



Technoeconomic analysis of co-hydrothermal carbonization of coal waste and food waste

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Abstract

The aim of this research was to evaluate the technoeconomic prospect of hydrochar production through co-hydrothermal carbonization of coal waste (CW) and food waste (FW). A process flow diagram was developed that considered seven reactors, six pumps, and other necessary equipment for producing 49,192 kg/h hydrochar. Three different cases were considered for the economic analysis. Case II considered both CW and FW transportation cost while cases I and III considered only FW and only CW transportation, respectively. The economic analysis revealed the break-even costs to be \$62.24 per ton for case I, \$69.90 per ton for case II, and \$60.26 per ton for case III. The fixed capital investment (FCI) was \$11.4M for all the cases while total capital investment (TCI), working capital (WC), and manufacturing costs were higher for case II compared to cases I and III. A sensitivity analysis examined the effect of nine different variables on the break-even cost. The raw materials' cost as well as their transportation costs significantly affected the corresponding break-even cost. Additionally, increasing the hydrochar production capacity has drastically decreased the break-even cost. However, the analysis also revealed that excessive increase of production capacity can have negative impact on the process economics.

Keywords Food waste · Coal waste · Co-hydrothermal carbonization · Technoeconomic analysis · Sensitivity analysis

1 Introduction

Coal is one of the major fossil fuel energy sources which contributed almost one fourth of the total energy demand of the world in 2011 [1, 2]. This demand is predicted to increase by 17.6% by 2040 due to huge population growth [3, 4]. However, coal extraction is not an efficient process as almost

30–40% of total extracted coal cannot be used as energy source due to high ash, high sulfur, and high heavy metal concentration [4, 5]. This waste portion of the extracted coal is known as coal waste and possesses serious environmental threat such as acid mine drainage, water contamination, and low fertility of land [4]. Turning this huge amount of coal waste into an energy source will go a long way to meet the increasing energy demand. Now, to use coal waste in energy production, it is necessary to choose another suitable energy source that can help to reduce the harmful effects of coal waste. Biomass resources can help in this regard as this resource is abundant and harmful effect on the environment is much lower than fossil fuel [4]. The USA generated 262.4 million dry tons of municipal solid waste (MSW) in 2015, and food waste was 15.1% [6]. However, low energy content will have adverse effect on the overall fuel property [7]. A pretreatment process can be a possible solution to this issue.

Hydrothermal carbonization (HTC) is a pretreatment process where the biomass goes through a conversion process under high temperature (180–260 °C) and high pressure for 5–30 min [8, 9]. Water acts as a reaction medium for this process due to its non-polar solvent characteristic at such high temperature [10]. When HTC process is performed on two

Highlights • Technoeconomic prospect of co-HTC of coal waste and food waste were evaluated.

- Break-even cost varied within \$62.24, \$69.90, and \$60.26 per ton for three separate cases.
- Sensitivity analysis considered nine parameters to analyze the effect on break-even cost.
- Raw material purchasing and transportation cost were key factors in economic analysis.

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different biomass in the same reactor at the same time, the process is called co-hydrothermal carbonization or co-HTC. Co-HTC process can upgrade the fuel properties of coal as well as coal waste and can reduce the concentration of toxic elements (e.g., sulfur, chloride) [4, 11]. Co-HTC process can convert the heterogenous feedstocks into a homogenous mix as well as can affect the chemical structure which helps improving the fuel quality [12–14]. When coal waste was treated with food waste at 230 °C for 30 min, the HHV and elemental carbon percentage increased to a maximum value of 23.0 MJ/kg and 49.5%, respectively [11]. Additionally, the co-HTC treatment decreased the sulfur content from 8.5 to 1.4% and leached significant amount of chloride. Also, the combustion behavior analysis of the co-HTC hydrochar showed a similar trend as the raw bituminous coal meaning that it has the potential to be used with coal in coal-fired power plants [11].

Researchers have experimentally proven that hydrothermal carbonization can be utilized successfully to upgrade the fuel quality of both coal and biomass [4, 15–17]. However, it is essential to analyze the technoeconomic feasibility of co-HTC process in upgrading fuel sources. The feasibility check should consider two different aspects of the co-HTC process—the energy required for the process and the process economics. Both factors are important in case co-HTC process is considered for a scale-up.

Although process energy and economic feasibility of co-HTC is scarce in literature, several attempts have been reported for HTC. For instance, McGaughy et al. [18] performed a process simulation for HTC of one-ton food waste per day at three different temperatures. Additionally, an energy calculation was done to compare the energy duty of the process with respect to the drying process. Results showed that the energy output to input ratio (EOIR) for HTC process at 230 °C was 3.95 while EIOR for only drying the feedstock was 2.22 [18]. A similar study by Zhao et al. [19] revealed that a complete HTC process of sewage sludge at 200 °C needed for only 42% of the total energy input required by the drying process. Additionally, 47.6% of total process energy output was available for external usage after covering all utility requirement of the process plant [19].

