Development of the Process Model and Optimal Drying Conditions of Biomass Power Plants

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ABSTRACT: An empty fruit bunch (EFB) is a byproduct of the palm oil production process with an undried moisture content of 60–70%, which is too high for use as direct combustion fuel. Drying processes are generally considered essential for the recent use of EFBs as power plant fuels because their high moisture content decreases the boiler efficiency. The lower moisture content of dried EFBs increases the heating value and boiler efficiency but creates a trade-off with the energy required for the drying process. This study developed an EFB-based 10 MW power plant model by integrating economic evaluations in order to obtain optimal drying conditions. A hot air dryer was used in the drying process. The EFB evaporation behavior was predicted by reflecting the drying kinetics of EFBs in Aspen Plus. The optimum drying conditions were found to be a steam recirculation ratio of 0.25 and drying time of 23 min, creating dried EFBs with a 9.91% moisture content, which reduced costs by 5.48% relative to the undried base scenario. In addition, the developed model was compared to the drying process of a real power plant currently under construction in Indonesia. This drying process reduces the EFB moisture content from 48 to 20%.

1. INTRODUCTION

The use of biomass as power plant fuel is globally well accepted due to fossil fuel depletion and global warming. Indonesia, for instance, has plans to reduce the proportion of fossil fuel power plants by 2025. Various types of biomass will be used as a fuel for power plants, including wood, grain, and municipal solid wastes; however, the most suitable appears to be empty fruit bunches (EFBs), a type of biomass occurring in palm oil production in Indonesia. Because Indonesia is the largest producer of palm oil, it produces most suitable appears to be empty fruit bunches (EFBs), a type of biomass occurring in palm oil production in Indonesia. Because Indonesia is the largest producer of palm oil, it produces 30 million EFBs,2 making the EFB fuel costs lower than other fuels until properly dried, as their typical moisture content (MC) ranges from 60 to 70%.3 Such high MCs cause numerous problems, including lower combustion temperatures and burning stability, higher CO and volatile organic compound (VOC) emissions, and difficult boiler operations. In addition, the boiler efficiency is reduced by increasing heat losses including flue gas, chemical unburned carbon, and mechanical unburned carbon losses.

The multiple disadvantages of using raw EFB biomass as power plant fuel can be solved by removing moisture from the biomass. Drying processes are broadly divided into two categories: mechanical drying, which can reduce moisture by up to 50 wt % through shredding, grinding, pressing, and filtering; and thermal drying with direct or indirect dryer use, which reduces MCs to less than 50 wt %. Thermal drying has large energy and cost requirements because of the high specific heat of moisture. Assuming a reference temperature is 15 °C, the evaporation energy of water per kilogram ranges from 2.48 to 2.57 MJ. Therefore, many studies have been conducted to improve the utilization of solid fuel while minimizing the drying energy.

Primary researches on this subject are most of the experimental studies focusing on the drying kinetic mechanism of biomass and the characteristic of dried products. The mathematical modeling studies on biomass drying make it possible to predict drying behavior, diffusivity, the temperature profile of biomass, and VOCs occurred drying. Several previous studies have reported that the integration of biomass drying and power plants has improved the thermal efficiency of boilers.

