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Process energetics for the hydrothermal carbonisation of human faecal wastes

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1 **Process Energetics for the Hydrothermal Carbonisation of Human Faecal** 2 **Wastes**

3
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12 **Abstract**

13
14 Hydrothermal carbonisation (HTC) has the capability to convert wet biomass such as sewage
15 sludge to a lignite-like renewable solid fuel of high calorific value. However, to date
16 assessment of the energy efficiency of the HTC process has not been fully investigated. In
17 this work, mass and energy balances of semi-continuous HTC of faecal waste conducted at
18 200°C and at a reaction time of 30 min are presented. This analysis is based on recovering
19 steam from the process as well energy from the solid fuel (hydrochar) and methane from
20 digestion of the liquid product. The effect of the feedstock solids content and the quantity of
21 feed on the mass and energy balance were investigated. The heat of reaction was measured at
22 200°C for 4 h with the wet faecal sludge, and the higher heating value was determined for the
23 hydrochar. The results indicated that preheating the feed to 100°C using heat recovered from
24 the process would significantly reduce the energy input to the reactor by about 59%, and
25 decreased the heat loss from the reactor by between 50–60%. For feedstocks containing 15–
26 25% solids (for all feed rates), after the process is in operation, energy recycled from the
27 flashing off of steam and combustion of the hydrochar and would be sufficient for preheating
28 the feed, operating the reactor and drying the wet hydrochar without the need for any external

29 sources of energy. Alternatively, for a feedstock containing 25% solids for all feed rates,
30 energy recycled from the flashing off of steam and combustion of the methane provides
31 sufficient energy to operate the entire process with an excess energy of about 19–21% which
32 could be used for other purposes.

33
34 **Keywords:** Bio-energy; Biomass; Hydrochar; Renewable energy; Sewage sludge

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38 1. Introduction

39 It is reported that about 1 billion tons of human faeces are generated each year [1], a
40 figure which will increase in line with the projected growth in population. It is estimated that
41 globally over 90% of sewage is discharged untreated [2] and that faecal contamination of
42 water sources causes almost 4 billion cases of diarrhoea each year, killing nearly 2.2 million
43 children under the age of five [3]. Even in those parts of the world where faecal waste
44 routinely undergoes treatment management of sewage sludge continues to present
45 environmental and health challenges. Management of faecal wastes, therefore, remains a
46 critical problem requiring proper mitigation techniques.

47 Hydrothermal carbonisation (HTC) is an effective method for converting wet biomass at
48 relatively mild reaction temperatures into a coal-like material commonly referred to as
49 ‘hydrochar’ along with aqueous products and gases - primarily CO₂ [4-6]. HTC has been
50 shown to produce substantial energy yields from various types of waste biomass of
51 agricultural origin [6-8]. However, attention has recently turned towards the HTC of sewage
52 sludges for energy generation [9-12]. One added benefit of this is that the process
53 temperatures typically employed in HTC result in the destruction of any pathogens present in
54 the sludge [13]. Most of the investigations cited above have concentrated on producing
55 renewable solid fuels and analysing the combustion properties of such fuels [8-10,14,15]. In

56 addition, some of these studies have focused on evaluating the energy efficiency of the
57 production of solid fuel by considering mechanical dewatering, drying and biofuel recovery
58 ratios [11,12]. Furthermore, the prospects for recovering energy from the liquid by-products
59 and the gas/vapour phase have also received considerable attention. The energy input of HTC
60 could be improved by combining HTC with anaerobic digestion of water-soluble by-products
61 [16-18] or their (and that of combustible gases and certain fractions of the hydrochar) wet air
62 oxidation [19-22]. The latter has been reported to improve subsequent anaerobic
63 biodegradability of the liquid products for enhance biogas yields.

64 HTC of biomass is widely reported [4,5,23] to be an exothermic process. In early reports
65 [23,24], it was stated that about 20–30% of the energy stored in the biomass is released as
66 heat during the HTC process, whilst between 60–90% of the heating value of the feedstock
67 remains in the hydrochar [24]. Heating values of hydrochars following HTC of sewage and
68 wastewater sludge range between 15–29 MJ kg⁻¹ [9-12,25] i.e. similar to that of lignite or
69 sub-bituminous coal [26,27]. In order to maximise energy yield after the process is in
70 operation, it is essential that the energy released during the process be recovered and utilised.
71 The hydrochar can be directly combusted to provide additional energy, or as recommended in
72 some studies, be blended with coal to improve the devolatilisation and ignition properties of
73 coal [14,15]. Erlach and Tsatsaronis [28] reported that flashing off steam from hot slurry
74 remaining in the reactor following treatment can improve the energy efficiency of the HTC
75 process. Stemann and Ziegler [29] proposed recovering heat from the hot compressed water
76 following the mechanical dewatering of the hydrochar to further improve the energy
77 efficiency. However, the amount of energy recovered from the process by such strategies
78 may not be sufficient to sustain the process, as energy is required to heat the faecal waste,
79 which typically contains about 90% water, to the reaction temperature, and also to dry the
80 hydrochar. Moreover, heat may be lost due to release of pressure at the end of the process

81 (particularly in batch HTC plants) and mixing of the material in the reactor following
82 treatment with the incoming cold feed (for semi-continuous and continuous plants). Heat
83 losses from the HTC reactor will occur as a result of radiation and convection but there is no
84 consensus as to their impact. Thorsness [30] reported significant heat losses during HTC of
85 municipal solid waste, whereas Namioka et al. [31] claimed that only insignificant heat losses
86 occurred during the HTC of sewage sludge. These conflicting claims emphasise the need for
87 a thorough investigation of the energetics of HTC processes.

88 The work presented here provides a framework for estimating energy utilisation, losses
89 and recovery within the HTC process for faecal waste treatment. Furthermore, consideration
90 is given to optimisation of the feed rate and the solids content in the faeces to determine the
91 best scale of operation for sustainability.

92 93 **2. Materials and Methods**

94 2.1. Materials

95 Primary sewage sludge (faecal sludge), was collected from Wanlip Sewage Treatment
96 Works (Leicestershire, UK). The faecal sludge contained 4.3% (wt.) solids as received. The
97 physical and chemical characteristics of the faecal sludge feedstock are shown in Table 1.

98 2.2. Experimental procedures

99 Triplicate batch HTC experiments were conducted using a 250 mL stainless steel reactor
100 (BS1506-845B, BTL Ltd, England, UK) immersed in an oil bath (B7 Phoenix II, Thermo
101 Scientific, UK) containing “THERMINOL ® 66” heating oil. HTC of primary sewage sludge
102 containing about 5% solids were carried out at 200°C for 30 min. The time taken for the
103 reactor to reach the reaction temperature was about 15 min. Further details of carbonisation
104 experiments are described in a previous work [16].