Kempegowda et al. [20] studied the technoeconomic analysis of electricity generation from HTC of wet biomass and concluded the process both economically and technically feasible with the cost of electricity of 0.2–0.4 \$/kWh. The integrated plant showed an efficiency of 21–40% depending on the HTC process parameters [20]. Saari et al. [21] performed a similar study in case of integrating hydrothermal carbonization in a wood-fired combined heat and power (CHP) plant and reported that combining HTC with a CHP plant resulted in a significant cost reduction and offered longer plant operating time [21]. Li et al. [22] made a comparative study of pyrolysis, anaerobic, and HTC process to convert rice husk to

energy and concluded HTC to be the economical depending on lower utility cost and higher solid loading.

Lucian et al. [12] modeled the process design of HTC of grape marc and compost containing 65% and 30% moisture, respectively. The HTC was performed at 180, 220, and 250 °C for 1, 3, and 8 h to determine to optimize operating condition and study the economic analysis of the process. Two hundred twenty degrees Celsius reaction temperature and 1 h of reaction time were revealed to be the optimum operating condition where the thermal and electrical energy consumption was 1.17 and 0.16 kWh per kg of hydrochar, respectively [12]. Additionally, the production cost and the break-even value for the pelletized hydrochar were \$157 and \$200 per ton of hydrochar, respectively, with a repayment period of 10 years [12].

In terms of process economics of co-HTC, Saba et al. [7] performed an economic analysis for co-HTC of coal and miscanthus blend and reported a break-even selling price of \$117 per ton to produce 110 MWe. The study was also stated that this break-even price could go even lower in case of a higher capacity plant. This study did not consider the transportation cost of coal or miscanthus. However, for waste materials such as coal waste and wet food waste, several scenario in terms of transportation could occur: (1) HTC performed at coal mine, where food waste and products need to be transported to coal-fired power plant; (2) HTC performed at coal-fired power plants, where both food waste and coal waste need to be transported; and (3) HTC performed at municipal solid waste treatment center (e.g., material recycle facility (MRF)), where coal waste and product need to be transported.

That is why the goal of this study was to perform an economic analysis of scaled-up hydrochar production process and analyze the effect of different parameters of hydrochar production process on the break-even cost through a sensitivity analysis for three abovementioned scenarios. The aim is to find out the most significant factors to reduce the overall production cost. This study is unique because it studied the economic feasibility of upgrading coal waste through co-hydrothermal carbonization process using food waste.

2 Materials and methods

2.1 Experimental methodology for co-hydrothermal carbonization

A bituminous coal waste, namely, 4Top, collected from south-east Ohio and food waste collected from the Central Food Facility of Ohio University were used as feedstocks in the co-HTC experiments. Both feedstocks were dried overnight in an oven at 105 °C. A 600-mL Parr reactor was used to perform co-HTC experiments. For co-HTC, a mixture of 50 wt% of coal refuse and 50 wt% of FW was mixed with

10 parts of deionized water. HTC runs were performed at 230 °C. A mixture of 50–50 wt% of coal waste and food waste was used for experimental simplicity as the mixture ratio does not affect the co-HTC output [4]. The reactor was heated at 3 °C per min to the desired temperature and held for 30 min. Two hundred thirty degrees Celsius and 30 min were chosen as the HTC temperature and reaction time because the authors' previous study determined that these process parameters produce hydrochar of optimum fuel quality from coal waste and food waste co-HTC [23]. The solid hydrochar was filtered from HTC process liquids and was dried overnight in an oven at 105 °C. The properties of co-HTC hydrochar are presented in Table 1.

2.2 Co-hydrothermal carbonization operation

Figure 1 shows the simplified process flow diagram of co-HTC operation. The coal waste and food waste feedstocks are mixed together in 1:1 ratio and process liquid from the previous run is added to increase the total moisture content. Later, the wet feedstock (slurry) is sent to positive displacement pump 1. The pump increases the pressure of the slurry to 200 PSIG before sending it to the U-tube heat exchanger to make sure the water does not get vaporized. Then, the heat exchanger increases the temperature of the slurry and send it to positive displacement pump 2. This pump later sends the slurry to the reactor after increasing the pressure of the slurry to the operating pressure of the reactor. There the feedstock is treated for 30 min at a reaction temperature of 230 °C. After treatment, the solid, liquid, and the gaseous product is sent back to the U-tube heat exchanger to recover the heat and use the excess pressure to send the product through leaf filter. The leaf filter separates the 80% of the process liquid from the product and the rest of the moisture is dried in the drying section. The end solid product contains approximately 11% of moisture which according to literature is acceptable limit for power plant [11]. The gas is vented in the atmosphere and the process liquid is sent back where it is mixed with the dry mixed coal waste and food waste to increase the moisture content before sending it to the positive displacement pump 1. The dried hydrochar is stored in a fixed roof tank. Table 2 [7, 24] lists the number of each unit required in the co-

hydrochar power plant. The heat exchanger and the pump will be operating at 80% efficiency.

2.3 Engineering economics methodology

The model described in the previous section is used to perform an economic analysis for a scaled-up hydrochar production through a co-HTC process using coal waste and food waste. The methodology described in this section followed design parameters mentioned in Turton et al. and was used by other studies [7, 24]. Data collected from previous studies and co-HTC runs of this study were used to perform a mass and energy balance. Later the mass and energy balance were used to determine necessary energy input, output, and hydrochar production which were then used to analyze the detailed cost of the production process.

