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Techno-economic analysis of wastewater sludge gasification: A decentralized urban perspective

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HIGHLIGHTS

- Evaluate potential technologies for conversion of waste water sludge to energy.
- Thermal systems analysis of air-blown and steam gasification of waste water sludge.
- Techno-economic analysis of electricity generation from sludge at small-scale plants.
- Air-blown gasification converts sludge to electricity with an efficiency greater than 17%.
- Favorable economics for energy recovery from sludge using air-blown gasification.

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ABSTRACT

The successful management of wastewater sludge for small-scale, urban wastewater treatment plants, (WWTPs), faces several financial and environmental challenges. Common management strategies stabilize sludge for land disposal by microbial processes or heat. Such approaches require large footprint processing facilities or high energy costs. A new approach considers converting sludge to fuel which can be used to produce electricity on-site. This work evaluated several thermochemical conversion (TCC) technologies from the perspective of small urban WWTPs. Among TCC technologies, air-blown gasification was found to be the most suitable approach. A gasification-based generating system was designed and simulated in ASPEN Plus® to determine net electrical and thermal outputs. A technical analysis determined that such a system can be built using currently available technologies. Air-blown gasification was found to convert sludge to electricity with an efficiency greater than 17%, about triple the efficiency of electricity generation using anaerobic digester gas. This level of electricity production can offset up to 1/3 of the electrical demands of a typical WWTP. Finally, an economic analysis concluded that a gasification-based power system can be economically feasible for WWTPs with raw sewage flows above 0.093 m^3/s (2.1 million gallons per day), providing a profit of up to \$3.5 million over an alternative, thermal drying and landfill disposal.

1. Introduction

Wastewater treatment sludge is a dilute mixture of microorganisms, suspended and dissolved organic matter, and mineral species in up to 99% water. Sludge is produced at a concentration of about 0.25 kg/m^3 of solids in mixed municipal and light industrial wastewater treated (Metcalf et al., 2010). In 2005, about 8.2 million dry metric tons of sludge was produced in the United States (Biosolids Generation, 1999). Sludge production was seen to increase 29% faster than the U.S. population growth from 1972 to 1998 (Biosolids Generation, 1999). Management of this process residual can present financial and environmental challenges for wastewater treatment plants (WWTPs). Operators of small urban WWTPs face the greatest difficulties as their operations do not benefit from the economies of scale which permit larger facilities to absorb the costs and plant footprint of anaerobic digestion. This work considers urban WWTPs serving sewage flows of up to 5.3 million gallons per day (MGD) (0.23 m^3/s).

A contemporary approach to sludge management considers sludge to be an income-generating recoverable resource (Murray et al., 2008). In analyzing thermochemical conversion (TCC)

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technologies, it is often useful to know the fuel's heating value, which is the amount of heat released during combustion. The higher heating value (HHV) treats water in the combustion products as a liquid, while the lower heating value (LHV) treats water in the combustion products as a vapor. On a dry basis, sludge has a LHV of about 15 MJ/kg, which is similar to that of a low-rank coal. For a 5.3 MGD plant, up to 825 kW_{th} would be available for conversion to electricity. This suggests that the value of sludge might best be recovered as a fuel for on-site electricity generation. TCC technologies subject sludge to chemical processes at high temperatures to convert the chemical energy in sludge into heat, more useful fuels, or both. However, no small scale (<100 kW_e) techno-economic analyses of TCC-generating systems have been found in the literature.

The objective of this research effort was to estimate the minimum WWTP capacity for which electric power generation by TCC of sludge may be feasible. Feasibility of a process was defined as the ability to produce electrical power on-site using currently available technology while creating a net present worth greater than zero. Four candidate TCC technologies were considered in this study. These include wet oxidation, direct combustion, pyrolysis, and gasification (air blown, steam blown, and supercritical water). The TCC technologies were evaluated with a focus on the unique demands of small-scale, decentralized WWTPs. A sludge fired generating system incorporating the most appropriate TCC process was simulated using the process modeling software ASPEN Plus®. The system was then analyzed for technical feasibility with currently available technology. Data from the simulations were used to inform an economic model that compared the generating system to a base case of thermal drying and landfill disposal, to determine at what scale the TCC-based generating system would be economically feasible.