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106
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Table 1
Proximate and ultimate analysis of feedstock and hydrochar generated

Parameter	Basis	Feedstock (dried)	Hydrochar	Liquid product
<i>Proximate analysis</i> ^a				
Moisture	%	8.17 ± 0.25	4.58 ± 0.21	
Ash	% db	27.54 ± 0.63	36.29 ± 0.30	
Volatile Matter	% db	68.56 ± 0.83	62.30 ± 0.86	
Fixed Carbon ^b	% db	3.90 ± 0.05	1.42 ± 0.56	
HHV	MJ kg ⁻¹ (db)	17.79 ± 0.09	18.49 ± 0.56	
<i>Ultimate analysis</i> ^c				
C	%	37.63 ± 1.60	37.85 ± 0.25	
H	%	5.79 ± 0.26	5.39 ± 0.15	
O ^d	%	51.30 ± 1.69	53.72 ± 0.18	
N	%	5.29 ± 0.17	3.04 ± 0.06	
Hydrochar yield	%		66.83 ± 1.00	
Carbon recovery ^e	%		57.05 ± 0.37	
Fixed carbon recovery ^f	%		24.65 ± 9.81	
Energy recovery ^g	%		70.64 ± 2.25	
COD	g L ⁻¹			21.30 ± 2.52

^a ASTM D7582-10. ^b100 – (Moisture + ash + volatile matter). ^c ASTM D5373-08.
^d Calculated as difference between 100 and total C/H/N.
^e (%C in hydrochar * hydrochar mass / %C in feedstock * dry feedstock mass) * 100 [16].
^f (%fixed carbon in hydrochar * hydrochar mass / %fixed carbon in feedstock * dry feedstock mass) * 100 [32]. ^g HHV of hydrochar * hydrochar mass / HHV of feedstock * dry feedstock mass) * 100 [32]. db = dry basis

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109 2.2. Experimental procedures

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 112 Scientific, UK) containing “THERMINOL ® 66” heating oil. HTC of primary sewage sludge
 113 containing about 5% solids were carried out at 200°C for 30 min. The time taken for the
 114 reactor to reach the reaction temperature was about 15 min. Further details of carbonisation
 115 experiments are described in a previous work [16].

116 2.2.1. Feedstock and product analysis

117 Proximate and ultimate analyses were conducted to characterise both the feedstock and
 118 hydrochar. Residual moisture, ash and volatile matter were analysed using a

119 thermogravimetric analyser (TA Instruments Q5000IR, Elstree, UK), according to ASTM
120 method D7582-10 [33]. Carbon (C), hydrogen (H) and nitrogen (N) contents were analysed
121 using a CHN Analyser (CE-440 Elemental Analyser, Exeter Analytical Inc., Coventry, UK),
122 according to ASTM D5373-08 [34]. Energy content of the PSS feedstock and hydrochars was
123 determined using a bomb calorimeter (CAL2K, Digital Data Systems, Randburg, South
124 Africa). The specific heat capacity of dry sewage sludge was determined theoretically by the
125 method of Namioka et al. [31] and that of the hydrochar was the value used by Stemann and
126 Ziegler [29]. Chemical oxygen demand (COD) in the liquid product was measured using a
127 COD analyser (Palintest 8000, Palintest Ltd, UK) at a wavelength of 570 nm, in accordance
128 with Standard Methods 5220 D – Closed Reflux Colorimetric Method [35]. All
129 determinations were conducted in triplicate.

130 2.2.2. *Heat of reaction measurement*

131 The heat of reaction was measured in triplicate using a heat flux differential scanning
132 calorimeter (DSC-Q10, TA Instruments, Crawley, UK) at 200°C for a reaction time of 4 h,
133 being the reported time for a complete reaction [36]. Approximately 8 mg of faecal sludge
134 (4.3% solids content) was heated in stainless steel high pressure capsules (TA Instruments,
135 USA). Empty sealed pans were used as the reference capsules. The experiments and evaluation
136 of the results followed ISO11357-5:1999 and ISO11357-1:2009 using the isothermal method
137 [37,38]. For comparison, additional measurements were conducted at reaction temperatures
138 of 160 and 180°C for 4 h. Before the heat of reaction measurements, the DSC was calibrated
139 using 10 mg of standard grade indium metal (LGC, Teddington, Middlesex, UK). The
140 nitrogen purge gas flow was set at 0.5 ml/min. The cooler temperature was held at 40°C by
141 DSC re-generated cooling system (TA Instruments).

142 As described by ISO 11357-1:2009, the sample and reference capsules were reweighed
143 after each run to determine if there were changes in mass that could have disturbed the

144 instrument baseline or created additional thermal effects [38]. The heat of reaction during the
145 isothermal stage was estimated by integrating the area between the peak and the baseline. The
146 results were evaluated using the method proposed by Funke and Ziegler [39], by defining the
147 interval of integration in order to reduce uncertainties in determining the virtual baseline.

148 2.3. Modelling

149 The model developed here was based on a semi-continuous HTC plant comprising eight
150 components; a feed tank (and pre-heater), a reactor, a flash vessel, a pressure filter, a solids
151 dryer, an anaerobic digester (AD), and two combustion units. The plant capacity was varied
152 between 4.00–400.00 kg/day of wet faecal waste, representing faeces generated by between
153 10 and 1000 people per day. The model assumed solids content in the faecal waste to vary
154 between 5%, 15% and 25%. The faecal waste was heated to 200°C for 30 min (via a heating
155 unit) for sterilisation. The treatment time used here was selected on the basis of results from
156 previous study [16], which indicated that HTC at 200°C for 30 min produced hydrochars
157 having optimal characteristics; with the HHV of the hydrochar at those conditions similar to
158 the HHV of sub-bituminous coal. Although batch or continuous systems are typically applied
159 in a HTC plant, continuous or semi-continuous systems promote efficient utilisation of the
160 heat of reactions well as effective application of adjacent equipment, and also pressure
161 changes in the reactor are prevented [29].

162 2.3.1 Heat recovery routes

163 The two processes that require energy inputs are: (1) heating of the faecal waste to the
164 reaction temperature of 200°C; (2) drying of the wet hydrochar to less than 5% moisture
165 before combustion to generate energy to power the plant. A previous study on filterability of
166 slurry following HTC showed that for slurry from carbonisation conducted at 200°C for 30
167 min, hot-filtering the slurry at about 100°C resulted in hydrochars having water contents of

168 about 50% [40]; such hydrochars would require drying before combustion. No additional or
 169 external source of energy was required for dewatering the hydrochar.