The net present value (NPV) of the plant will be calculated from Eq. 1 [7, 24] using total capital investment (TCI) and cash flow.

$$\text{NPV} = \text{TCI} + \sum_{k=1}^n F_k \cdot (1+i)^{-k} \quad (1)$$

Here, F_k is the annual after-tax cash flow, i is the interest rate, k is the year being evaluated, and n is the total number of years the plant is operating. The term NPV determines whether a plant will make profit or not. A positive value will indicate a plant profit whereas a negative value will mean the opposite. Also, this value will determine the appropriate selling price for the plant to break even.

The TCI was calculated from the overall fixed capital investment (FCI) and working capital (WC). The WC is related to the money required in the very early stage of the plant and is usually not depreciated and recovered in the final year. It was calculated using Eq. 3.

$$\text{TCI} = \text{FCI} + \text{WC} \quad (2)$$

$$\text{WC} = 0.1 \cdot (\text{FCI} + C_{\text{OL}} + C_{\text{RM}}) \quad (3)$$

where C_{OL} is the operating labor cost and C_{RM} is the operating raw material cost. FCI usually includes general cost, unforeseen cost, and supporting site cost and was calculated by summing up individual bare module costs (C_{BM}) [7, 20].

Table 1 Mass yield, ultimate analysis, ash content, and HHV of raw and HTC-treated feedstock at 230 °C

Feedstock	Mass yield (%)	C (%)	H (%)	N (%)	S (%)	O (%)	Ash (%)	HHV (MJ/kg)
Raw coal waste	–	18.6 ± 1.3	1.8 ± 0.2	0.6 ± 0.1	8.5 ± 1.6	4.1 ± 1.1	66.4 ± 1.0	19.9 ± 0.1
Raw food waste	–	39.3 ± 1.8	6.0 ± 0.2	1.5 ± 0.1	0.0 ± 0.0	44.0 ± 1.8	9.2 ± 0.3	19.2 ± 0.8
Coal waste-H230	97.1 ± 0.8	20.2 ± 1.2	1.6 ± 0.0	0.7 ± 0.2	7.9 ± 1.6	5.2 ± 1.0	64.4 ± 1.3	20.5 ± 0.5
FW-H230	32.5 ± 0.6	60.6 ± 1.7	5.7 ± 0.1	3.3 ± 0.6	0.0 ± 0.0	28.9 ± 1.7	1.5 ± 0.1	26.8 ± 0.5
Co-HTC-H230	60.6 ± 0.2	49.5 ± 1.6	4.5 ± 0.1	1.9 ± 0.2	1.4 ± 0.1	14.8 ± 1.6	27.9 ± 0.8	23.0 ± 0.8

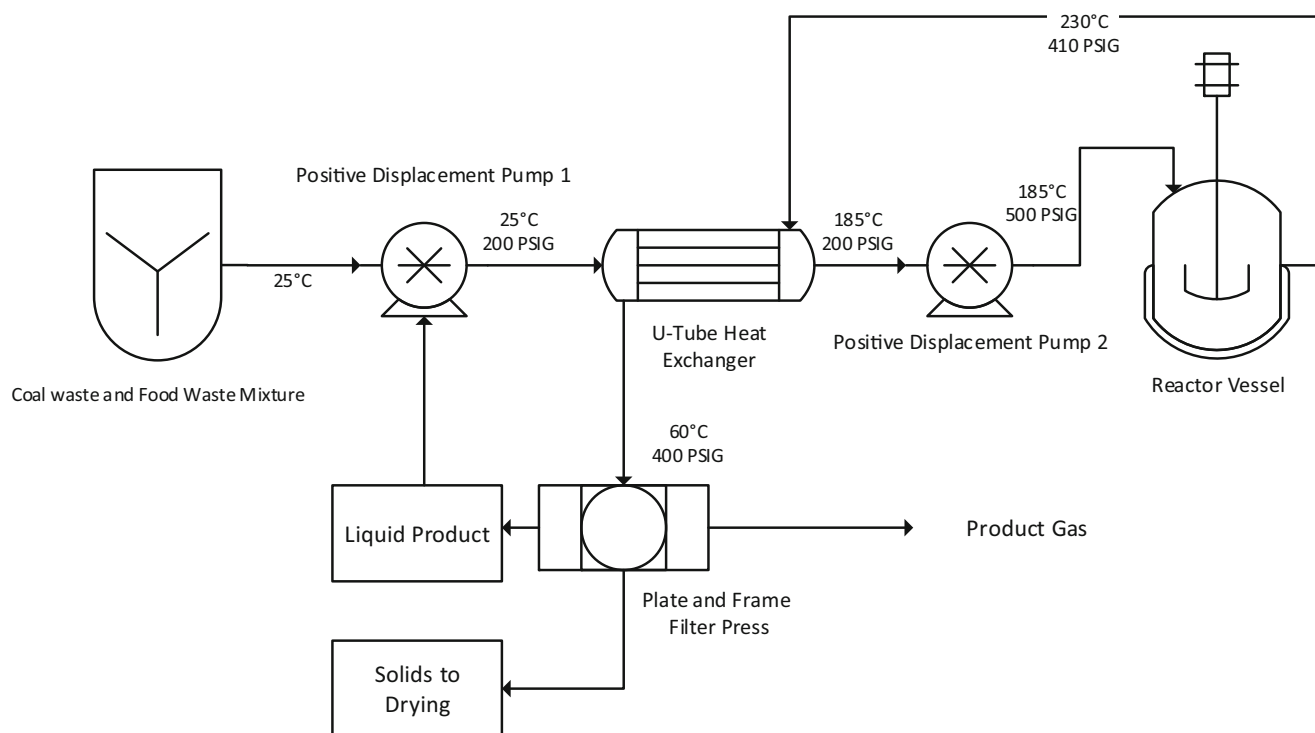


Fig. 1 Simplified process flow diagram of hydrochar production from co-HTC treatment of coal waste and food waste