1.1. Sewage sludge chemical characteristics

Technical fuels are commonly described by their proximate analyses (percentages of moisture, volatile matter, sulfur, fixed carbon, and ash, as well as HHV) and ultimate analyses (percentages of elemental carbon, elemental hydrogen, elemental nitrogen, sulfur, elemental oxygen, ash, and moisture). The composition and properties of the sludge used in this study along with several other resource streams commonly considered as gasification feedstocks are summarized in Table 1. These include wood (pine), corn straw, and municipal solid waste. The most notable differences between sludge and other resource streams are initial moisture, oxygen, and ash content. One challenge in exploiting sludge as a fuel is the need to remove substantial quantities of water. For example, simulations conducted in the course of this study predict that up to 60% of the chemical energy in dried sludge is required for thermal drying. Ash presents other challenges to TCC of sludge, particularly corrosion. The ash compositions of sludges from several different municipal wastewater treatment processes in the Denver, Colorado, area were studied and found to contain 1,830 mg/kg of Na, 5350 mg/kg of K, and 19,600 mg/kg Ca (averaged between all samples) (Ramey et al., 2014). Ash can be aggressively corrosive at high temperature (Cummer and Brown, 2002); however, ash high in Na and K salts may provide a benefit to the system by catalyzing tar cracking reactions at temperatures above 500 °C (Fonts et al., 2012). A further complication of sludge not represented by proximate and ultimate analyses is heavy metals content, including Cd, Hg, Pb, and Zn. At flame temperatures achieved during sludge or char combustion, metals may vaporize and be entrained with the exhaust gas stream. Metals remaining in ash and char may also be leached by acidic water if disposed of in a landfill (Fytili and Zabaniotou, 2008). Furthermore, due to the presence of fats, oils, and greases, the mass fraction of oxygen in sludge is somewhat lower than other biomass materials, which partially offsets the reduction in mass-basis heating value caused by high ash content.

1.2. Thermochemical conversion processes

TCC processes are routinely used to transform solid fuels into heat or a more valuable technical fuel such as synthesis gas or bio-oil (Fytili and Zabaniotou, 2008; Werther and Ogada, 1999; Bridgwater, 2003). For consideration in this study, candidate TCC technologies were required to be at least in the pilot plant stage of development and able to support the production of electricity on-site. TCC technologies considered in this study include wet oxidation, direct combustion, pyrolysis, and gasification. Selection of the most appropriate TCC process was also based on process energy efficiency, environmental considerations, and applicability to small scale installations. A short summary of these technologies with a view to decentralized WWTPs is presented in the following paragraphs.

1.2.1. Wet oxidation

Wet oxidation is an exothermic process which takes place in an oxygenated aqueous phase at pressures ranging from 5 to 30 MPa (725–4350 psi) and temperatures from 150 to 600 °C (Khan et al., 1999). Organic species are oxidized to H_2O, CO_2, volatile fatty acids, and simpler organic compounds such as formaldehyde (Khan et al., 1999). The effluent solution must be recycled to the WWTP headworks to treat the reaction products (Khan et al., 1999). Wet oxidation in wastewater treatment is used to decompose organic materials which are resistant to biological treatment processes.

Table 1
Chemical composition of sludge compared to other common waste streams. Data are dry wt.% unless otherwise indicated

Analysis	Wastewater sludge[a]	Pine wood[b]	Corn straw[c]	Municipal[d] solid waste
Initial moisture (Wet basis)	80 (Dewatered)	12	6.17	8.8
Fixed carbon	9.03	16	13.75	11.79
Volatile matter	71.3	71.5	75.95	82.8
Ash	19.67	0.5	5.93	5.98
C	42.92	51.6	43.83	51.81
H	6.04	4.9	5.95	5.76
O	24.51	42.6	45.01	30.22
N	5.91	0.9	0.97	0.26
S	0.95	Not detected	0.13	0.36
LHV [MJ/kg]	16.7	20.2 (HHV)	17.75	21.3

[a] Ramey et al. (2014).
[b] Franco et al. (2003).
[c] Gai and Dong (2012).
[d] He et al. (2009).