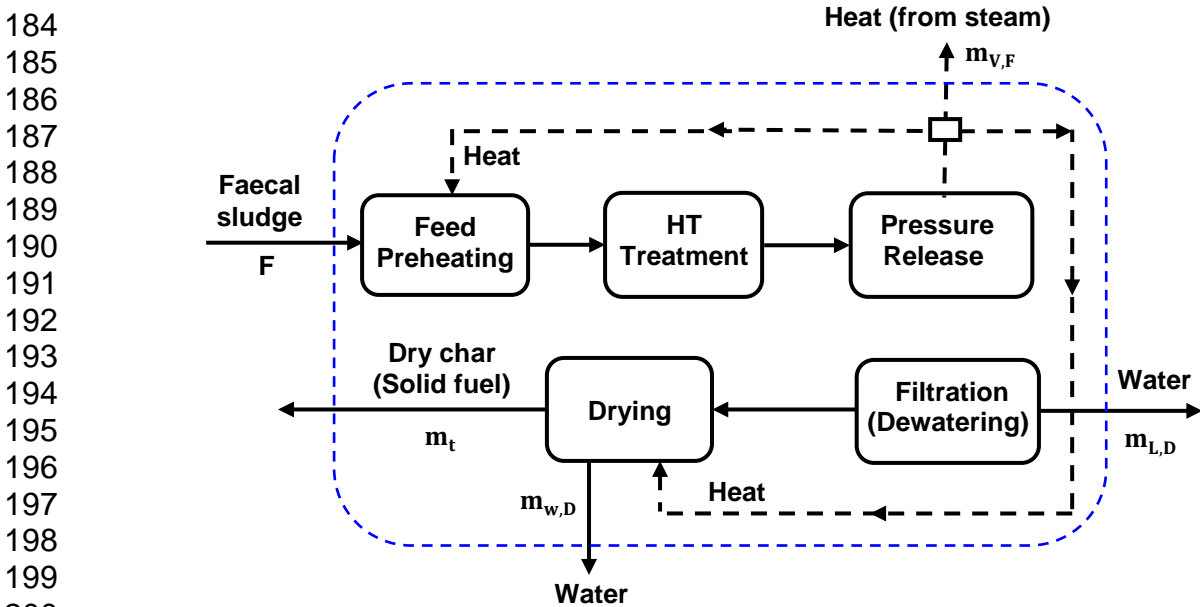
170 After the HTC process has been initiated, energy can be recovered from three
 171 mechanisms: (1) steam from the flash tank; (2) combustion of the hydrochar; (3) combustion
 172 of biogas produced from anaerobic digestion (AD) of the liquid product. Estimated methane
 173 yields from previous work showed that a yield of 52% was attainable from such liquid
 174 products following HTC at 200°C for 30 min [16]. The overall energy recycled will be used
 175 to preheat the feedstock, dry the hydrochar, and heat the reactor.

176 *2.3.2 Mass and energy balances*

177 The total mass balance was evaluated for the operations taking place within the
 178 boundaries shown in Figure 1 as follows:

179
$$F = m_t + m_{L,D} + m_{w,D} + m_{V,F} \tag{1}$$

180 where F is the mass of wet faecal feedstock (kg), m_t is mass of dried hydrochar obtained at
 181 reaction time t (kg), $m_{L,D}$ is the mass of liquid after filtration of the carbonised slurry (kg),
 182 $m_{w,D}$ is the mass of water evaporated from the wet hydrochar during drying (kg), and $m_{V,F}$ is
 183 the mass of steam or water vapour recovered from the flash vessel (kg).



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Figure 1 – Basis of mass balance calculations.

202 The energy balance was modelled as follows:

$$203 \quad \left(\begin{array}{c} \text{Energy} \\ \text{input} \end{array} \right) = \left(\begin{array}{c} \text{Energy to} \\ \text{heat reactor} \end{array} \right) + \left(\begin{array}{c} \text{Energy to heat} \\ \text{faecal sludge} \end{array} \right) + \left(\begin{array}{c} \text{Energy to} \\ \text{heat water} \end{array} \right) + \left(\begin{array}{c} \text{Heat of} \\ \text{reaction} \end{array} \right) - \left(\begin{array}{c} \text{Heat} \\ \text{loss} \end{array} \right) \quad (2)$$

204 That is, energy balance before energy is recovered from the process is

$$205 \quad V_o \rho_o c_{p,o} (T - T_0) + (H_U \times \tau_h) = [m_r \cdot c_{p,r} (T - T_0)] + [m_{FS} \cdot c_{p,FS} (T - T_0)] \\ + [m_w (H_{L(T)} - H_{w,(T_0)})] + m_o \cdot \Delta H_R - [A_r U_r t_h (T - T_0)] \quad (3)$$

206 where: $c_{p,o}$ is the specific heat capacity of heating oil ($\text{kJ kg}^{-1} \text{K}^{-1}$); $c_{p,FS}$ is the specific heat
207 capacity of dry sewage sludge (kJ kg K^{-1}), used to represent that of faecal sludge; $c_{p,r}$ is the
208 specific heat capacity of the reactor (kJ kg K^{-1}); ρ_o is the density of heating oil (kg m^{-3});
209 m_{oil} is the mass of heating oil (kg); m_r is the reactor mass (i.e. density of reactor material x
210 reactor volume, kg); $m_{FS} = m_o$, is the mass of solids in the wet sludge as (kg); m_w is the
211 mass of water in the sludge (kg); $H_{w(T_0)}$ is the enthalpy of water at initial temperature (kJ kg^{-1});
212 $H_{L(T)}$ is the enthalpy of water at the saturated liquid temperature, T (kJ kg^{-1}); H_U is heating
213 unit utility (kW); τ_h is oil holding time (min); ΔH_R is the heat of reaction during holding
214 period (kJ kg^{-1}); U_r is the overall reactor heat transfer coefficient ($\text{W m}^{-2} \text{K}^{-1}$); A_r is the
215 reactor heat transfer area (m^2); and t_h is reaction time – including heat up time (45 min).

216 The reactor heat transfer area is given by

$$217 \quad A_r = \pi L(D + t_I) + 2\pi r^2 \quad (4)$$

218 where: D is reactor diameter (m); L is reactor height (m); r is the reactor radius (m); and t_I is
219 the insulation thickness (m).

220 Ignoring fouling factors, the overall heat transfer coefficient of the reactor is:

$$221 \quad \frac{1}{U_r} = \frac{1}{h_m} + \frac{1}{h_I} + \frac{1}{h_A + h_R} \quad (5)$$

222 where: h_A is the heat transfer coefficient at the reactor wall ($\text{W m}^{-2} \text{k}^{-1}$); h_I is the conduction
223 coefficient of insulation ($\text{W m}^{-2} \text{k}^{-1}$); h_m is the conduction coefficient of the metal walls of
224 the reactor ($\text{W m}^{-2} \text{K}^{-1}$); h_R is the radiation coefficient of side walls ($\text{W m}^{-2} \text{K}^{-1}$).

225 The heat transfer coefficient at the reactor wall was calculated by modifying the equation
 226 proposed by Kato et al. [41] as follows:

$$227 \quad h_A = 0.138 \times (N_{Gr})^{0.36} \times [(N_{Pr})^{0.175} - 0.55] \times k_{air}/L \quad (6)$$

$$228 \quad N_{Gr} = (L^3 \times \rho_{air}^2 \times g \times \beta \times \Delta T) / \mu_{air}^2 \quad (7)$$

$$229 \quad N_{Pr} = c_{p_{air}} \times \mu_{air} / k_{air} \quad (8)$$

230 where: N_{Gr} is the Grashof number, N_{Pr} is the Prandtl number, k_{air} is the thermal
 231 conductivity of air at 25°C (W m⁻¹ K⁻¹); $c_{p_{air}}$ is the specific heat capacity of air at 25°C (kJ
 232 kg⁻¹ K⁻¹); μ_{air} is the viscosity of air at 25°C (kg m⁻¹ s⁻¹); ρ_{air} is the density of air at 25°C (kg
 233 m⁻³); β is the coefficient of thermal expansion of air at 25°C (K⁻¹), g is the gravitational
 234 constant (m s⁻²), L is the reactor height (m); ΔT is the average temperature (°C).

235 The average temperature drop is described in Eq. (9) using the expression proposed by
 236 Kumana and Kothari [42].

$$237 \quad \Delta T = (T - T_A) / 4 \quad (9)$$

238 where T_A is the outside air temperature (°C).

