}\dot{W}_k = \sum_{in}(c_{in} \cdot \dot{E}_{in})_k + c_{Q,k}\dot{E}_{Q,k} + \dot{Z}_k \tag{14}$$

The cost balance can be drawn up in terms of the fuel and product formulation as is shown in equation 15 (Lazzaretto & Tsatsaronis, 2006). The exergy destroyed in each component can be calculated with the fuel cost as presented in equation 16, where $\dot{C}_{D,k}$ is the cost of exergy destruction of component *k*.

$$\dot{C}_{P,k} = \dot{C}_{F,k} + \dot{Z}_k - \dot{C}_{D,k} \tag{15}$$

$$\dot{C}_{D,k} = c_{F,k}\dot{E}_{D,k} \tag{16}$$

The cost of each *k* component can be expressed as a cost per unit of time $\dot{Z}_k$ (\$/h), given as equation 17 and 18, where CRF is the capital recovery factor, φ is the maintenance factor, N is the number of operating hours in a year, i is the interest rate and n the system life, assumed to be 20 years (Shokati, Mohammadkhani, Yari, Mahmoudi, & Rosen, 2014).

$$\dot{Z}_k = \frac{Z_k \cdot CRF \cdot \varphi}{N} \tag{17}$$

$$CRF = \frac{i \cdot (1+i)^n}{(1+i)^n - 1} \tag{18}$$

The exergoeconomic assessment is accomplished using exergoeconomic factor expressed in equation 19; with the parameters of equipment cost expressed in cost per time and the cost flow of exergy destruction. If f_k is high, it is necessary an investigation to reduce the capital investment for the *k* component, if f_k is low, it is necessary to look for an improvement in the efficiency of the component increasing the capital investment.

$$f_k = \frac{\dot{Z}_k}{\dot{Z}_k + \dot{C}_{D,k}} \tag{19}$$

The market price of the feedstock must be defined (i.e. leaves, pseudostem), however, as the resource is a production waste, it does not have a cost or a monetary trend established in the market. Moreover, the pyrolysis plant is being evaluated onsite the production of banana in Urabá, Antioquía, Colombia, thus the feedstock will not have a cost, as other studies establish it (Osorio & Rubiano, 2019). Table 4 summarizes the necessary data for the economic analysis. The annual operating hours are assumed to be 5760 due to is the common time for a continuous plant (Kohl et al., 2015; Torres et al., 2020).

Table 4. Economic data

Parameter	Symbol	Units	Value
Operating hours	N	h	5760
Maintenance factor	φ	-	1.10
Interest rate [a]	i	%	1.750
Electricity tariff [b]	-	\$/kWh	0.14

Note: [a] *Central Bank of Colombia,* [b] *Enel* (ENEL, 2020)

3. Results and Discussion

3.1. Mass and energy balances

The material and energy flows for the overall plant are summarized in Tables 5 and 6. Results were obtained from the simulation in Aspen Plus®. The air stream was calculated according to the requirement for drying and for burning the char and the non-condensable gases. As a part of the non-condensable gas was used to provide the heat required for the pyrolysis, just the part that was not employed is shown as an output. The mass balance evidences that from 1,446.92 kg/h feedstock the process obtained 156.06 kg/h of bio-oil, which a high amount is taking into account that in the drying process the feedstock was reduced to 388.26 kg /h on dry basis. The exhaust gas is a mix of outlet gases from combustion (1,659.26 kg/h) and the NCG (1,066.00 kg/h) which both can be used for energy generation applications with a combined heat and power (CHP) plant (Sahoo et al., 2020). It should be noted that the reactor presents a significant loss of energy, even though the char and NCG provided the energy, it was unavoidable to present energy losses. Due to this loss, that occurs when the feedstock is converted into a conventional component in the simulator, bio-oil presents a low energy content, besides the energy lost the condenser could infer in the result as other studies presented a quenched system where the bio-oil was recirculated and sprayed to the hot gases (Lv et al., 2018). Another significant output stream of energy is the wet air from the drying process, which, even though, is a high amount of energy the low temperature (47.5°C) make the stream useless for another heating process and also the spend is no worthy.

Table 5. Summary mass balance.

Stream	Input (kg/h)	Output (kg/h)
Feedstock	1,446.92	
Air	67,832.62	
Nitrogen	1,048.30	
Coolant	10,000.00	
Cool Air	10,000.00	
Wet Air		77,242.82
Exhaust gas		2,725.26
Cool Water		9,827.54
Bio-oil		156.06
Ash		31.24
Total	90,327.84	90,327.84

Table 6. Summary energy balance.

Sink	Input (kW)	Output (kW)
Feedstock	4.499.80	
Heat Air	462.41	
Compressor	753.99	
Wet Air		3,527.07
Dryer losses		601.35
Reactor losses		423.09
Exhaust gas		391.01
Non-Condensable Gases		83.68
Bio-oil		63.42
Ash		6.90
Total	5,096.54	5,096.54

3.2. Model validation

Simulation results were validated against literature values to have an idea of the accuracy of the model as shown in Table 7, the experimental works were performed at similar conditions to those of the simulated model. The simulative results are closer to Ozbay et.al.'s (Ozbay et al., 2019) results than the others authors, this can be attributed to differences in the reactor type, experimental set-up and feedstock source. However, there were no errors during the simulation, demonstrating that the model is viable.