Wet oxidation has been used to produce process heat, but there is limited information in the literature on electricity generation from wet oxidation. A wet oxidation process in a U.S. wastewater plant was capable of supplying process heat at 115.5 °C; sufficient for heating mesophilic digesters (Griffith and Raymond, 2002).

1.2.2. Hydrothermal carbonization

The hydrothermal carbonization process describes a complex set of reactions occurring in non-oxygenated water under pressures up to 2 MPa and temperatures up to 400 °C which transform organic feedstock into a carbonaceous char commonly called hydrochar or synthetic coal. Hydrolysis reactions cleave large organic molecules and biomolecules; decarboxylation and dehydration reactions remove oxygen, nitrogen, and some hydrogen from carbon backbones. The simpler carbon containing molecules then tend to polymerize into a poly-aromatic, hydrophobic solid with hydrogen/carbon and oxygen/carbon ratios similar to those of lignite (He et al., 2013). The product contains less nitrogen, sulfur, ash, and soluble metals than the feedstock, and can be recovered as a precipitate and used as a solid fuel. The process may generate significant wastewater requiring further treatment, especially for removal of heavy metals. Hydrothermal carbonization is especially suited for wet feedstock: little or no energy is required for dewatering, and the low activation energies of the hydrolysis reactions allows the process to occur at relatively low temperatures (He et al., 2013). Thus, hydrothermal carbonization promises to be an energy efficient TCC technology, especially for wet feedstocks such as sludge. While hydrothermal carbonization has been considered as a candidate sludge stabilization process for some time, only recently has research focused on producing solid fuel from sludge.

1.2.3. Combustion

Direct combustion involves burning sludge, usually pre-dried, in a furnace to decompose organic material, concentrate ash, and produce high temperature flue gas (Werther and Ogada, 1999). Direct combustion temperatures can range from 800 to 2000 °C, typically at atmospheric pressure (Werther and Ogada, 1999). Sludge incineration has a long history as a waste stabilization mechanism, although examples of electricity generation from incineration in small scale installations have not been found. One of the greatest challenges in sludge combustion is stack emissions control. Emissions include fly ash, SO_x, NO_x, HCl, Hg, Cd, dioxins, and furans (Fytili and Zabaniotou, 2008). Remediation of these emissions requires capital intensive stack gas cleaning facilities (Van Caneghem et al., 2012). Furthermore, heat exchange surfaces may be fouled by metals and alkali salts vaporized during combustion (Werther and Ogada, 1999). Another possible mode of sludge combustion, co-combustion, involves burning dried sludge as an amendment to a utility fuel such as coal. Because co-combustion does not produce on site power it was outside the scope of this analysis.

1.2.4. Pyrolysis

Pyrolysis processes subject sludge to temperatures between 450 and 650 °C in an anoxic environment, usually circulating flue gas or pyrolysis gas. The process produces three fractions, a solid char, oily water-miscible liquid (bio-oil), and combustible gas (Fytili and Zabaniotou, 2008). Reactor conditions and fuel composition can influence how the products partition into each phase. Ash tends to catalyze cracking of oil precursors into gas species. The high ash content of sludge results in maximum oil yields of about 50 wt% (Inguanzo et al., 2002). The slow kinetics of char gasification at the relatively low temperatures needed for maximum oil yield results in char yields often as high as 30 wt% (Fonts et al., 2012). Pyrolysis processes are typically operated to maximize bio-oil yield for upgrading to transportation fuel. The oil, once refined, can be stored and transported (Wright et al., 2010).

1.2.5. Gasification

Gasification is the process of converting carbonaceous feedstock into a combustible synthesis gas, or syngas, by partial oxidation in a net reducing atmosphere at high temperature. A complex series of reactions typically produces syngas composed of H_2, CO, CH_4, CO_2, H_2O, N_2, minor species, and tars (Franco et al., 2003). Wastewater sludge gasification takes place at close to atmospheric pressure and temperatures from 700 to 1000 °C (de Andres et al., 2011; Nipattummakul et al., 2010).