239 The conduction coefficient of the metal wall and insulation are defined as follows:

$$240 \quad h_m = \frac{k_M}{t_M} \quad (10)$$

$$241 \quad h_I = \frac{k_I}{t_I} \quad (11)$$

242 where: k_M is the thermal conductivity of the metal of the reactor at 200°C (W m⁻¹ °C⁻¹); k_I is
 243 the thermal conductivity of the insulation used (calcium silicate at 200°C, W m⁻¹ °C⁻¹); t_M is
 244 the reactor wall thickness (m).

245 The radiation coefficient of side walls is defined by Perry and Chilton [43] as follows:

$$246 \quad h_R = 0.1713 \times \varepsilon \times \frac{[(T_{IS} + 460/100)^4 - (T_A + 460/100)^4]}{T_{IS} - T_A} \quad (12)$$

$$247 \quad T_{IS} = T_A + 0.25(T - T_A) \quad (13)$$

248 where: ε is the surface emissivity of the aluminium jacket, T_{IS} is the outside temperature of
 249 the insulated surface ($^{\circ}\text{C}$), 460 is the temperature conversion factor from K to $^{\circ}\text{F}$. T_A and T_{IS}
 250 were converted from $^{\circ}\text{C}$ to K during calculation of h_R .

251 After the process is started, energy recycled from steam is calculated using the following
 252 equation as:

$$253 \quad E_s = [m_w \times (\hat{H}_{w(T)} - \hat{H}_{w(T_0)})] - m_0 \times \text{HHV}_f \quad (14)$$

254 where: E_s is the steam energy (kJ h^{-1}); HHV_f is the heating value of the solids in the sludge
 255 (kJ kg^{-1}); $\hat{H}_{w(T)}$ is the specific enthalpy of water at the reaction temperature, T (kJ kg^{-1}); and
 256 $\hat{H}_{w(T_0)}$ is the specific enthalpy of water at reference temperature, 25°C (kJ kg^{-1}).

257 The total mass of steam produced from flashing, $m_{V,F}$ was then calculated as:

$$258 \quad m_{V,F} = \left(1 - \frac{\hat{H}_{S(T)} - \hat{H}_{w(T)}}{\hat{H}_{S(T)} - \hat{H}_{w(120^{\circ}\text{C})}} \right) \times m_w \quad (15)$$

259 where: $\hat{H}_{S,T}$ is the specific enthalpy of steam at reaction temperature, T (kJ kg^{-1}); and
 260 $\hat{H}_{w,120^{\circ}\text{C}}$ is the specific enthalpy of water at 120°C (kJ kg^{-1}).

261 The energy from the steam for preheating the feedstock, assuming negligible heat losses,
 262 is given by:

$$263 \quad m_{V,PF} \cdot H_{V(T)} = [m_{FT} \cdot c_{p,FT} (T_f - T_0)] + [m_{FS} \cdot c_{p,FS} (T_f - T_0)] \\ 264 \quad + [m_w (H_{w(T_f)} - H_{w(T_0)})] \quad (16)$$

265 where: $c_{p,FT}$ is the specific heat capacity of feed tank ($\text{kJ kg}^{-1} \text{K}^{-1}$); $m_{V,PF}$ is the mass of
 266 steam from flash vessel for preheating feed (kg); m_{FT} is the mass of feed tank; T_f is the feed
 267 pretreatment temperature ($^{\circ}\text{C}$); and $H_{w(T_f)}$ is the enthalpy of water at pretreatment
 268 temperature (kJ kg^{-1}), $H_{V(T)}$ is the specific steam energy, kJ kg^{-1} (i.e. $E_s/m_{V,F}$).

269 The energy for drying wet hydrochar, assuming negligible heat losses, is given by:

$$270 \quad m_{V,D} \cdot H_{V(T)} = [m_{H,D} \cdot c_{p,H} (T_D - T_H)] + (m_{W,H} \times h_{\text{vap}}) \quad (17)$$

271 where: $m_{V,D}$ is the mass of steam for drying (kg); $m_{H,D}$ is the mass of wet hydrochar fed to
 272 the drying process (kg); $m_{W,H}$ is the mass of water in hydrochar (kg); $c_{p,H}$ the specific heat
 273 capacity of hydrochar ($\text{kJ kg}^{-1} \text{K}^{-1}$); h_{vap} is the latent heat of vaporisation of water (kJ kg^{-1});
 274 T_D is the drying temperature ($^{\circ}\text{C}$); and T_H is the temperature of wet hydrochar entering the
 275 dryer ($^{\circ}\text{C}$).

276 The energy required to heat the reactor and contents after completion of one process
 277 cycle is given by:

$$278 \quad Q_{char} + Q_{CH_4} = [m_r \cdot c_{p,r} (T - T_r)] + [m_{FS} \cdot c_{p,FS} (T - T_f)] + \Delta H_R m_0$$

$$279 \quad + [m_w (H_{L,(T)} - H_{w,(T_f)})] - [A_r U_r t_h (T - T_f)] \quad (18)$$

280 where: Q_{char} and Q_{CH_4} are the energies produced from combustion of dry hydrochar and
 281 methane (kJ), respectively; T_f is the temperature of preheated feedstock ($^{\circ}\text{C}$); and T_r is the
 282 temperature of the reactor body after completion of a process cycle (120°C); and $H_{w(T_f)}$ is the
 283 enthalpy of water at the preheat temperature, T_f (kJ kg^{-1}).

$$284 \quad Q_{char} = m_t \times \text{HHV of char} \quad (19)$$

$$285 \quad Q_{CH_4} = m_{CH_4} \times H_{CH_4} \quad (20)$$

286 where H_{CH_4} is the heat of combustion of methane (50125 kJ kg^{-1}) from stoichiometric
 287 combustion equation. m_{CH_4} is the mass of methane (kg) obtained from the relationship
 288 between mass of CH_4 generated and that of COD removed during anaerobic digestion (1 g
 289 COD removed = 0.25 g CH_4 produced, which is equivalent to 1.4 L CH_4 at STP [44-46], and
 290 on the basis that 90% of the COD was converted to CH_4 [47]. That is by proportion: {mass of
 291 CH_4 produced (kg) = mass of COD (kg) x 0.9 x 0.25 (kg)}/1 kg COD. Table 2 gives the
 292 physical and thermodynamics properties of the HTC reactor, faecal waste and products.