Table 7. Validation Scheme

	Current Study	Ozbay et al. (Ozbay et al., 2019)	Basu (Ghosh et al., 2019)	Hussain et al. (Sellin et al., 2016)
Bio-Oil	40.20%	48.5%	52%	27%
Char	24.75%	28%	30%	23.3%
Gas	35.07%	23.5%	18%	49.6%

3.3. Exergy analysis

Table 8 shows the result of exergetic analysis for each component in the system. Each component is valued based on the percentage of exergy destroyed. The overall exergetic efficiency of the plant is 61.76%. The dryer accounts for 35.84% and 50.06% of the total exergy destruction and exergetic efficiency, respectively. The low efficiency is due to the irreversibilities produced by the mixture of the biomass with the dry air and the evaporation of the water contained in the biomass. The pyrolysis process accounts for 32.96% of the total exergy destruction, this is mainly caused by the irreversible thermal chemical conversion reaction in the reactor. The product distribution and exergetic efficiencies can be related and the operating conditions could be investigated (Lv et al., 2018). The pyrolysis efficiency was of 45.61%, similar to a pyrolysis studio with efficiency of 57.3% in the reactor (J. F. Peters et al., 2014), the difference may be related with type of reactor, feedstock and operational conditions. The input exergy of the pyrolysis is destroyed due to the heat lost and the variety of concentrations of the streams (Torres et al., 2020). The compressor only represents 1.26% of the total exergy destruction, the potential for further is reducing their electricity consumption by more efficient equipment is related with high costs

(Petrakopoulou, Tstsaronis, Morosuk, & Carassai, 2012). The condenser and the combustion chamber present similar exergy destructions, 9.05% and 10.41%, respectively. Although, the combustion chamber has a better efficiency than the condenser, this could be for the non-condensable gases that exit the condenser. Cyclones have the highest exergetic efficiency as much as 98%, thus the exergy destroyed is the lowest from the plant. Compared with other pyrolysis studies the efficiency of the plant (61.76 %) is similar to other studies of fast pyrolysis with a total efficiency of 68.17 % (Lv et al., 2018), however compared with another study about a combined plant of heat and power that integrated a pyrolysis plant was 46 % due to the other elements that the plant has apart from the pyrolysis system.

Table 8. Results of exergy analysis

Component	$E_F(kW)$	$E_P(kW)$	$E_D(kW)$	$E_D(\%)$	$\eta_E(\%)$
Dryer	759.23	380.09	379.15	35.84%	50.06%
Pyrolysis	641.03	292.35	348.68	32.96%	45.61%
Cyclone	99.91	97.69	2.22	0.21%	97.78%
Compressor	134.32	120.95	13.37	1.26%	90.04%
Combustion Chamber	359.14	249.07	110.07	10.41%	69.35%
Heat Exchanger	256.46	208.11	48.35	4.57%	81.15%
Cyclone2	140.87	138.27	2.60	0.25%	98.15%
Condenser	235.69	139.96	95.73	9.05%	59.38%
Cooling Tower	139.59	82.01	57.58	5.44%	55.40%
Total	2,766.25	1,708.50	1,057.75	100.00%	61.76%

3.4. Exergoeconomic analysis

The investment cost of the plant is presented in Table 9. The capacity of the components was obtained according to the data from the simulation, those cost were carried out as explained above with equations 12 and 13. The cost of the cooling tower was obtained from the Aspen Process Economic Analyzer (APEA) from Aspen Plus®. The total Purchased Equipment Cost (PEC) was $2,613.68 kUSD, and according with the cost estimation for Fixed Cost (FC) and operational and maintenance cost, the total cost investment was $4,809.16 kUSD. Comparing the investment with other authors could present variations due to the different equipment and factors employed, but scaling the capacity of the present study to compare it with Kun et al. (Kun, He, Guan, & Zhang, 2018), the total capital investment should be higher ($6,510.00), but that difference could be attributed that this report took into account high factors for installation cost, contingency and also cost of buildings.

Table 9. Investment cost calculation for the plant.

Equipment	Capacity	CEPCI [c]	Scale factor, j	Cost (kUSD$)
Dryer [a]	1.50 ton/h	678.8	0.67	150.06
Pyrolysis reactor [a]	2.00 ton/h	678.8	0.65	783.02
Cyclone [b]	1.50 ton/h	593,1	0.60	74.76
Compressor [a]	0.14 MW	678.8	0.85	375.28
Combustion Chamber [a]	800.00 kW	678.8	0.60	586.58
Heat exchanger [a]	0.15 MW	678.8	0.60	217.02
Cyclone 2 [b]	1.70 ton/h	593.1	0.60	80.59
Cooler [a]	0.25 MW	678.8	0.60	294.86
Cooling Tower				51.50
Total PEC				*2,613.68*
FC	Cost estimation (% of PEC)			
Installation cost	35			914.79
Contingency	25			653.42
Total FC				4,181.88
OP	Cost estimation (% of FC)			
Maintenance	10			418.19
Personnel	5			209.09
Total OP				627.28
TCI	***FC+OP***			***4,809.16***

Note: taken from [a] Kun et al. (Kun et al., 2018) *(2015), [b] Wright et al.* (M. M. Wright et al., 2010) *(2018), [c] the index of each year*.