Luk et al. compared the energy efficiency when each of hot air dryer (HAD) and superheated steam dryer (SSD) integrates with the biomass power plant. The way of single drying, both HAD and SSD enhanced energy efficiency for condition that was not through drying process, 21.56 to 25.99 and 23.18%, respectively. The overall energy efficiency combined with HAD and SSD was 26.90%. Li et al. studied capital costs...
and operating costs in detail using two types of heat sources as flue gas or superheated steam. The use of flue gas as a drying medium resulted in lower capital costs. However, there was a possibility of pollutant emission. The superheated steam has relatively high capital costs, although it has short drying time, environmental protection, and good heat recovery. Literature provided by the Northwest Combined Heat and Power Application Center suggests that the optimum biomass MC for direct combustion is 10–15%. The boiler efficiency in this range was 80%, which was higher than the 74% when the MC was 45%. Gebreegziabher et al. have developed a mathematical model of drying to optimize the annual profit of process varying with woodchip size. The annual profit for wood chip sizes of 0.0005, 0.005, and 0.05 m were investigated, respectively. Of these, the profit was the highest at 0.0005 m. The above-cited studies presented a useful perspective of integration of the drying process with power plant and thermal efficiency related to the MC during drying. However, they lack proper discussion on the specific topic regarding how optimum drying condition and MC are obtained. Although Gebreegziabher et al. developed the drying model to study the trade-off between the heating value and MC, they focused on operating conditions dependent on drying temperature and particle size while limiting the biomass MC to 17% (minimum calorific value 15 MJ/kg). In addition, the drying kinetics needed to implement these models apply the commonly used Fick’s model of diffusion, not optimized for EFB. In another paper by Gebreegziabher et al. a study was conducted to minimize the cost by heat integrating the drying process developed in the previous paper with a biomass power plant. However, it does not reflect changes in boiler efficiency according to MCs of fuel. Therefore, the objective of this work developed an optimization model to investigate the optimum drying conditions and MC applying the drying kinetics obtained from the EFB drying experimental data and the variation of boiler efficiency with moisture. As shown in Figure 1, this model with integrated drying process and power plant is utilized to examine this trade-off varying with parameter such as drying time or steam recirculation ratio. Applying drying kinetics due to EFB material properties, which is previously reported, to the model allowed prediction of EFB evaporation behavior, temperature, and relative humidity (RH) of air, and a more detailed economic analysis was conducted.

2. PROCESS MODELING

2.1. Process Overview. A 10 MW small-scale biomass power plant under construction in Indonesia was simulated in Aspen Plus. As shown Figure 1, the overall process contains a shredding process, drying process, boiler, and steam cycle. A lower volume of dried EFB than raw EFB is required to generate 10 MW because dried EFB increases the heating value and boiler efficiency. The simulation was designed to vary the amount of EFB used with a target of 10 MW power generation. A description of the process is given below.

1. 60% MC EFB is finely crushed by a shredder to 5 mm pieces, reducing the MC to 48%. The 48% MC EFB enters a hot air rotary dryer.

2. Air entering the rotary dryer raises the temperature via heat exchange with a portion of the steam exiting the turbine. The heated air comes into direct contact with EFBs inside the dryer, evaporating moisture from the material and exchanging heat to reduce the EFB MC to 20%. Ideally, air and EFBs are well mixed. The steam (191 °C, 12 atm) used for drying is fully condensed (188 °C, 12 atm) after heating the air, and the condensed water enters a tank to preheat the feed water (FW).

3. Dried EFBs are burned in the boiler, and the flue gas is discharged at 200 °C. The heat generated by burning EFBs creates 433 °C, 60 atm steam.

4. The high-temperature, high-pressure steam is discharged at 0.107 atm after turning the turbine and producing 10 MW electricity. A portion of the steam exiting the turbine extraction valve is used for drying. The low-pressure steam (0.107 atm) discharged from turbine is fully condensed by the condenser and recirculated in the steam cycle.
2.2. Drying Process. 2.2.1. Dryer Model. Dryers are classified as co-current or counter-current depending on the directions in which the solid and air flow.\textsuperscript{23} Counter-current dryers can produce solids with lower MC; however, the driest solids contact the hottest air, creating a potential fire risk when the dryer operates at high temperatures if the solids are flammable.\textsuperscript{23,24} As biomass generally has a fire risk, this study selected co-current dryers (using a direct rotary dryer and air heater) for investigation. In co-current dryers, the solid and hot air are in direct contact with each other, and the transfer phenomenon moves moisture from the solid to the air. This evaporation process requires a large amount of energy because moisture has a high specific heat. The dryer was assumed to be in a steady state with no heat loss. The dryer unit was modeled in Aspen Plus.\textsuperscript{26}