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294

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Table 2
Physical and thermodynamic properties of the reactor and operational data

Parameter	Notation	Unit	Value
Reactor and other units			
Specific heat capacity of stainless steel	$c_{p,r}; c_{p,PT}$	$\text{kJ kg}^{-1} \text{K}^{-1}$	0.502
Density of stainless steel reactor (at 25°C)	ρ_r	kg/m^3	8027.2
Reactor diameter ^a	D	m	0.502
Reactor/jacket height ^a	L	m	0.741
Reactor jacket diameter	D_j	m	1.34
Reactor thickness	t_M	m	3.18×10^{-02}
Insulation thickness	t_I	m	0.062
Thermal conductivity of insulation at 200°C ^b	k_I	$\text{W m}^{-1} \text{°C}^{-1}$	0.068
Thermal conductivity of stainless steel at 200°C	k_M	$\text{W m}^{-1} \text{°C}^{-1}$	17.0
Surface emissivity of jacket aluminium	ε		0.05
Volume of heating oil ^a	V_o	m^3	
Feedstock and products			
Specific heat capacity of dry sewage sludge	$c_{p,FS}$	$\text{kJ kg}^{-1} \text{K}^{-1}$	1.7 ^c
Specific heat capacity of hydrochar	$c_{p,H}$	$\text{kJ kg}^{-1} \text{K}^{-1}$	1.45 ^d
Specific heat capacity of water at 25°C	$c_{p,w}$	$\text{kJ kg}^{-1} \text{K}^{-1}$	4.187
Enthalpy of vaporisation of water	h_{vap}	kJ kg^{-1}	2270
Specific enthalpy of water at 25°C	$H_{w(T_o)}$	kJ kg^{-1}	104.8
Specific enthalpy of saturated water at 200°C	$H_{L(T)}$	kJ kg^{-1}	859.0
Specific enthalpy of water at 100°C	$H_{w(T_f)}$	kJ kg^{-1}	419.1
Specific enthalpy of steam at 120°C	$H_{V(T_F)}$	kJ kg^{-1}	2706.0
Specific enthalpy of steam at 200°C	H_S	kJ kg^{-1}	2790.0
Operational data			
Specific heat capacity of heating oil	$c_{p,o}$	$\text{kJ kg}^{-1} \text{K}^{-1}$	1.57
Density of heating oil (at 25°C)	ρ_o	kg m^{-3}	1005.86
Heating unit utility	H_U	kW	3.0
Holding time	τ_h	min	45 ^e
Reaction temperature	T	°C	200
Reference (feedstock) temperature	T_o	°C	25
Temperature of preheated feedstock	T_f	°C	100
Temperature of steam from flash tank	T_{FT}	°C	120
Drying temperature	T_D	°C	120
Temperature of hydrochar to dryer	T_H	°C	100
Density of air (at 25°C)	ρ_{air}	Kg m^{-3}	1.25
Specific heat capacity of air (at 25°C)	c_{pair}	$\text{kJ kg}^{-1} \text{K}^{-1}$	1.005
Viscosity of air (at 25°C)	μ_{air}	$\text{kg m}^{-1} \text{s}^{-1}$	1.98×10^{-5}
Thermal conductivity of air	k_{air}	$\text{W m}^{-1} \text{K}^{-1}$	0.0257
Coefficient of thermal expansion of air (at 25°C)	β	K^{-1}	3.43×10^{-3}
Gravitational constant	G	m s^{-2}	9.81

^a Varies based on reactor size or the number of faeces to be fed. ^b Calcium silicate.

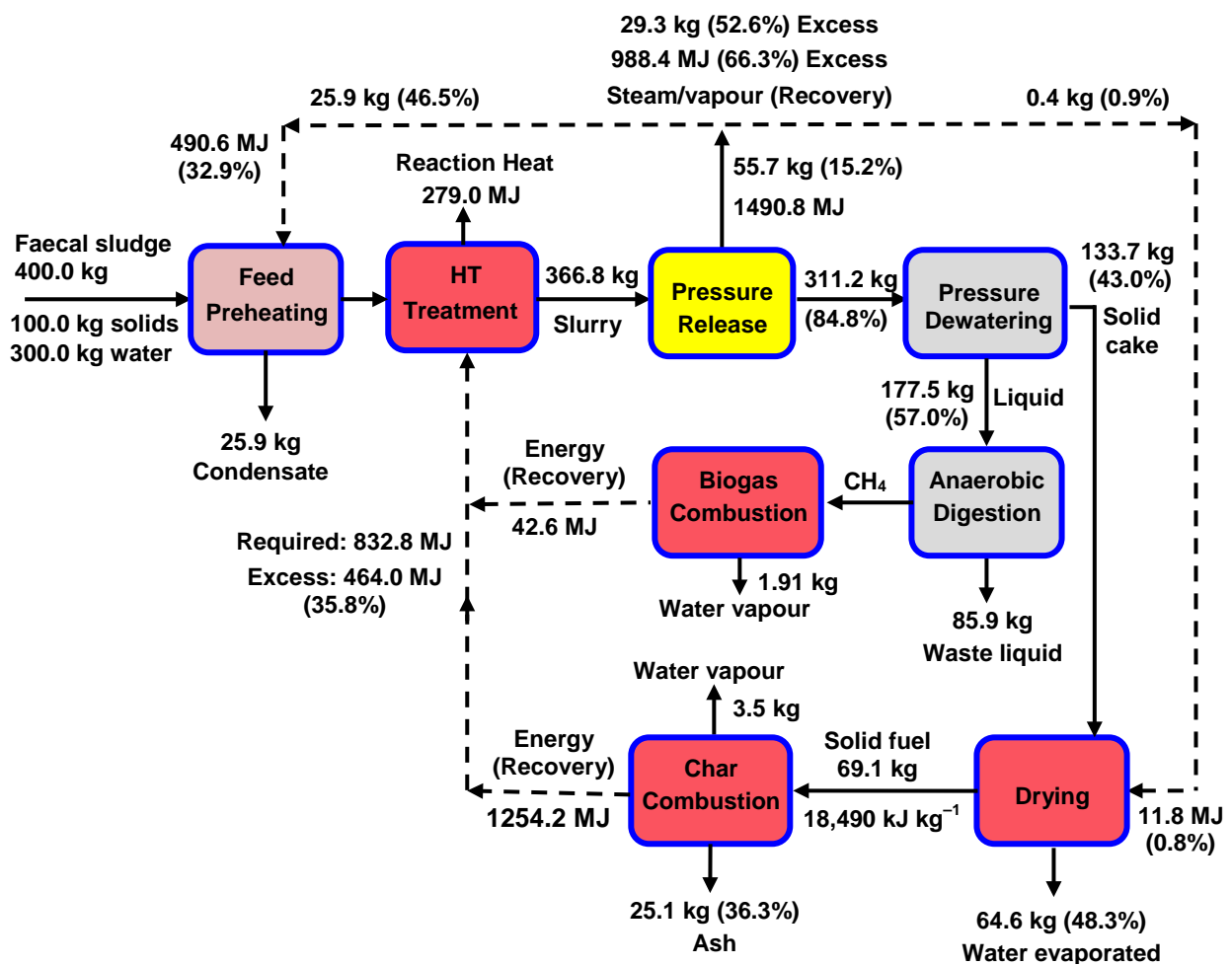
^c [31]. ^d [29]. ^e Heat up time and reaction time.

297 **3. Results and Discussion**

298 3.1. Mass balance

299 Figure 2 summarises the mass balance for the HTC process carried out at 200°C and for a
 300 reaction time of 30 min using faecal sludge with solid concentrations of between 5–25%, on a
 301 per day basis and assuming a 12 h process operation. The hydrochar yield was about 67% of
 302 the initial solids in the faeces following carbonisation under these conditions. After drying,
 303 the hydrochar contained about 5% water (using the results of residual moisture content in

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325 **Figure 2** – Energy balance on a semi-continuous HTC system based on heat recycled from
 326 both the process and products using feedstock with 25% solids and a daily feed rate
 327 equivalent to 400.00 kg.