Three oxidants are commonly used in gasifiers: air, pure oxygen, steam, or a mixture of steam with either air or oxygen. A major disadvantage of air-blown gasification is that atmospheric nitrogen acts as a diluent, reducing the syngas heating value. Steam blown gasifiers require an external heat source due to the overall endothermic nature of the process; however, the main advantage of steam blown gasification is a greater syngas LHV of 15–20 MJ/kg. Additionally, steam gasification tends to show superior carbon conversion; defined as the ratio of reactant carbon to carbon in the syngas, $\eta_C = m_{\text{carbon,gas}}/m_{\text{carbon,feed}}$ (Franco et al., 2003). Oxygen blown gasification produces a gas with intermediate LHV, 10–12 MJ/kg (Bridgwater, 2003), and is generally the cleanest syngas with regard to tars. Oxygen blown reactors do not have the heating demands of raising steam to reactor temperature, although the oxygen separation plant can consume a substantial quantity of energy. Oxygen blown gasification was not considered in this analysis due to cost and technology uncertainties associated with the scaling of air separation apparatus for small systems.

Gasification in supercritical water is another sludge conversion process that may eventually be viable for syngas production. Supercritical water gasification processes operate above the critical point of water, 22.1 MPa (3205 Psi) and 374 °C. Actual operating temperatures may approach 1000 °C. The solubility of organic material in supercritical water improves contact with the gasification medium, enhancing product yield. Supercritical water gasifiers produce negligible tar and boast η_C greater than 90% (Fiori et al., 2012).

Air-and steam-blown gasification were chosen for further analysis due to favorable operating characteristics. The relatively low temperatures used for biomass gasification reduce the vaporization of inorganic salts and heavy metals in comparison with direct combustion processes. The chemical environment of a gasifier also inhibits dioxin, furan, NO_x, and SO_x formation. In comparison to sludge pyrolysis, gasification partitions most of the feedstock potential energy into a single syngas stream which can be prepared as an engine fuel using simpler means than those needed for bio-oil. Furthermore, gasification has been successfully applied to a variety of feedstocks, including sludge at the laboratory scale (Arjharn et al., 2013). No clear advantage is found between air and steam blown gasification from a qualitative comparison. The greater LHV of steam process syngas may be offset by the heating duty to raise steam to reactor temperature. Therefore, both alternatives were simulated to compare net power output and technical feasibility.

2. Methods

Effective thermal system design must consider process energy consumption, energy recovery, emissions, and cost. In the case of wastewater sludge, it is expected that the energy requirement to dry sludge and heat it to reactor temperature will constitute the bulk of process energy demands. Simulation results reveal that up to 75% of the energy content of sludge is required to dry sludge from 80 to 10 wt% moisture in a convective dryer and then heat it to 850 °C. Thus, thermal integration is of primary importance to system design. However, emissions and cost concerns may constrain design options. Char combustion approaches to gasifier heating, such as chemical looping combustors, present the same stack gas clean-up challenges as direct combustion (Van Caneghem et al., 2012; Inguanzo et al., 2002). Therefore, char combustion approaches to gasifier heating were not considered.

A system process flowsheet, shown in Fig. 1, was developed to incorporate four process areas: convective drying, gasification, gas cooling and cleaning, and an engine-generator. Sludge is dried by direct contact with hot exhaust gases recycled from the engine, syngas cooler, and burners. Dried sludge is briquetted (pressed into pellets or bricks), then fed to the gasifier where it is converted to syngas, char, and ash. Water present in the semi-dry sludge is converted to steam within the gasifier. Hot syngas is cooled by air in a gas-to-gas heat exchanger and then scrubbed in a direct contact water scrubber to condense tars and remove acidic gases. The cool, scrubbed syngas stream is used to fire an engine-generator; however, a portion of the syngas can be combusted in a utility burner for dryer heating. The system recovers heat by recycling all exhaust gas and hot gas from the syngas cooling heat exchanger into the convective dryer. Char and ash are removed from the gasifier at gasification temperature and rejected from the system as waste (not shown).