$$FCI = \sum_j^m C_{BM,j} + 0.03 \sum_j^m C_{BM,j} + 0.15 \sum_j^m C_{BM,j} + 0.5 \times \sum_j^m C_{BM,j} \quad (4)$$

$$C_{BM} = C_p^\circ \cdot F_{BM} \quad (5)$$

where C_p° is the base cost of equipment and F_{BM} is the cost modifier to determine C_{BM} . F_{BM} includes direct and indirect cost as well as material and pressure rating modifier. C_p° was determined from Eq. 6.

$$\log(C_p^\circ) = K_1 + K_2 \cdot \log(A) + K_3 \cdot (\log(A))^2 \quad (6)$$

Here, the K_1 , K_2 , and K_3 are constants and A is primary design parameter. The values for the constant and the

parameter along with F_{BM} used in this economic analysis are presented in Table 2 [7, 24].

The F_k was determined by summing up the after-tax net profit and depreciation (d) for the year (n), depicted in Eq. 7.

$$F_k = (R - COM_d - d)(1 - t) + d \quad (7)$$

where R is revenue, COM_d is the cost of manufacturing, d is depreciation, and t is the tax rate. Number of products produced each year multiplied by estimated selling price generated the revenue, R . Equation 1 gave the number of products produced when the equation was solved for NPV = 0. COM_d was calculated from Eq. 8 [7, 24], which includes operating labor costs or fixed operating and maintenance (O&M) cost (C_{OL}), utilities (C_U), waste treatment costs (C_{WT}), material transportation cost (C_T), and general plant upkeep. C_U , C_{WT} , and C_{RM} are also known as variable O&M cost. All the O&M parameters are given in Table 3 [7, 24–30]. Multipliers were

Table 2 Constants used to determine cost associated with equipment shown in process flow diagram [7, 20]

Model unit	Quantity	Units of A	k_1	k_2	k_3	Modifier
Jacketed agitated reactor	7	m ³	4.1052	0.532	−0.0005	4.00
Positive displacement pump (pre)	2	kW	3.8696	0.3161	0.122	5.66
Positive displacement pump (post)	4	kW	3.8696	0.3161	0.122	6.51
U-tube heat exchanger	2	m ²	4.1884	−0.2503	0.1974	4.59
Leaf filter press	2	m ²	3.8187	0.6235	0.0176	1.80
Api-fixed roof tank (FW storage)	1	m ³	4.8509	−0.3973	0.1445	1.00
Api-fixed roof tank (product storage)	1	m ³	4.8509	−0.3973	0.1445	1.00

Table 3 Fixed and variable operation and maintenance (O&M) parameters as well as economic parameters used for determining cost associated with manufacturing [20–27]

Fixed O&M parameters		Variable O&M parameters		Economic parameters	
C_{OL} multiplier	2.76	Variable O&M multiplier	1.23	Tax rate (%)	25
Cost per laborer (\$)	52,700	Food waste (\$/ton)	5	Annual interest rate (%)	10
N_{np}	9	Coal waste (\$/ton)	5	Plant life (years)	20
P	2	Water utility (\$ m ⁻³)	1.12	Streaming factor	0.9
Op labor	53	Wastewater disposal (\$/ton)	0.74	Salvage value	0
		Cost of natural gas (\$/m ³)	0.132	Depreciation	7-year MACRS
		Natural gas energy content (MJ/m ³)	38.64	2016 CEPCI	541.7

used for C_{OL} , C_U , C_{WT} , and C_{RM} to account for administrative cost as well as any fluctuations or indirect cost related to utility, waste treatment, and transportation, respectively. General plant maintenance was assumed to be an 18% cost of the initial FCI.

$$COM_d = 0.18FCI + 2.73C_{OL} + 1.23(C_U + C_{WT} + C_{RM}) \quad (8)$$

The base number of operators per shift was calculated from Eq. 9 using the fixed O&M parameters presented in Table 3.

$$N_{OL} = (6.29 + 31.7 \cdot P^2 + 0.23 \cdot N_{np})^{0.5} \quad (9)$$

Here, P is the total number particulate handling unit operations and N_{np} is the total number of non-particulate handling unit operations.

This study assumed a raw material purchasing cost of \$5/ton for both coal waste and food waste. This study also considered three different cases for the economic analysis.

Case I: The hydrochar production was performed in the coal mine where the coal waste was produced. So, there was no transportation cost for coal waste. Only food waste was transported to the location. Coal mine located in Vinton County of southeast Ohio was selected as the location.

Case II: The hydrochar production was performed in a power plant location away from the coal mine where the coal waste was produced. So, both coal waste and food waste were transported to the power plant. A power plant located at Gallia County, Ohio, was selected in this study as the location of the power plant where the produced hydrochar can be used for electricity generation.