328 hydrochar – Table 1). As the quantity of faeces treated increases from 4.00 to 400.00 kg per
329 day, the amount of hydrochar produced after drying increases from 0.12–13.80 kg, 0.48–
330 41.52 kg, and 0.72–69.12 kg when the solid content in the feedstock increases from 5%, 15%
331 and 25%, respectively (Table A.1 in the appendix). The amount of steam from the flash tank
332 increases significantly as the quantity of faeces treated increases from 4.00 to 400.00 kg per
333 day, but decreased as the solids content in the faeces increased or the liquid fraction
334 decreased. The amount of steam released from the flash tank ranged from 0.60–70.56 kg for
335 5% solids concentration, 0.60–63.24 kg for 15% solid concentration, and 0.60–55.68 kg for
336 25% solids concentration (Table A.1).

337 As shown in Figure 2, the amount of steam released from the Flash tank represents about
338 15% of the total quantity of material (slurry) fed to the Flash tank from the HT reactor, while
339 the remaining 85% consists of solids and water. For feedstock containing 15–25% solids,
340 only about 47% of the amount of steam released is required to preheat the feed to 100°C and
341 dry the hot hydrochar containing about 50% moisture. Therefore, 53% of the steam generated
342 is available for other purposes. The amount of water evaporated during drying of the wet
343 solids ranged between 0.12–12.96 kg, 0.36–38.64 kg, and 0.60–64.56 kg for faeces
344 containing 5%, 15%, and 25% solids respectively as the quantity of faeces treated was raised
345 from 4.00–400.00 kg per day (Table A.2). The mass of estimated methane yields decreased as
346 the solids content in the feedstock increases and ranged from 0.01–1.42 kg, 0.01–1.13 kg, and
347 0.01–0.85 kg for faeces with 5%, 15%, and 25% solids respectively. Previous studies
348 reported that HTC of biomass wastes followed by AD could enhance methane yields from
349 72–222% [16-18]; whilst carbonising digestate from AD by HTC has been reported to
350 improve energy recovery from biomass wastes [11,25,48] with the combined AD-HTC
351 doubling the energy recovery compared to AD process alone [48].

352 The mass of liquid waste that remained after anaerobic digestion ranged between 1.44–
353 143.28 kg, 1.20–114.60 kg, and 0.84–85.92 kg for faeces containing 5%, 15%, and 25%
354 solids respectively; representing about 48% of the liquid filtrate fed to the anaerobic digester,
355 which is recycled. The mass of condensed steam following preheating of the faecal feed
356 decreases as the solids content in the faeces was reduced as a result of the decrease in heat
357 energy required at higher solids contents (Section 3.2). For further processing, the water
358 vapour must be condensed and the condensates sent to the evaporator and sorption stage for
359 recovery of inorganic salts. The mass of ash after combustion of the hydrochar was obtained
360 by multiplying the ash content of the hydrochar following HTC at 200°C for 30 min (about
361 36% d.b., Table 1) by the mass of hydrochar after drying (Table A.2).

362 3.2 Energy balance

363 The energy balance of the HTC process is summarised in Figure 2 and the details
364 presented in Tables A.3 and A.5 in the appendix. The data in Table A.3 are based on the
365 assumption that no heat was recovered from the process or the products, and that the only
366 energy required was that used to heat the reactor to the reaction temperature and for
367 completion of the process; whilst that in Table A.5 are based on energy recovery from the
368 process and the products when the process is in operation. The results clearly indicate that the
369 total amount of energy required to heat feedstock containing a lower amount of solids was
370 higher than that required for feedstock with a higher solids content. This was particularly so
371 for feedstock containing 5% solids. Also, the total energy input to the reactor increases as the
372 quantity of feedstock increases. Although the energy required to heat the reactor increased
373 from 13.92–1458.96 MJ as the quantity of faeces treated per day was increased from 4.00–
374 400.00 kg, there was no change when the solids content in the faeces increased. Of the total
375 energy input, when no heat was recovered from the process or the products the energy
376 required to heat the reactor represented about 63–61% (for 5% solids in faeces), 64–62% (for

377 15% solids in faeces), and 65–63% (for 25% solids in faeces) of the total energy input; that is
378 a slight decrease as the faeces treated per day was increased, and increased as the solids
379 content of the faeces increases (noting that energy input decreased as the faeces solid content
380 increased, as explained earlier). The energy required for heating all the faecal material
381 represents about 21–20% (for faeces with 5% solids), 20–19% (for faeces with 15% solids),
382 and 19–18% (for faeces with 25% solids) of the total energy input to the reactor, that
383 represents an increase with increases in the feed water content. These results are in keeping
384 with those of previous studies. For example, Thorsness [30] found that energy input in the
385 form of steam increased by approximately 15% as the MSW feed water content increased
386 from 25 to 35%. Stemann and Ziegler [29] also reported that the amount of energy required
387 to heat biomass to the reaction temperature depended significantly on the water content of the
388 biomass. In their study on the energetic assessment of the HTC of woody biomass, increasing
389 the water content of the feedstock resulted in increases in energy input of between 2.2% and
390 7.3% of the energy of the hydrochar.

391 Heat losses from the insulated reactor increased as the number of person equivalents
392 increased, but did not change with increases in the solids content and accounted for between
393 about 0.5–3% of the total energy input. This serves to indicate the importance of thermal
394 insulation, and heat losses on a commercial scale may become significant if proper insulation
395 is not provided, and this would adversely affect the overall energy efficiency of the process.
396 Thorsness [30] reported that heat loss effects from the walls of the reactor were significant,
397 with an increase in input steam flow rate requirement of about 40% due to adiabatic
398 conditions. Stemann and Ziegler [29] reported that heat losses from the reactor ranged from
399 0.005–0.2 MW, which accounts for about 0.2% of the system power of the HTC plant, and
400 falls within the range of values obtained in this study. The heat transfer parameters used to

401 estimate the heat loss from the insulated reactor are presented in Tables 2 and A.4 (in the
402 appendix).

403 The heat of reaction measured over an interval of 4 h using the DSC were -0.20 MJ kg^{-1}
404 (± 0.01) at 160°C , -0.32 MJ kg^{-1} (± 0.03) at 180°C , and -0.70 MJ kg^{-1} (± 0.08) at 200°C . The
405 heats of reaction were measured over a period of 4 h, as this was the time previously reported
406 [36] as being the time for complete reaction. Funke and Ziegler [36] reported that the amount
407 of energy released increases under severe carbonisation conditions. In their study because
408 cellulose required severe reaction conditions to carbonise, it took longer for the heat to be
409 released than the 30–40 min reported for wood and glucose. For a treatment time of 30 min,
410 the use of such a value measured in 4 h may represent a slight overestimate but in the overall
411 scheme the error involved would be insignificant. The heat of reaction measured at 200°C for
412 4 h was closer to the value of -0.79 MJ kg^{-1} reported for HTC of digestate from anaerobic
413 digested waste that was estimated based on measured higher heating value (HHV) and
414 combustion reactions [25], but significantly lower than the value of -1.6 MJ kg^{-1} reported for
415 cellulose [5,25], and -1.07 MJ kg^{-1} and -1.06 MJ kg^{-1} for cellulose and glucose using DSC
416 measurements [36]. The reaction heat increased as the amount of faeces undergoing treatment
417 was increased, and was calculated by multiplying the mass of faeces fed into the reactor by
418 the heat of reaction measured at 200°C . The heat of reaction alone cannot sustain the
419 carbonisation reaction as it represents only about 19–20% of the total energy required if the
420 feed had not been preheated (Table A.3 in the appendix), and between 33–35% of the energy
421 if the feed was preheated to 100°C (Table A.5 in the appendix).