2.1. Assumptions and limitations

This analysis was intended to determine the feasibility of producing net electrical power and heat from waste water solids under steady-state conditions. The system boundary crosses sludge feed, dryer exhaust, char and ash waste, and net electrical work streams. Sludge was assumed to enter the system at 80 wt% water content from upstream mechanical dewatering by centrifuge. The energy consumption of the centrifuge was excluded from the energy balance because this hardware is already in place for the thermal drying necessary for sludge stabilization in the base case. Sludge composition was taken from Ramey et al. (2014) and is shown in Table 1. It was assumed that all components represent real mechanical equipment which is optimized for the mass and energy flows at every point of analysis. Surface heat losses from all components except the gasifier and engine were neglected. All components which are unlikely to contribute substantially to the overall plant energy balance were not specifically modeled; plant parasitic power was assumed to be 10% of gross power production. Pressures for all components except the engine were set to atmospheric pressure (0.1013 MPa).

2.2. Modeling and simulation

System simulation was implemented in the chemical engineering process modeling software ASPEN Plus®. Thermophysical properties for all components except sludge and ash were estimated by the Boston-Matthias modified Peng–Robinson equation of state (Damartzis et al., 2012). Sludge and ash thermophysical properties were estimated by the ASPEN Plus internal coal models *HCOALGEN* and *DCOALIGT*, which are also appropriate for biomass.

2.2.1. Dryer

Sludge drying assumes that free water is evaporated and the sludge is unaffected, except for a change in temperature. Arlabosse et al. (2005) suggested that energy needed to dry sludge is nearly that of the enthalpy of vaporization of pure water for sludge water content in the 10–99 wt% (wet basis) range. The sludge dryer consists of a stoichiometry reactor, *RStoic*, to convert some weight of wet sludge to water, vaporize it, and then a flash operation to separate the vapor from dry solids.

2.2.2. Gasifier

In keeping with common thermodynamic equilibrium methods (Puig-Arnavat et al., 2010) a gasifier model was constructed of two principal unit operations, a yield reactor,*RYield*, and a Gibbs equilibrium reactor, *RGibbs*. The yield reactor models fuel pyrolysis by decomposing sludge into its constituent elements, C, H, O, N, S, ash, and water at gasification temperature.A separation block after *RYield* separates a fraction of the hot reactant carbon into a waste char stream to simulate incomplete carbon conversion. *RGibbs* determines an equilibrium distribution of reactants and products by calculating an equilibrium constant for each reaction in a linearly independent set. The three-reaction global chemistry model used to simulate gasification is given by Reactions (1)–(3).