\begin{align}
X_{i+1} &= X_i - \frac{M \times N_p}{M_i} \times \frac{\Delta \tau}{t} \\
Y_{i+1} &= Y_i - \frac{M \times N_p}{M_i} \times \frac{\Delta \tau}{t} \\
T_{o,i+1} &= T_{o,i} - \frac{Q \times N_p - M \times N_p \times h^v \times \rho \times \xi_{p,s} + X \times C_p,M}{M \times (C_{p,s} + X \times C_{p,M})} \times \frac{\Delta \tau}{t} \\
T_{G,i+1} &= T_{G,i} - \frac{Q \times N_p}{M_i \times C_{p,G}} \times \frac{\Delta \tau}{t} \\
M &= \dot{v} \times \rho \times \beta_G \times A_p \times [Y^* - Y] \\
Q &= \alpha_G \times A_p \times (T_{G,i} - T_o) \\
N_p &= \sum_{i=1}^{n} \frac{x_i \times M_s \times t}{\rho \times \bar{v} \times d_{p,i}} \\
A_p &= \sum_{i=1}^{n} \frac{x_i \times \xi \times d_{p,i}^2}{2}
\end{align}

where $M_{i}$ and $M_{G,i}$ are the mass flow rate solid and gas in the dryer, respectively; $T_{o,i}$ is the inlet temperature of solids; $T_{G,i}$ is the inlet temperature of gas; $h^v$ is the enthalpy of evaporation; $C_{p,s}$ is the specific heat capacity of the dry solids; $C_{p,M}$ is the specific heat capacity of the moisture; $C_{p,G}$ is the specific heat capacity of the dry gas; $X$ is the dry-based MC of the solids; $\dot{v}$ is the normalized drying rate of single particles; $\rho$ is the solids density; $\rho_G$ is the gas density; $\beta_G$ is the coefficient for mass transfer between the surface of the particle and the gas; $Y^*$ is the MC of the gas at adiabatic saturation; $X$ is the MC that gas actually has at the considered position in the dryer; $\alpha_G$ is the heat-transfer coefficient gas and particles; and $x_i$ is the mass fraction of the particles within the class $i$.

2.2.2. Drying Kinetics. The falling rate drying curves, critical MC ($X_{c}$), equilibrium MC, and heat- and mass-transfer coefficients of the EFB were required to develop a drying kinetics model in Aspen Plus. $Q$ and $M$ were obtained from the drying kinetics model.

The drying curve includes several stages. First, the solid is heated by a heat source such as hot air. Next, moisture on the surface of the solid evaporates during the constant rate drying period. This drying rate remains the same regardless of the material under constant drying conditions because the moisture on the surface of the material evaporates. When surface moisture has been removed, moisture inside the solid evaporates during the falling rate drying period. The point when the rate changes from the constant rate drying period to the falling rate drying period is called the critical MC. As drying continues, moisture in the dry air reaches an equilibrium with moisture in the solid, after which no more moisture evaporates. This point is called the equilibrium MC. Because the particle structure inside the material affects the drying rate during the falling rate drying period, different materials have different drying curves in the falling rate drying period. To more accurately predict the evaporative behavior of EFBs, an EFB falling rate drying curve was modeled in Aspen Plus.\textsuperscript{20}

\begin{equation}
\dot{v} = \left[1 - \left(1 - \frac{X}{X_{c}}\right)^{1.506}\right]^{1.237}
\end{equation}

where $X$ denotes the normalized MC. The critical MC depends not only on the material and shape but also on the velocity and temperature of the drying air. Because many factors affect the critical MC, the mean values of critical MC shown in Table 1 were used.\textsuperscript{26} The equilibrium MC is influenced by RH and temperature of surrounding air and usually approaches zero at high temperatures and low RH.