422 Energy recovered from the steam in the flash tank increased as the solids content in the
423 faeces and mass of faecal sludge was increased. About 4.1, 0.9, and 0.8% of the total energy
424 recovered from steam was used to dry the hydrochar to approximately 5% moisture content
425 for faeces containing 5, 15, and 25% solids respectively; whilst 65 and 33% of the energy

426 was used to preheat faecal sludge with solid contents of 15 and 25% respectively (Table A.3
427 in the appendix). Preheating the feed reduced the energy required to heat the reactor by about
428 59%. However, energy recovered from steam for faeces containing 5% solids was not
429 sufficient to preheat the feed before it was fed to the HTC reactor. For faeces containing 25%
430 solids, about 63–64% of the total energy from combustion of the hydrochar and methane
431 were used to power the reactor; indicated that the surplus energy could be utilised for other
432 purposes.

433 Alternatively, the energy generated from combustion of the methane and the excess
434 energy recovered from the flashing off of steam alone (71–81%) were sufficient for powering
435 the entire HTC system; hence, the hydrochar can be used for other applications such as
436 addition to soil as a soil conditioner and carbon sequestration or combustion for syngas
437 production. It must be noted that higher methane yields were obtained when the solids
438 content of the faeces was low. Preheating the faeces to 100°C before it was fed to the reactor
439 reduced the heat losses from the reactor to between 50–60%, and also decreased the total heat
440 input required to heat the reactor and faecal content to the reaction temperature of 200°C by
441 59% (Table A.5 in the appendix). Zhao et al. [12] reported that about 48% of the heat
442 generated from hydrochar combustion could be recovered, while the total energy recovery for
443 HTC processing at temperatures above 200°C was approximately between 40% and 60% if
444 the reactor was preheated and when ignoring preheating, respectively. However, this could be
445 lower as heat losses were not considered in their study.

446 3.3. Sensitivity analysis

447 The amount of faeces to be treated, and the concentration of solids in the faeces
448 significantly determined the overall process energetics (Tables A.3 and A.5 in the appendix).
449 The latter were varied as the input parameters in the HTC process from 4.00–400.00 kg per
450 day for the feed rate, and 5–25% for the solids content. For a higher solids content (15–25%)

451 sufficient energy is recovered from flashing off the steam, which can be used to preheat the
452 feed to 100°C and drying the wet hydrochar with 50% moisture to 5%, with excess energy of
453 up to 34% (for 15% solids) and 66% (for 25% solids in faeces). For faeces with 25% solids
454 content energy from combustion of the hydrochar was enough to operate the reactor, leaving
455 a surplus of between 33–35%. This decreased as the amount of faecal waste increased from
456 4.00–400.00 kg/day. For HTC of feedstock with 15% solids, using the energy from
457 combustion of the hydrochar and the surplus energy from steam (10.08–1021.08 MJ per day,
458 Table A.5) would be sufficient to operate the HTC reaction. A feedstock containing 5%
459 solids produces the highest amount of methane (Table A.2 in the appendix), about 125 and
460 167% more than that produced from feedstock with 15 and 25% solids respectively; and
461 consequently generating more energy from its combustion. For all solids contents and
462 feedstock rates, the amount of energy generated from the combustion of methane alone was
463 insufficient to operate the reactor. However, for a feedstock containing 15% solids the excess
464 energy from flashing off of steam and the energy from combustion of both methane and the
465 hydrochar were sufficient to operate the HTC reactor with surplus energy of about 21–22%.
466 Also, for a feedstock containing 25% solids energy from combustion of methane and the
467 hydrochar were sufficient to operate the reactor leaving excess of about 36–37%. It must be
468 noted that the methane yield was based on an empirical estimation and that prediction of the
469 percentage of CH₄ in biogas is difficult and depends on the pH in the anaerobic digestion
470 reactor, which is influenced by the equilibrium CO₂. This is because carbon dioxide is
471 partially soluble in water, and so is partly dissolved in the liquid phase or converted to
472 bicarbonate depending on the pH; but the CH₄ produced is practically insoluble in water and
473 is mostly present in the gas phase. As a result, the estimated CH₄ yield will generally be
474 lower than the fraction of CH₄ in biogas produced from experimental anaerobic digestion
475 tests.

476 **4. Conclusions**

477 The solids contents of the feedstock and the amount of feed material had a significant
478 effect on the material and energy balances of the HTC of faecal sludge. Although feedstocks
479 of lower solids content produced more steam, the steam energy from feedstock with 5%
480 solids was not sufficient for preheating the feed although it was enough for drying the wet
481 hydrochar. In a process where the liquid products were not digested for methane production
482 and for feedstocks containing 15 and 25% solids, once the process has started energy
483 recovery from flashing off steam, and combustion of the char would be sufficient for
484 operating the entire process without the need for any external sources of energy.
485 Alternatively, for a feedstock with 25% solids content and all feed rates, 79–81% of the
486 energy from combustion of methane and the excess energy recovered from flashing off of
487 steam were sufficient for sustaining the process, and the remaining 19–21% could be utilised
488 for other purposes; hence the hydrochar could be used for carbon sequestration when applied
489 to soil or for other applications such as gasification for syngas production. Further
490 investigations would need to be conducted at different reaction temperatures to fully establish
491 the effect of temperature on the energetics of the process. Also, studies into a detailed life-
492 cycle and economic analysis of the process would be useful to confirm the sustainability of
493 the process.

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650 **Appendix**

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Table A.1

Mass balance of faecal sludge HTC as a function of feedstock quantity and solids content

5% Solids in faeces								
Feedstock ^a (kg)	Faecal Sludge (kg)		Flashing (kg)		Dewatering (kg)		Drying (kg)	
	Solids	Water	Slurry	Steam	Solid cake	Liquid	Hydrochar	Evaporated water
4.00	0.20	3.80	3.24	0.60	0.24	3.00	0.12	0.12
8.00	0.40	7.60	6.48	1.44	0.60	5.88	0.36	0.24
40.00	2.00	38.00	32.28	7.08	2.64	29.64	1.32	1.32
200.00	10.00	190.00	161.40	35.28	13.32	148.08	6.84	6.48
400.00	20.00	380.00	322.80	70.56	26.76	296.04	13.80	12.96
15% Solids in faeces								
4.0	0.60	3.40	3.12	0.60	0.84	2.28	0.48	0.36
8.0	1.20	6.80	6.36	1.32	1.56	4.80	0.84	0.72
40.0	6.00	34.00	31.68	6.36	8.04	23.64	4.20	3.84
200.0	30.00	170.00	158.52	31.56	40.08	118.44	20.76	19.32
400.0	60.00	340.00	317.04	63.24	80.16	236.88	41.52	38.64
25% Solids in faeces								
4.0	1.00	3.00	3.12	0.60	1.32	1.80	0.72	0.60
8.0	2.00	6.00	6.24	1.08	2.64	3.60	1.44	1.20
40.0	10.00	30.00	31.08	5.52	13.32	17.76	6.96	6.36
200.0	50.00	150.00	155.64	27.84	66.84	88.8	34.56	32.28
400.0	100.00	300.00	311.16	55.68	133.68	177.48	69.12	64.56

^a On a per day basis, and assuming that the plant operates 12 hours a day. Solids in the faeces reduced by 66.8% following carbonisation.