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \tag{1}$$

$$C + H_2O \rightleftharpoons CO + H_2 \tag{2}$$

$$C + 2H_2 \rightleftharpoons CH_4 \tag{3}$$

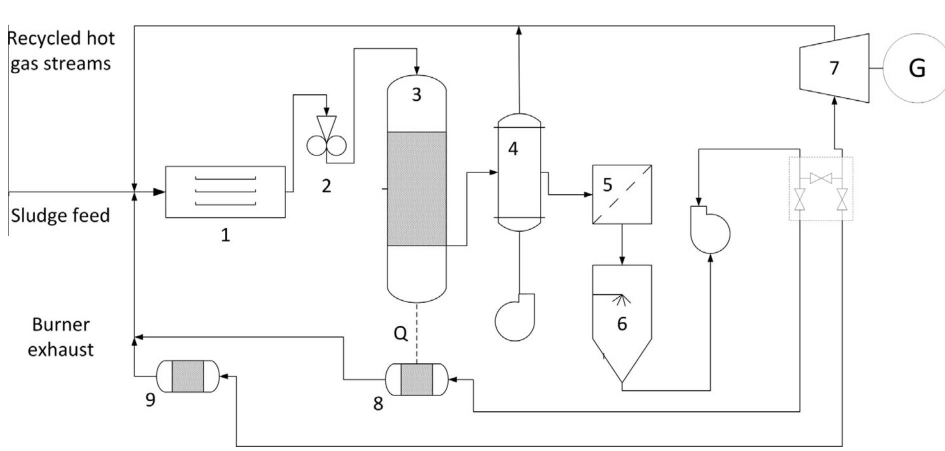

Fig. 1. System flowsheet. 1: dryer, 2: briquetter, 3: gasifier, 4: syngas cooler, 5: filter, 6: wet scrubber, 7: engine generator, 8: gasifier heating burner, 9: dryer heating burner, Q: heat supplied to gasifier from burner.

Equilibrium is restricted by assigning a reaction temperature approach, $T_{equilibrium} = T_{reactor} + \Delta T_{approach}$ to each reaction. Restricting equilibrium accounts for kinetic limitations in real systems and allows for calibration to experimental data.

Overall, gasification chemistry is endothermic. The energy balance for the air-blown system is satisfied by admitting air into *RGibbs*, simulating partial combustion of the fuel. *RGibbs* then transfers heat to *RYield* at reactor temperature. Air flow is varied until the heat duties of *RYield* and *RGibbs* sum to zero. This approach was implemented through a *DesignSpec* block which dynamically calculates an equivalence ratio (ER), allowing only the minimum amount of fuel to be burned. For the steam blown system, a similar *DesignSpec* block diverts some of the product syngas to a burner. Heat is transferred from hot flue gas to *RGibbs* at 20 °C above reactor temperature. In both cases, 3% of the reactant sludge power as $\dot{m}_{sludge}LHV_{sludge}$ is added to the heat duty of *RGibbs* to account for surface heat losses (Doherty et al., 2009).

2.2.3. Gas processing

Syngas exiting a gasifier must be cleaned to remove fly ash, char, and vapor phase tars. Syngas may be acceptably cleaned to fire an internal combustion engine by bag filtration and a water scrubber (Anis and Zainal, 2011). Syngas is first passed through a 20 °C approach heat exchanger where it is cooled to 150 °C. In order to avoid assuming a particle size distribution and pressure drop, no bag filter was modeled. Instead, solid carbon and ash are removed in separation blocks. A water scrubber then cools the gas to 70 °C, which is below the condensation point of most tars (Cummer and Brown, 2002). Water use by the scrubber was not modeled in detail, only the amount of water necessary to cool the syngas to 70 °C is circulated in a closed loop. In practice, a certain amount of water will be wasted from the scrubber. Exact water make-up and wastewater flows will depend upon the scrubbing apparatus used; although the simulated water consumption needed to cool the gas was on the order of 0.005 kg/s for a 5 dry metric ton/day system. It may be possible to skim condensed tars off the surface of scrubbing water in a surge tank and recycle them to the gasifier, thereby improving efficiency.

2.2.4. Engine generator

ASPEN Plus does not provide blocks which can accurately simulate a reciprocating internal combustion engine. In particular, the only fluid expander, a turbine block (*Turb*), cannot operate in positive-displacement mode. Thus, the engine was modeled as a gas turbine (which is easily approximated) and calibrated to the performance of a real spark ignition engine burning wastewater sludge syngas.

The engine model was built using an assembly of compressor (*Compr*), combustor (*RGibbs*), and expander (*Turb*) blocks. *Compr* was specified as an adiabatic, positive displacement compressor with a 10:1 compression ratio. Combustion was modeled in an adiabatic *RGibbs* block operating at the compressor outlet pressure. A *Calculator* block was used to fix a stoichiometric air to fuel ratio for any syngas composition. The turbine was specified as an isentropic turbine operating adiabatically, exhausting into 0.1013 MPa. Heat loss to the coolant was assumed to be equal to shaft power output (Pulkrabek, 2004). Additionally, heat lost to the environment from hot engine surfaces was assumed to be equal to 10% of the shaft power. Coolant was directed to a 20 °C approach heat exchanger preheating air for the syngas cooler. Heat lost to the coolant and environment was extracted from the turbine exhaust.