The exergoeconomic analysis was carried out according to each equipment of the plant. Data of the analysis is presented in Table 10 with indicators such as specific fuel cost ($\dot{c_F}$), the destruction cost rate ($\dot{C}_D$), the cost of the unit in terms of time ($\dot{Z}_k$), and the exergoeconomic factor (f). To develop the analysis model, the specific exergy costing (SPECO) method was used to calculate the cost of each stream of the system, also electricity cost data was required to determine the work cost of some equipment. Each value of these costs were expressed in terms of specific fuel cost ($\$/kJ$). These costs let us know how much of the exergy was consumed to obtain such exergy in each stream, and as the higher of the exergetic costs, the lower the thermodynamic efficiency of the process (Bejan, Tsatsaronis, & Moran, 1996), results show that the combustion chamber (4.55×10^{-4}), the cooling tower (4.69×10^{-4}), and the condenser (3.35×10^{-4}) are the equipment with the lower thermodynamic efficiency due to their specific fuel costs.

In order to identify which equipment should be optimized, the exergoeconomic factor is used. A high value of the factor indicates that the cost of the unit should be reduced even if the efficiency is also reduced. A low value of the factor indicates that the cost of the unit should increase in order to increase the efficiency. As was mention above, the units with the higher specific fuel cost also present a low *f* value which confirms the necessity of the efficiency improvement in these equipment. For that reason, it is undoubted that the exergetic efficiency of these units should be improved in order to reduce the overall cost of the system. Furthermore, the dryer also presents a low *f* value even though the specific fuel cost is also low, this is because the dryer is the unit with the most exergy destruction. In general terms, the low value indicates that the cost of destroying the exergy is considerably higher than the cost and operation of the unit. Although, in the open literature it could not be found typical values for the equipment employed in the current study to carry out a comparison on the exergoeconomic values, but the analyses made above gives an idea about which components should be studied and improved in order to increase their efficiency. Summarising, the equipment that would need an improvement will also increase its investment cost. As mentioned above, the exergoeconomic analysis allows the evaluation of the impacts of fuels on each equipment of the system and to determine the irreversibilities that are generated in its processes, which is an essential part to understand which are the operations of the plant that would need a deeper evaluation for improve or design in order to achieve better efficiency.

Table 10. Results of the exergoeconomic analysis.

Equipment	$\dot{c}_F(\$/kJ)$	$\dot{Z}_k(\$/h)$	$\dot{C}_D(\$/h)$	$f(\%)$
Dryer	3.89×10^{-5}	2.96	53.08	5.22
Pyrolysis	5.72×10^{-5}	15.25	71.82	17.51
Cyclone	8.06×10^{-5}	1.46	0.64	69.34
Compressor	3.89×10^{-5}	7.31	1.87	79.60
Combustion Chamber	4.55×10^{-4}	11.42	180.19	5.96
Heat Exchanger	4.69×10^{-6}	4.23	0.82	83.80
Cyclone2	2.78×10^{-4}	1.57	2.60	37.61
Condenser	3.35×10^{-4}	5.74	115.47	4.74
Cooling Tower	4.69×10^{-4}	1.00	97.31	1.02

4. Conclusion

A thermoeconomic evaluation of the fast pyrolysis process using banana wastes has been presented. The simulation of the whole process was developed using Aspen Plus®. Software, results have been validated with experimental data. The yield of bio-oil (48.5%) evidences the potential of this product. System components have been analyzed individually. The exergoeconomic analysis has been done to identify the components that should be studied to improve their efficiency. The overall exergetic efficiency is 61.76%. The process of drying and pyrolysis are the ones that contribute the most to the exergy destruction with 35.84% and 32.96%, respectively. Besides, the condenser and the cooling tower presented the lower exergoeconomic factors, whereby it is necessary to increment the investment in order to increment its efficiency. On the whole, this study showed how fast pyrolysis technique with banana waste could be an opportunity of new fuel resources, with good yield results of products. However, to determine the potential for improvements in the plant it is necessary an advanced exergoeconomic analysis to quantify the avoidable and unavoidable economic losses. Besides that analysis, as in this study only one type of plant was presented, in the future would be interesting the analysis of other types of equipment with different characteristics or the relation of operational conditions of the process with energy and exergetic efficiency, and in that way determine the profitable of the plant. According to this, conclude if the plant is profitable would be an early assumption, due to more studies and investigations regarding the efficiencies should be made.

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