Case III: The hydrochar production was performed near the food waste processing facility where the food waste was stored. So, there was no transportation cost for food waste. Only coal waste was transported to the location.

Central Food Facility located in Ohio University, Athens, Ohio, was selected as the location.

The goal on each case was to produce 49,192 kg/h of co-HTC hydrochar. The reason of choosing this goal is because this exact amount of hydrochar can produce 110 MWe in case of power generation using solely hydrochar as the source [7].

2.4 Sensitivity analysis of hydrochar production

Sensitivity analysis was performed to evaluate the effect of different parameters on the break-even selling price. For the analysis, nine (9) different parameters were considered. These were hydrochar production, food waste and coal waste travel distance, food waste and coal waste purchase cost, cost of natural gas, waste treatment, hydrochar HHV, and FCI. When one parameter was changed, all the others were kept constant to evaluate the effect of the changed parameter on break-even price of hydrochar production. The parameters were changes from a lower sensitivity bound (LSB) to a higher sensitivity bound (HSB) to see the effect. Later, the break-even prices for each parameter were plotted in a sensitivity diagram.

3 Results and discussion

3.1 Process flow conditions for hydrochar production in co-HTC plant

Operational data for a co-hydrothermal carbonization plant operation are presented in Table 4. These operation data were determined for producing 49,192 kg/h of hydrochar from co-HTC operation of coal waste and food waste at 230 °C. The plant required an input of 81,175 kg/h of feedstock on dry basis to produce 49,192 kg/h of hydrochar. Fifty percent of this total required input was coal waste while the rest of it was food waste. The coal waste did not contain any moisture whereas the food waste contained approximately 70% of

Table 4 Experimental parameters derived from co-HTC plant operation

Parameters	Co-HTC
Hydrochar production (kg/h)	49,192
Reaction temperature (°C)	230
CW and FW solid ratio	1:1
Total feed water content (%)	85
Gas production ($\times 10^{-3}$ kg/kg solid feed)	4.9
Total process feed (kg/h)	541,172
Total solid feed (kg/h)	81,175
Total water feed (kg/h)	459,996
Produced hydrochar on dry basis (kg/h)	49,517
Total process liquid after treatment (kg/h)	491,655
Moisture dried from post filter hydrochar (kg/h)	1396

moisture. However, it was necessary to increase the moisture content of the mixture to 85% to make sure the positive displacement pumps can pump the feedstock. Also, it was reported in previous literatures that the feedstock is required to be completely submerged in water before the co-HTC operation begins [31, 32]. For this reason, the total water feed required for the desired operation was 459,996/h which is higher than the required amount. This huge amount of required water was supplied from the produced process liquid from the co-HTC operation. The co-HTC operation produced 491,655 kg/h of process liquid during operation which was used to increase the moisture content of the initial feedstock. The process produced 0.0049 kg of gas for per kg of solid feed [33]. As a result, a gas flow of 398 kg/h was produced during the operation. The co-HTC operation produced 49,517 kg/h of hydrochar which is slightly more than the required (Table 2). These process flow data were used to determine the design parameters and equipment sizing for the economic analysis. The equipment are shown in a simplified process flow diagram in Fig. 1 and design parameters along with the number of equipment required for this co-HTC operation are mentioned in Table 2.

3.2 Capital cost analysis

Table 5 lists the capital cost associated with the co-HTC hydrochar production process. The capital cost analysis was done using 2016 pricing. Total capital investment (TCI) was calculated from the summation of fixed capital investment (FCI), working capital (WC), and bare module cost. The bare module cost was calculated from the summation of the cost associated with individual systems and operations of the co-HTC plant, namely, reactor system, pump system, initial water utility cost, heat recovery system, solid product recovery and dewatering system, and storage system. This was done using Eq. 5 where the base cost of the equipment was multiplied

with the cost modifier F_{BM} . The base cost of individual equipment was determined using the constants and design parameters presented in Table 2. To account for the material rating modifier part of F_{BM} , SS-316 was selected as the construction material [7]. The total bare module cost was \$7.4M. The cost distribution illustrates how different systems of the plant contributed to the bare module cost. The reactor system cost approximately \$4.1M which accounted for 55.2% of the bare module cost. This was the highest among all the systems. Using 7 individual reactors for the hydrochar production was the main reason for this high cost associated with the reactor system. Each reactor operated at 500 PSIG and had a volume of 15 m³. The pumping system accounted for \$1.6M which was 22.81% of the total bare module cost. The co-HTC plant used six individual positive displacement pumps in two locations. Two pumps were located before the U-tube heat exchanger shown in Fig. 1. This two operated at 200 PSIG and 150 m pressure head. The rest of the pumps were installed after the U-tube heat exchanger and before the reactors. These operated at 300 PSIG and with 210 m pressure head. The heat recovery, filtration, and storage system were 6.57, 7.78, and 3.77% of the total bare module cost. The bare module cost was used later to calculate the FCI which was \$11.95M. Later, three different cases were considered to calculate WC and TCI.