| no. | $X_c$ | $T$ (°C) | $V$ (m/s) |
|-----|-------|----------|-----------|
| T1  | 0.84  | 200      | 0.89      |
| T2  | 0.62  | 220      | 0.89      |
| T3  | 0.68  | 240      | 0.89      |
| T4  | 0.7   | 260      | 0.89      |
| T5  | 0.63  | 280      | 0.89      |
| V1  | 0.7   | 200      | 0.79      |
| V2  | 0.77  | 200      | 0.89      |
| V3  | 0.82  | 200      | 0.99      |
| V4  | 0.95  | 200      | 1.09      |
| AVG | 0.745 |          |           |

Heat- and mass-transfer coefficients are calculated using eqs 10–17.\textsuperscript{25,27,28} The diffusion coefficient vapor in the gas ($\delta_{g}$), dynamic viscosity of the gas ($\eta_g$), thermal conductivity of the gas ($\lambda_g$), and the gas density used to calculate the Lewis (Le), Schmidt (Sc), Reynolds (Re), and Sherwood (Sh) number were obtained through Aspen Plus stream analysis. The diameter in
the original mass-transfer coefficient and Reynolds number equation was replaced with sauter mean diameter \((d_{32})\) and median particle diameter for EFBs, respectively. As shown in eqs 10–17, the heat, mass-transfer coefficient, and evaporation are improved by higher drying temperatures, higher velocities, and smaller particle sizes. The air velocity in the dryer \((u_G)\) was typically 0.5–2.5 m/s, and the mean value was used to calculate the Reynolds number.

\[
\beta_G = \frac{Sh \times \delta_G}{d_{32}}
\]

\[
\alpha_G = \beta_G \times \rho_G \times C_{p,G} \times Le^{1-n}
\]

\[
Sh = \sqrt{Sh_{\text{laminar}}^2 + Sh_{\text{turbulent}}^2}
\]

\[
Sh_{\text{laminar}} = 0.664 \times \sqrt{Re} \times \sqrt{Sc}
\]

\[
Sh_{\text{turbulent}} = \frac{0.037 \times Re^{0.8} \times Sc}{1 + 2.443 \times Re^{-0.1} \times (Sc^{2/3} - 1)}
\]

\[
Re = \frac{u_G \times d_p \times \rho_G}{\eta_G}
\]

\[
Sc = \frac{\eta_G}{\delta_G \times \rho_G}
\]

\[
Le = \frac{C_{p,G} \times \rho_G \times \delta_G}{C_{\lambda}}
\]

2.2.3. Air Heater. As shown in Figure 1, the air exchanges heat with the portion of steam exiting the turbine, with the heat of condensation in the steam heating the air. Air and steam flow counter-currently to remove the maximum amount of heat from the steam. Steam (191 °C, 12 atm) is completely condensed and becomes water (188 °C, 12 atm). The temperature of the hot air leaving the heat exchanger (HEX) cannot exceed 191 °C, the temperature at which the steam enters.

The temperature of the air exiting the heat exchanger depends on the air flow rate and can range between 30 and 191 °C. EFB MC changes over time were simulated, reflecting the drying kinetics. Figure 2 shows the simulation results and indicates that hotter air required less time to produce a lower-moisture-content EFB. Therefore, the most efficient drying air temperature is highest in the possible temperature range.

Although the temperature range peaked at 191 °C, the highest possible air temperature was set at 181 °C with consideration of typical minimum temperature difference. The dry air flow rate was designed to vary with the steam recirculation ratio such that increasing the steam recirculation ratio would increase the dry air flow rate to match 181 °C.

2.3. Combustion Process. 2.3.1. Combustion Model. The EFB heating value, which varies with the MC, is important when calculating the amount of EFB required to generate 10 MW power. As it was impossible to obtain heating values for all MCs in this experiment, heating values were estimated for different MCs using the Aspen Plus combustion model. The combustion process was modeled using the Rgibbs reactor. In an Rgibbs reactor, combustion reactions are conducted to minimize the Gibbs free energy. The experimental results of proximate and ultimate EFB analysis shown in Table 2, calculated using EFBs imported from Indonesia, are required to model biomass combustion reactions. If an EFB heating value corresponding to an arbitrary MC was specified in the model, the EFB heating value at a different MC could be obtained by burning EFB in the model. The experimental in Table 2 was used to specify the heat of combustion (heating value) at an arbitrary MC (20%). The model-estimated higher heating value (HHV) of EFB is plotted as black lines in Figure 3, and the experimentally obtained HHV of EFB is represented by the gray dot. As shown in Figure 3, the combustion model is valid because the difference between the model and experimental results was insignificant.