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Table A.2

Mass balance of faecal sludge HTC resulted from recovered and waste materials

5% Solids in faeces							
Feedstock ^a (kg)	Feed Pre-heating (kg) Condensed Steam	Anaerobic Digestion (kg)			Methane ^b Combustion (kg) Water vapour	Char Combustion (kg)	
		COD	Methane	Waste Water		Ash	Water vapour
4.00	0.28	0.06	0.01	1.44	0.03	0.05	0.01
8.00	0.67	0.13	0.03	2.88	0.06	0.12	0.01
40.00	3.29	0.63	0.14	14.28	0.32	0.48	0.12
200.00	16.41	3.15	0.71	71.64	1.60	2.52	0.36
400.00	32.81	6.31	1.42	143.28	3.19	5.04	0.72
15% Solids in faeces							
4.00	0.28	0.05	0.01	1.20	0.03	0.12	0.02
8.00	0.61	0.10	0.02	2.28	0.05	0.36	0.04
40.00	2.96	0.50	0.11	11.52	0.26	1.56	0.24
200.00	14.68	2.52	0.57	57.36	1.28	7.56	1.08
400.00	29.41	5.04	1.13	114.60	2.55	15.00	2.04
25% Solids in faeces							
4.00	0.28	0.04	0.01	0.84	0.02	0.24	0.04
8.00	0.50	0.08	0.02	1.68	0.04	0.48	0.12
40.00	2.57	0.38	0.09	8.64	0.19	2.52	0.36
200.00	12.95	1.89	0.43	42.60	0.96	12.48	1.68
400.00	25.89	3.78	0.85	85.92	1.91	25.08	3.48

^a On a per day basis, and assuming that the plant operates 12 hours a day. ^b From reaction stoichiometry: 1 kg CH₄ makes 2.25 kg (16/36) kg H₂O.

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Table A.3

Energy balance of faecal sludge HTC without heat recovery

5% Solids in faeces								
Feedstock (kg)	Hydrothermal Treatment (MJ)					Flashing (MJ)		
	Total input	Energy to heat reactor	Energy to faeces & water			Reaction heat	Heat loss	
			faeces	water	total			
4.00	14.28	9.00	0.06	2.88	2.94	2.76	0.48	0.60
8.00	28.68	17.88	0.12	5.76	5.88	5.64	0.72	1.20
40.00	144.84	89.52	0.60	28.68	29.28	27.96	1.80	5.64
200.00	728.40	447.48	3.00	143.28	146.28	139.56	4.92	28.44
400.00	1458.96	894.96	6.00	286.56	292.56	279.00	7.56	56.76
15% Solids in faeces								
4.00	14.04	9.00	0.24	2.52	2.76	2.76	0.48	7.68
8.00	28.32	17.88	0.36	5.16	5.52	5.64	0.72	14.76
40.00	143.04	89.52	1.80	25.68	27.48	27.96	1.80	77.40
200.00	719.28	447.48	8.88	128.16	137.04	139.56	4.92	387.00
400.00	1440.72	894.96	17.88	256.44	274.32	279.00	7.56	773.76
25% Solids in faeces								
4.00	13.92	9.00	0.36	2.3	2.66	2.76	0.48	14.88
8.00	27.96	17.88	0.60	4.6	5.20	5.64	0.72	29.88
40.00	141.24	89.52	3.00	22.7	25.70	27.96	1.80	149.04
200.00	710.16	447.48	14.88	113.2	128.08	139.56	4.92	745.44
400.00	1422.36	894.96	29.76	226.3	256.06	279.00	7.56	1490.76

On per day basis, and assuming that the plant operates 12 hours a day. The faeces are not preheated before fed to the reactor.

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Table A.4

Heat transfer parameters

Feedstock ^a (kg)	L (m)	D (m)	A_{Γ} (m ²)	N_{GR}	N_{Pr}	h_I (w m ⁻² K ⁻¹)	h_M (w m ⁻² K ⁻¹)	h_A (w m ⁻² K ⁻¹)	h_R (w m ⁻² K ⁻¹)	U_{Γ} (w m ⁻² K ⁻¹)
4.00	0.15	0.10	0.09	1.87×10^7	0.77	1.10	534.59	4.06	0.16	0.87
8.00	0.18	0.13	0.14	3.72×10^7	0.77	1.10	534.59	4.14	0.16	0.87
40.00	0.32	0.22	0.36	1.86×10^8	0.77	1.10	534.59	4.32	0.16	0.88
200.00	0.54	0.38	0.97	9.29×10^8	0.77	1.10	534.59	4.51	0.16	0.89
400.00	0.68	0.48	1.51	1.86×10^9	0.77	1.10	534.59	4.59	0.16	0.89

^a On a per day basis, and assuming that the plant operates 12 hours a day.

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Table A.5

Assessment of energy balance of faecal sludge HTC with heat recovery

5% Solids in faeces									
Feedstock (kg)	Preheating ^a (MJ)	Hydrothermal Treatment (MJ)			Flashing ^c (MJ)	Drying (MJ)	Combustion (MJ)		Excess ^d (MJ)
	Steam input	Input ^b	Reaction heat	Losses	Excess energy	Steam input	Char	Methane	
4.00	5.04	8.40	2.76	0.24	Deficit	0.02	2.52	0.71	Deficit
8.00	10.08	16.80	5.64	0.36	Deficit	0.05	5.04	1.42	Deficit
40.00	50.52	84.84	27.96	1.08	Deficit	0.24	25.08	7.11	Deficit
200.00	252.84	426.60	139.56	2.76	Deficit	1.20	125.40	35.56	Deficit
400.00	505.56	854.40	279.00	4.32	Deficit	2.40	250.80	71.12	Deficit
15% solids in faeces									
4.00	5.04	8.28	2.76	0.24	2.52	0.12	7.56	0.57	2.37
8.00	9.96	16.56	5.64	0.36	4.68	0.12	15.00	1.14	4.26
40.00	49.80	83.76	27.96	1.08	26.88	0.72	75.24	5.69	24.05
200.00	249.00	421.20	139.56	2.76	134.52	3.48	376.32	28.44	118.08
400.00	498.12	843.60	279.00	4.32	268.56	7.08	752.52	56.88	234.36
25% Solids in faeces									
4.00	4.92	8.16	2.76	0.24	9.72	0.12	12.60	0.43	14.59
8.00	9.84	16.32	5.64	0.36	19.80	0.24	25.08	0.85	29.41
40.00	49.08	82.68	27.96	1.08	98.76	1.20	125.40	4.26	145.74
200.00	245.28	415.80	139.56	2.76	494.28	5.88	627.12	21.32	727.22
400.00	490.56	832.80	279.00	4.32	988.44	11.76	1254.24	42.63	1452.51

^a Includes energy to heat feed tank, faeces and water to 100°C. ^b Energy from combustion of the hydrochar and methane. ^c Only part of the energy is used to preheat the feedstock and dry the char, and the remainder represents a surplus; "deficit" indicates that the energy is used only to dry the char but is insufficient to preheat the feed.

^d Surplus energy recovered from steam, and from combustion of both char and methane after using part of the energy to operate the reactor. On per day basis, and assuming that the plant operates 12 hours a day.