The WC was calculated using Eq. 3 which considered cost of raw materials, C_{RM} . C_{RM} included both the raw material purchasing cost and the raw material transportation cost. For case I, the food waste was collected from the surrounding counties of Vinton. So, the WC for case I was \$2.2M which along with FCI and bare module cost made the TCI approximately \$14.15M. However, in case II, food waste along with coal waste was transported to power plant in Gallia. So, there was an additional transportation cost for coal waste which increased the WC by \$0.2M compared to case I. The TCI in case II was \$14.3M. Case III assumed transportation of coal waste to a food waste facility in Athens which brought the WC and TCI down to \$2.1M and \$14.09M, respectively. Case III showed the minimum WC and TCI among the three cases.

3.3 Manufacturing cost analysis

Fixed and variable operation and maintenance (O&M) cost of hydrochar production is presented in Table 5. Later, these base costs were multiplied by the respective multipliers presented in Table 3 to account for the indirect cost associated with in the process. The total manufacturing cost was then calculated by incorporating the fixed capital investment upkeep with fixed and variable O&M costs. The total manufacturing cost was \$22.4M, \$25.4M, and \$21.7M for case I, case II, and case III, respectively. The reason behind this difference in these three cases was the transportation cost associated with coal waste and food waste which made the raw material cost

Table 5 Summarized estimated cost for hydrochar production from co-HTC plant

	Cost			Cost Distribution
Reactor System	\$ 4,084,172			55.19%
Pumping Systems	\$ 1,688,141			22.81%
Initial Water Utility cost	\$ 226,283			3.10%
Heat Recovery System	\$ 485,853			6.57%
Solid Product Filtration and Dewatering	\$ 575,703			7.78%
Storage	\$ 278,628			3.77%
Bare Module Cost	\$ 7,400,000			-
FCI	\$ 11,949,000			-
	Case I	Case II	Case III	
Working Capital (WC)	\$ 2,202,447	\$ 2,442,273	\$ 2,140,190	
Total Capital Investment (TCI)	\$14,151,447	\$14,391,273	\$ 14,089,190	

	\$/year			\$/tonne		
Estimated FCI upkeep	\$ 2,150,820			\$ 5.5		
Labor costs	\$ 7,582,001			\$19.4		
Utilities	\$ 285,378			\$ 0.7		
Waste treatment	\$3,437,862			\$ 8.8		
	Case I	Case II	Case III	Case I	Case II	Case III
Raw Materials	\$ 8,976,758	\$ 11,926,623	\$ 8,211,000	\$ 23.0	\$ 30.6	\$21.0
Total Costs	\$ 22,432,819	\$ 25,382,684	\$ 21,667,061	\$ 57.5	\$ 65.0	\$ 55.5

Operation and Maintenance Cost			
Fixed O&M costs			
COL (\$/year)	\$ 2,777,290		
Variable O&M costs			
CU (\$/year)	\$ 232,015		
CWT (\$/year)	\$ 2,795,010		
	Case I	Case II	Case III
CRM (\$/year)	\$ 7,298,178	\$ 9,696,442	\$ 6,675,611

\$8.9M, \$11.9M, and \$8.2 M, respectively. The other costs were common in all the cases. The majority of the manufacturing cost came from the raw material cost. This raw material cost includes the transportation cost as well as their purchasing cost which was \$5/ton for both the raw materials. Now this raw material cost to produce per ton of hydrochar was \$23, \$30.6, and \$21 for case I, case II, and case III, respectively. Previous researchers have also reported similar results where raw material cost contributed to a significant portion of the manufacturing cost [34]. The second largest contribution was

from labor cost. The utility cost was low compared to other costs as process liquid was used in the operation instead of water which minimized the water utility cost. The FCI upkeep was \$2.2M which accounted for 8–10% of the total manufacturing cost. The total manufacturing cost was \$57.5, \$65, and \$55.5 per ton in case I, case II, and case III. Case II appeared to be the most expansive scenario as this included both the CW and FW transportation cost whereas rest of the two scenarios only included one raw material transportation cost.

3.4 Sensitivity analysis

The TCI and manufacturing cost determined in the previous sections were later used to evaluate the break-even selling price. The net present value equation was set to zero in both the cases which resulted in a break-even selling price of \$62.24 for case I, \$69.90 for case II, and \$60.26 for case III. Values presented in Tables 3 and 4 were used for this purpose. Both CW and FW transportation costs associated in case II made the break-even price high. To evaluate how different parameters affected the break-even price, a sensitivity analysis was performed. For the sensitivity analysis, different parameters, namely, hydrochar production, food waste and coal waste travel distance, food waste and coal waste purchase cost, cost of natural gas, waste treatment, hydrochar HHV, and FCI, were changed to their highest and lowest values within a range and then compared to the base case scenario. Table 6 presents the range within which the parameters were altered as well as the resulting break-even cost for each alteration and a comparison with the base case scenarios. Later the change of break-even cost (%) for each parameter is presented in Fig. 2. It is evident from Table 6 and Fig. 2 that the most dominating factors were the raw material cost as well as their travel distance and the hydrochar production capacity in both cases. For this study, the raw material cost was assumed to be \$5/ton for both coal waste and food waste. However, when the food waste price was increased to \$30/ton, the break-even cost drastically increased by 41%, 36%, and 42% for case I, case II, and case III, respectively. Similarly, the break-even cost increased by 74%, 66%, and 76% when the coal waste cost purchasing price was increased to \$50/ton for case I, case II, and case III, respectively. However, when the raw material cost was lowered down to \$−10 per ton, the break-even cost decreased significantly. The \$−10 signifies that the collection