2.3.2. Boiler Efficiency. The HHV of EFB obtained from the combustion model represented the heat generated when EFBs were completely burned; however, during actual combustion the high MC lowers the combustion temperature, causing

**Table 2. Proximate and Ultimate Analysis EFB Properties**

| proximate analysis (wet basis) | ultimate analysis (dry basis) |
|-------------------------------|-------------------------------|
| volatile matter 67.4 wt %     | C 45.9 wt %                  |
| fixed carbon 9.8 wt %         | H 6.1 wt %                   |
| ash 2.8 wt %                  | N 0.43 wt %                  |
| moisture 20 wt %              | S 0.13 wt %                  |
| O 47.44 wt %                  | HHV (MC = 20%) 15.45 MJ/kg   |

Figure 2. EFB drying curves at various air temperatures.

Figure 3. HHVs according to MC.
incomplete combustion. This results in several thermal losses and reduces the amount of heat transferred to produce high-temperature, high-pressure steam. This study applied known boiler efficiencies according to the MC of wood to the process model. For flue gas at 200 °C, the actual heat transferred considering flue gas loss and chemical and mechanical unburned carbon losses could be presented as

\[ \eta_{\text{boiler}} = -0.0025x_\text{m}^2 + 0.0145x_\text{m} + 84.612, \]

\[ R^2 = 0.9941 \] (18)

where \( \eta_{\text{boiler}} \) and \( x_\text{m} \) denote the boiler efficiency and the MC of wood, respectively.

2.4. Steam Cycle. 2.4.1. Steam Cycle Model. The steam cycle process shown in Figure 1 shows that 47 °C, 2 atm FW is raised to 12 atm by pump 1 and enters the FW preheat tank, where it is preheated by completely condensed water from the air heat exchange. The preheated water is then raised to 67 atm by pump 2, and this high-pressure water enters the boiler where it is converted to steam (433 °C, 60 atm). The high-temperature, high-pressure steam then turns the turbine, and a portion is recirculated to heat the air. After 10 MW electricity is produced, the steam exiting the turbine is discharged at 47 °C, 1 atm) for this heat. Similarly, with 200 °C, air 2 flow rate satisfying 10 MW power generation. Assuming depreciation over the 20 year lifespan of a power plant, AEC were estimated by eq 20. The AOC were calculated by eq 21, multiplying the EFB fuel cost by the annual number of operating days and the EFB (MC, 60%) mass flow rate (\( m_{\text{EFB,60%}} \)). The fuel price was contracted with local Indonesian EFB suppliers as 14.08 $/ton.

\[ \text{AEC} = \left( C_{\text{dryer}} + C_{\text{shredder}} + C_{\text{furnace}} + C_{\text{hex}} \right) / 20 \] (20)

\[ \text{AOC} = 14.08 \, \text{$/ton} \times m_{\text{EFB,60%}} \times 24 \times 320 \] (21)

\[ \text{ATC} = \text{AEC} + \text{AOC} \] (22)

where \( C_{\text{dryer}}, C_{\text{shredder}}, C_{\text{furnace}} \), and \( C_{\text{hex}} \) are the equipment cost of dryer, shredder, furnace, and heat exchanger, respectively. Because AEC are influenced by various factors such as usage, capacity, and materials, they must be roughly estimated. The total capital investment in the power plant neglected labor costs, installation costs, transportation costs, and pipeline installation costs but included the cost of equipment affected by the amount of EFB: the dryer, shredder, heat exchanger, and furnace. Turbines were not considered because they remained the same at 10 MW power generation, and details such as pump blowers were ignored. The equipment cost estimate equation for dryers, shredders, and heat exchanger was developed in 2003, whereas the furnace cost equation was developed in 2002. Equations 27–29 reflect the cost index in 2017.