In order to approximate the performance of a real engine using this model, a calibration was performed by adjusting the polytropic efficiency of *Compr* and isentropic efficiency of *Turb* until engine efficiency matched the experimental data (Szwaja et al., 2013) using the experimental syngas composition and equivalence ratio. The mechanical efficiency of both components was set to 87%, a representative figure for low speed engines (Pulkrabek, 2004). Engine calibration was then held constant throughout the simulations, allowing efficiency to vary with model gas composition. Electricity generation by an induction alternator was assumed to be 99% efficient.

2.3. Economic feasibility

Economic feasibility was evaluated by calculating the present worth (PW) of the gasification plant. Present worth discounts net cash flows over the lifetime of the plant to some specified point in time, taken to be analysis year 2010. Positive PW indicates an economically feasible project (Gribik et al., 2007). Net cash flow was calculated by considering an annual cost difference between the gasification plant and a base case. The base case was thermal stabilization in a natural gas fired dryer. The centrifuge described in Section 2.1 would likely be required by the thermal drying process to reduce dryer size and cost (Werther and Ogada, 1999). Thus, the base case was composed of a centrifuge, wet sludge conveyance equipment, and a natural gas fired dryer. This choice represents a likely sludge management scheme for urban wastewater treatment because odor and footprint concerns restrict the ability of small urban WWTPs to stabilize sludge by digestion or solar drying. Both the gasifier plant and thermal dryer were expected to operate at 90% uptime. Net annual cash flow, F_n, is the difference between the annual cash flows of the gasification plant and the base case:

$$F_n = \begin{aligned}&[(R_{elec} + R_{fuel}) - (C_{capital} + C_{mfg})]_g \\ &-[(C_{capital} + C_{mfg})]_b\end{aligned} \tag{4}$$

where R_{elec} and R_{fuel} are positive valued revenues for electricity and natural gas, respectively, and $C_{capital}$ and C_{mfg} are negative valued costs for financing and manufacturing, respectively. Subscript g refers to the gasifier plant and b refers to the base case. As the gasification plant incurs negligible variable cost (no supplemental fuel, etc.), its costs were considered to be only capital investment and finance charges. Gasification plant revenues include savings from on-site power generation and relieved dryer fuel costs. As thermal stabilization produces no value-added products, its annual cash flow was only its costs. Capital investment and finance charges for the drying equipment were assumed to be identical for both cases. Thus the capital costs of the drying equipment cancel from both terms in Eq. (4), and the cash flow of the base case was only the purchase price of natural gas. On an annual compounding basis, PW of a project is given by Peters et al. (2003)

$$PW = \sum_{n=1}^{k} F_n(1+i)^{-n} \tag{5}$$

for a project evaluated every year n for k years. Plant operating life was expected to be 20 years. Variable i represents the minimum acceptable rate of return (MARR). Similar to previous studies, we chose a MARR value of 10% (Gribik et al., 2007).

Cost estimation proceeded according to the percentage of delivered equipment costing method which assumes that the total cost of building and commissioning a plant (the total capital investment, or TCI) is a function of the total purchased equipment cost (TPEC). Purchased equipment was taken to mean major process equipment items such as the gasifier, dryer, engine-generator, etc. Cost estimates by this method can be considered accurate to ±30% (Peters et al., 2003).

First, a set of empirically determined cost factors were multiplied by TPEC to estimate direct, indirect, and working costs. Direct costs are equipment installation, construction supplies, and other hardware and construction related costs. Indirect costs account for engineering, supervision, contingency costs, and construction expenses such as tools and temporary facilities. Contingency costs in process plant construction represent funds set aside for any unexpected cost during plant construction and commissioning. Working capital accounts for capital held in the production process including raw and finished material in stock, cash on hand, and other business related capital. Cost factors used in this analysis are summarized in Table 2. Each of these costs are then added to TPEC to determine TCI (Eq. (6)).

$$\mathrm{TCI} = \mathrm{TPEC}\left(1 + \sum C_{\mathrm{direct}} + \sum C_{\mathrm{indirect}} + \sum C_{\mathrm{working}}\right) \quad (6)$$