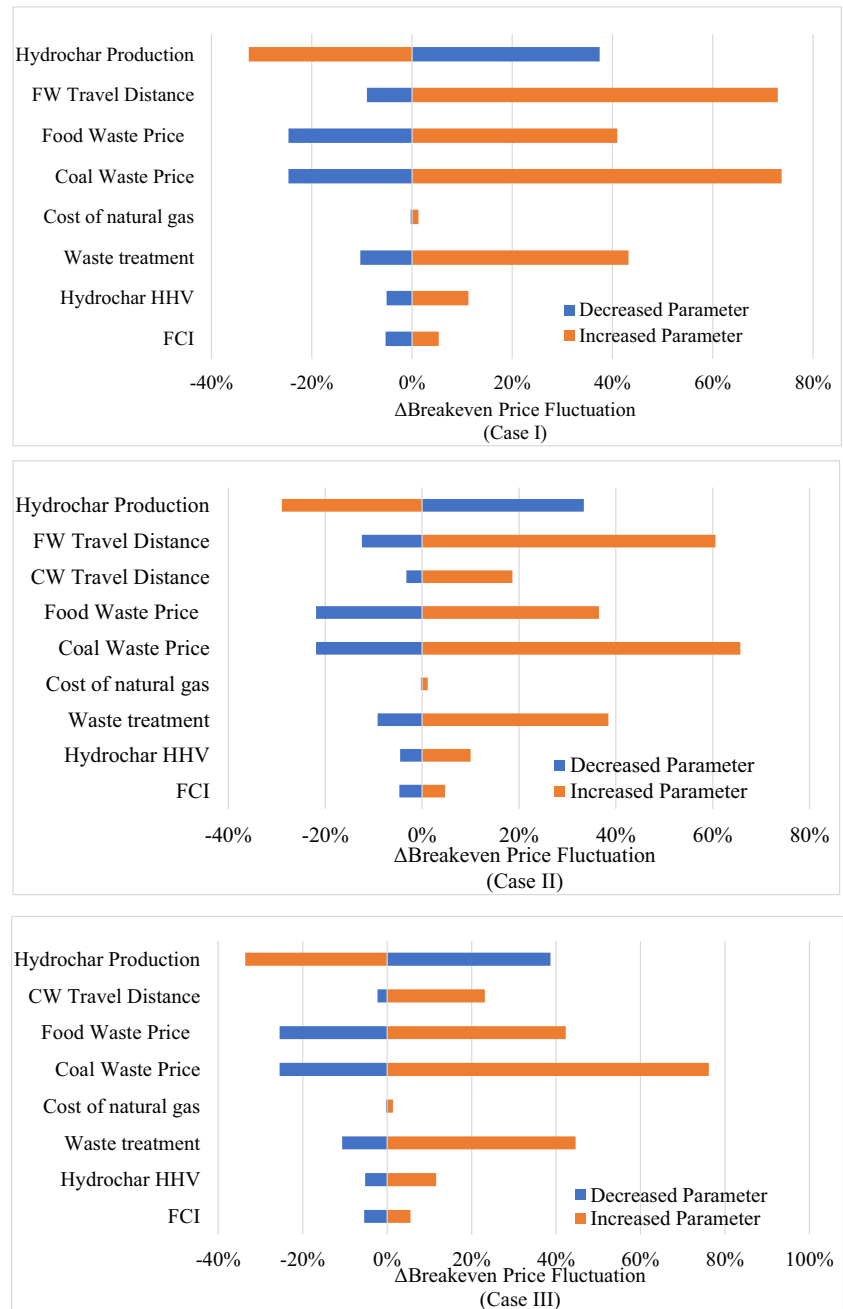
of raw materials would earn \$10 per ton. Additionally, when the food waste travel distance was increased to 50 miles from the base case scenario, the break-even cost increased approximately \$107 and \$112/ton for cases I and II, respectively. Similar increase was observed for coal waste travel distance in case II and case III. These scenarios prove that it is possible to reduce the break-even cost for co-HTC hydrochar production by switching to low-cost raw materials as well as minimum travel distance. Previous researchers observed the same trend with raw material cost and recommended switching to lower cost materials for reducing the price [7]. The wastewater treatment cost also had a significant effect on the cost. When the cost was increased to \$3/ton, the overall break-even cost increased by 43%, 38%, and 45% for cases I, II, and III, respectively. Hydrochar HHV, cost of natural gas, and FCI had comparatively less effect on the break-even cost. So, raw materials and plant location should be selected carefully to ensure lower purchasing cost and minimum travel distance to minimize the cost.