The dryer cost was influenced by the lateral surface area of the dryer (\( A_{\text{dryer}} \)). If the EFB (MC, 48%) mass flow rate (\( m_{\text{EFB,48%}} \)) and drying time are known prior to dryer area estimates, the dryer volume (\( V_{\text{dryer}} \)) can be estimated. It was assumed that EFBs would occupy 20% of the dryer volume with a density of 500 kg/m³. If the EFBs do not shrink during the drying process, the dryer volume occupied by EFBs is represented by eq 23. The volume of solid in the dryer (\( V_s \)) is affected by time as well as the EFB feed rate.

\[ V_s = \frac{m_{\text{EFB,48%}}}{\rho} \times t \] (23)

When the dryer operates at a 20% load, the dryer volume is calculated by eq 24. The \( L/D \) ratio of the dryer should be known to determine the area of the dryer. As typical rotary dryers have a length to diameter (\( L/D \)) ratio of 4–10. The selected \( L/D \) ratio in this study was 7, the average value. The dryer diameter can be expressed by a volume equation such as eq 25, after which the area is calculated by eq 26.

\[ V_{\text{dryer}} = \frac{V}{0.2} \] (24)

\[ D^3 = \frac{4V_{\text{dryer}}}{7\pi} \] (25)

\[ A_{\text{dryer}} = 7\pi D^2 \] (26)
The type of heat exchanger was selected as air-cooled (finned-tube), which raises the air temperature entering the dryer. The heat exchanger cost is a function of area ($A_{\text{hex}}$), which can be calculated as eq 29. $T_{\text{LM}}$ is a log mean temperature difference, which is obtained from inlet and outlet temperature of the fluids when being heat exchanged. The overall heat-transfer coefficient ($U$) was typically $700–850 \text{ W/m}^2\text{K}$, and the mean value was used.\(^{34}\) The cost of heat exchanger was represented by eq 30.

\[
Q = UA_{\text{hex}}T_{\text{LM}}
\]

\[
C_{\text{hex}} = 6844A_{\text{hex}}^{0.4}
\]

Only combustion was considered in furnace costs, ignoring detailed costs such as piping, steam drums, soot blowers, fans, deaerators, and pumps included in the boiler. Furnace costs varied according to the heat-transfer rate ($H$) and were estimated using eq 31. Finally, we summarize the parameters for the objective function in Table 3.

\[
C_{\text{furnace}} = 247H^{0.88}
\]

### Table 3. Summary of Parameters for the Objective Function

| Parameter | Unit | Value |
|-----------|------|-------|
| $\rho$ | kg/m$^3$ | 500 |
| $V_i$ | m$^3$ | $m_{\text{EFB,60\%}} \times \frac{t}{\rho}$ |
| $V_{\text{dry}}$ | m$^3$ | $V_i/0.2$ |
| $D$ | m | $(4V_{\text{dry}}/\pi)^{1/3}$ |
| $A_{\text{dry}}$ | m$^2$ | $7\pi D^2$ |
| $C_{\text{dry}}$ | $\$ | $8055A_{\text{dry}}^{0.63}$ |
| $C_{\text{furnace}}$ | $\$ | $247H^{0.86}$ |
| $T_{\text{LM}}$ | K | 53.6 |
| $U$ | W/m$^2$ K | 775 |
| $A_{\text{hex}}$ | m$^2$ | $Q/UT_{\text{LM}}$ |
| $C_{\text{hex}}$ | $\$ | $6844A_{\text{hex}}^{0.4}$ |
| $C_{\text{shredder}}$ | $\$ | $4125m_{\text{EFB,60\%}}^{0.78}$ |
| EFB cost | $\$/ton | 14.08 $ |
| AEC | $\$/year | $(C_{\text{dryer}} + C_{\text{shredder}} + C_{\text{furnace}})/20$ |
| AOC | $\$/year | $(14.08 \times \$\$/ton) \times m_{\text{EFB,60\%}} \times 24 \times 320$ |
| ATC | $\$/year | AEC + AOC |