The co-HTC plant production capacity also had a significant effect on the break-even selling price of the hydrochar. The base case showed a hydrochar production capability of 49,192 kg/h for which the break-even costs were \$62.24 for case I, \$69.90 for case II, and \$60.26 for case III. However, when the production capability was scaled down to 24,596.27 kg/h, the cost increased significantly by 37%, 33%, and 39%, and consequently, the price decreased by 33%, 29%, and 34% when the hydrochar production rate was increased to 245,962.73 kg/h, for cases I, II, and III, respectively. The scenario was further illustrated in Fig. 3 where the break-even cost was plotted against hydrochar production rate for all the cases. The graph shows an exponential trend for break-even cost with respect to plant production capacity. For a plant capacity of 24.6 ton/h, the break-even

Table 6 Parameter range for sensitivity analysis and change of break-even cost for LSB and HSB

Items	Baseline scenario			Sensitivity range	LSB break-even cost (\$/ton)			HSB break-even cost (\$/ton)		
	Case I	Case II	Case III		Case I	Case II	Case III	Case I	Case II	Case III
Baseline scenario break-even price (\$/ton)	62.24	69.90	60.26	—	—	—	—	—	—	—
FW travel distance (mile)	5.46	8.52	—	0–50	56.64	61.20	—	107.67	112.23	—
CW travel distance	—	7.44	4.37	0–50	—	67.62	58.89	—	82.93	74.2
Hydrochar production (kg/h)	49,192			24,596.27–245,967.23	85.55	93.23	83.56	41.97	49.65	39.99
Food waste price (\$/ton)	5.00			−10 to 30	46.90	54.59	44.92	87.73	95.41	85.74
Coal waste price (\$/ton)	5.00			−10 to 50	46.90	54.59	44.92	108.14	115.83	106.16
Cost of natural gas (\$/m ³)	0.13			0.092–3.0	62.07	69.72	60.09	63.06	70.71	61.08
Waste treatment (\$/ton)	0.74			0.2–3	55.82	63.47	53.84	89.14	96.89	87.16
Hydrochar HHV (MJ/kg)	23.00			20–30	59.09	66.74	57.11	69.24	76.89	67.26
FCI (\$10 ⁶)	11.95			8.0–15.95	58.97	66.62	56.99	65.56	73.21	63.58

Fig. 2 Change in break-even price of co-HTC hydrochar with respect to baseline break-even cost for cases I, II, and III



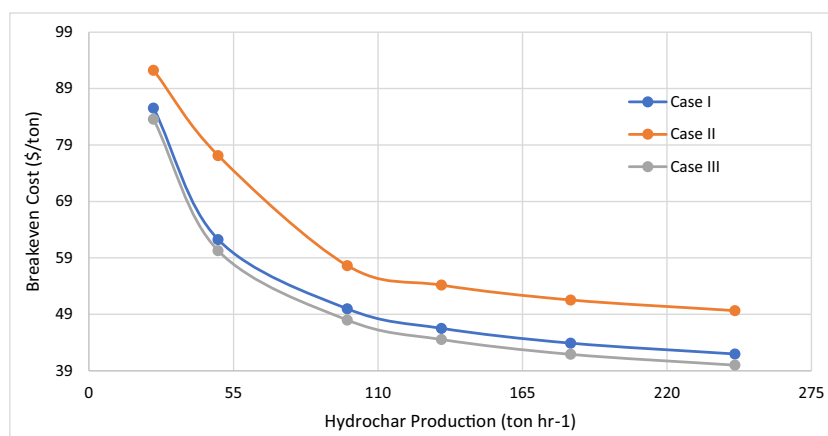
selling price increased to \$85.55, \$92.23, and \$83.56/ton from \$62.24, \$69.90, and \$60.26/ton for base case scenario of 49.19 ton/h production in case I, case II, and case III, respectively. However, when the production was increased by an order of magnitude to 245.96 ton/h, the cost reduced to \$41.97, \$49.65, and \$39.99/ton. So, this means that increasing the production of the plant will decrease the break-even cost. However, the graph suggests that further increase of plant capacity may not have significant effect on the break-even cost as the cost reduction became less and less significant with the increase. Also, increasing the plant capacity will have additional equipment cost and manufacturing cost which can

reduce the revenue generating from the plant. So, the plant capacity should be increased carefully to make sure that it reduced the break-even cost but does not affect the revenue.

4 Conclusions

This study was focused on analyzing the technoeconomic feasibility of hydrochar production through co-hydrothermal carbonization of coal waste and food waste mixture. A process flow diagram was designed with appropriate equipment and design parameters for production of 49,192 kg/h hydrochar

Fig. 3 Change of break-even cost with respect to hydrochar production



considering three different scenarios. The scenarios were based upon the raw material transportation. Case II showed the maximum WC and TCI while case III was the lowest. The break-even cost was \$62.24 for case I, \$69.90 for case II, and \$60.26 for case III. The raw material transportation cost affected the break-even cost as well as the manufacturing cost which turned out to be \$22.4M, \$25.4M, and \$21.7 M, for case I, case II, and case III, respectively. Later a sensitivity analysis was performed depending on 9 different variables to see how they affect the economic feasibility of hydrochar production. The sensitivity analysis revealed the raw material purchasing cost and raw material transportation cost to be the most influential variable. Additionally, when the hydrochar production capacity was increased 5 times the base case scenario, the break-even price decreased significantly for all cases compared to base case scenarios. However, a further analysis of break-even cost with respect to hydrochar production revealed that excessive increase of hydrochar production might not have any significant effect on break-even cost due to increased equipment and manufacturing cost. The maximum break-even cost was \$69.90 for case II while the lowest was \$60.26 for case III. A subsidy in the food waste might be required to reduce the price of co-HTC hydrochar to be economically competitive with coal.

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