#### 3.2. Case Study Description

Case studies were conducted to optimize power of 10 MW using the objective function of the process model mentioned above. The most efficient (case A) and most inefficient (case B) conditions among various circumstances, which evaporate the moisture of EFBs to 20%, were compared to save ATC of real power plant currently being constructed in Indonesia, which reduce the MC of EFBs from 48 to 20%. Furthermore, given only the most efficient drying conditions which satisfy several MCs such as 15, 25, 30, and 35%, the optimal point was investigated. Herein, case 1, case 2, case 3, and case 4 corresponded to the extent of drying performed to MC 15, 25, 30, and 35%, respectively. Measurements were carried out by altering the steam recirculation ratio ($0–0.43$) and drying time ($1–60 \text{ min}$) to identify the optimum combination. In all cases, EFBs (MC, 60%) were processed by the shredder into 5 mm particles with 48% MC. The base case, in which EFB from the shredder was used as fuel without drying, produced 10 MW of electricity.

### 4. RESULTS AND DISCUSSION

The results were derived from the case studies of designed model. Figure 4 shows a contour image of the annual cost required to generate 10 MW power when the steam recirculation ratio and drying time are varied. Figure 5 represents the MC of the dried EFB in the same range. Table 4 summarizes the detailed results for several cases as well as cost savings. Besides, the heat and mass balance results for the cases are provided in Figures S1 and Tables S4–S11 of the Supporting Information. In the base case, the hot air mass flow rate is zero; this is because the steam recirculation ratio used for drying is zero. To produce 10 MW of electric power, approximately 19,784 kg/h of 60% MC EFB was required; however, the shredder produced 15,219 kg/h. Without considering the energy used in the drying process, the amount of work required to operate the steam cycle was 71 kW.

The case A, case B, dots, and triangles in Figure 4 show the points where the MC of the dried EFB reached 20%, indicating that there are many drying conditions capable of satisfying the desired EFB of 20% MC. If you see Figure 4 with Figure 5, it
helps you to understand easily. The triangle indicates that the relative air humidity reached nearly 100%, preventing efficient moisture evaporation even for prolonged drying times. Although the triangle with a shorter drying time has higher MC than those with longer ones, these differences may be disregarded and the MCs considered as 20% because the discrepancies are only one decimal point different. Among these, cases A and B represent the most efficient and most inefficient conditions, respectively. The MCs of cases A and B were approximately equal to those of the other conditions but demonstrated large differences in cost savings. This indicates that the drying method is more important than the amount of drying. 

Cases 1, 2, 3, and 4 represent the most efficient conditions satisfying MC requirements of 15, 25, 30, and 35%, respectively. In Figure 4, the star represents the optimal point of the annual cost graph. At this optimal point, the steam recirculation ratio, drying time, and MC of the dried EFBs were 0.25, 23 min, and 9.91%, respectively. Table 4 shows the data for each case. As shown by the optimum cost savings points in cases 1, 2, 3, 4, and A, drying EFBs to the optimum point incurred the least cost. However, there were no significant differences in cost savings compared to optimal EFB drying conditions at 10–20% MC if only the most efficient drying conditions were considered. The differences at over 20% MC were significant.

5. CONCLUSIONS

An EFB-based 10 MW power plant process model created in Aspen Plus was used to optimize the drying process. The model considered the drying kinetics according to EFB material properties. The $\dot{Q}$ and $\dot{M}$ values calculated from the drying kinetics were used to obtain the dryer material and energy balance as well as the EFB evaporation behavior over time. This model has a trade-off among steam, drying time, capital investment, and operating costs for drying. The optimal drying conditions were found to be 0.25 ratio, 23 min, and 9.91%, respectively. The optimization graph shows that approximately equal MCs can be obtained under different conditions; however, the cost savings differ. This indicates that the drying method is more important than the extent of drying performed. It was further observed that the use of 20% MC EFB is viable when only optimal operating conditions are considered for drying because the cost is not significantly reduced.

**Table 4. Results of the Case Studies**

| parameter | unit | base | optimum | case A | case B | case 1 | case 2 | case 3 | case 4 |
|-----------|------|------|---------|--------|--------|--------|--------|--------|--------|
| drying time | min | 0 | 23 | 13 | 4 | 17 | 9 | 5 | 4 |
| steam recirculation ratio | | | | | | | | | |
| wet EFB (60%) | kg/h | 19,822 | 18,264 | 18,464 | 19,501 | 18,346 | 18,610 | 18,610 | 18,819 | 19,041 |
| wet EFB (48%) | kg/h | 15,248 | 14,049 | 14,202 | 15,001 | 14,113 | 14,316 | 14,476 | 14,476 | 14,647 |
| EFB flow rate | kg/h | 15,248 | 8109 | 9339 | 9824 | 8711 | 9999 | 10,754 | 11,835 |
| evaporation amount | | | | | | | | | |
| wet EFB temperature | °C | 30 | 30 | 30 | 30 | 30 | 30 | 30 | 30 |
| dried EFB temperature | °C | 48 | 9.91 | 48.75 | 48.00 | 48.82 | 47.49 | 44.37 | 43.16 |
| hot air flow rate | m³/h | 125.72 | 102.47 | 163.42 | 113.92 | 91.35 | 80.55 | 59.87 |
| inlet air temperature | °C | 181 | 181 | 181 | 181 | 181 | 181 | 181 | 181 |
| outlet air temperature | °C | 50.45 | 48.75 | 89.03 | 49.52 | 48.45 | 50.81 | 46.90 |
| RH of hot air | % | 87.44 | 95.46 | 12.84 | 91.80 | 96.63 | 84.71 | 99.90 |
| turbine work | kW | 10,071 | 10,117 | 10,109 | 10,131 | 10,113 | 10,105 | 10,101 | 10,093 |
| other work | kW | 71 | 117 | 109 | 131 | 113 | 105 | 101 | 93 |
| net power | kW | 10,000 | 10,000 | 10,000 | 10,000 | 10,000 | 10,000 | 10,000 | 10,000 |
| dryer cost | $/year | 0 | 32,892 | 25,971 | 16,215 | 29,003 | 22,327 | 17,533 | 16,035 |
| Shredder cost | $/year | 2119 | 1989 | 2001 | 2092 | 1993 | 2013 | 2033 | 2049 |
| heat exchanger cost | $/year | 0 | 2312 | 2136 | 2520 | 2225 | 2044 | 1948 | 1742 |
| furnace cost | $/year | 89,164 | 100,123 | 98,171 | 102,731 | 99,133 | 97,238 | 96,329 | 94,585 |
| AEC | $/year | 91,283 | 137,315 | 128,280 | 123,588 | 132,354 | 123,622 | 117,843 | 114,411 |
| AOC | $/year | 2,143,540 | 1,974,962 | 1,996,555 | 2,108,756 | 1,983,872 | 2,012,447 | 2,034,990 | 2,058,988 |
| ATC | $/year | 2,234,823 | 2,112,277 | 2,124,835 | 2,232,314 | 2,116,225 | 2,136,069 | 2,152,833 | 2,173,400 |
| cost savings | % | 5.48 | 4.92 | 0.11 | 5.31 | 4.42 | 3.67 | 2.75 |

## ASSOCIATED CONTENT

### Supporting Information

The Supporting Information is available free of charge at https://pubs.acs.org/doi/10.1021/acsomega.9b03557.

Specification summary of the streams and units in the process and parameters for the calculation of mass- and heat-transfer coefficients (PDF)

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2817

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Notes

The authors declare no competing financial interest.

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