In Eq. (6), C represents cost factors. The cost factors were chosen to reflect the costs of a small scale generating plant (Peters et al., 2003). Neither the gasification plant nor the base case were taken to hold salable raw material or product in inventory. In the context of a wastewater treatment plant it was assumed that the gasification plant will contribute negligible non-labor overhead to that already accounted for in the base case. Commercial experience (Community Power Corporation, 2012) suggests that a gasifier system may demand less than eight hours per day of operations and maintenance labor and it was expected that a gasification plant will occupy the same building facilities needed by a thermal stabilization system. Thus, only limited additional construction is necessary, reflected in the small building and yard improvements factors. Finally, municipal WWTPs in the United States are tax exempt. Tax and depreciation are thus neglected.

Major equipment costs for all components except the gasifier were determined from literature sources (Wright et al., 2010) and supplier quotations. Costing a small-scale downdraft gasifier was challenging. Little literature exists on the techno-economics of small-scale biomass power, perhaps reflecting limited industrial experience with small scale TCC. Gasifier cost was estimated from a large scale atmospheric fluidized bed reactor (Wright et al., 2010). A fluidized bed reactor is likely to be more expensive than a downdraft reactor, making this substitution conservative. Costs were scaled by capacity via Eq. (7) (Wright et al., 2010), where C_{new} is the scaled cost, C_o is quoted cost, $\mathrm{Capacity}_o$ is the quoted capacity, and $\mathrm{Capacity}_{\mathrm{new}}$ is the capacity the cost is scaled to.

$$C_{\mathrm{new}} = C_o\left(\frac{\mathrm{Capacity}_{\mathrm{new}}}{\mathrm{Capacity}_o}\right)^{0.6} \quad (7)$$

Table 2
Cost factors used to compute total capital investment, TCI.

	Cost factor
Direct costs	
Installation	0.2
I&C	0.26
Piping	0.2
Electrical	0.2
Buildings	0.1
Yard improvement	0
Service facilities	0
Indirect costs	
Engineering	0.03
Construction	0.01
Legal	0.01
Contractor's Fee	0.19
Contingency	0.37
Working Capital	0.05
Fixed operating costs	
Maintenance	0.2
Laboratory	0.05
Supervision	0.15 × Labor
Overhead	0.25 × (Labor + Maintenance + Supervision)
Insurance	0.0125
Taxes	0
General Expenses	0.05

3. Results and discussion

3.1. Model calibration and verification

Gasification modeling suffers from the fact that in real systems gasification is a non-equilibrium process. Without known kinetics or model calibration it is impossible to predict syngas composition. Considering that this model is more concerned with syngas LHV than composition, calibration using the restricted equilibrium method discussed in Section 2 is an appropriate approach (Puig-Arnavat et al., 2010). A three-reaction restriction was considered a logical choice, since the eight major carbon gasification reactions comprise a linearly independent set of only three reactions (those used in the gasification model) (Franco et al., 2003). Five variables are required to specify the gasifier. These include temperature, pressure, fuel moisture, η_c, and $\Delta T_{\mathrm{approach}}$. The model was calibrated by substituting sludge composition, gasifier temperature, and fuel moisture with values found in the literature. The quantity of air entering the air-blown gasifier was set dynamically in the model in order to satisfy the gasifier energy balance. However, for model calibration, the air-blown process equivalence ratio (ER) was set to the experimental value. For the steam process, the steam/fuel ratio was accounted for in the fuel moisture setting.

Carbon conversion efficiency, η_c, is a function of various parameters, including temperature, pressure, fuel moisture, fuel composition, reactor design, and other factors. Data from the literature (de Andres et al., 2011; Kang et al., 2011) show a roughly linear increase in η_c with temperature up to approximately 850 °C and little temperature dependence above 850 °C. Carbon conversion efficiency was observed to be largely unaffected by other parameters including fuel moisture and ER. Thus, η_c was simplified in this model to depend only on temperature, and was specified as 60%, 75%, 80%, 88% (air-blown process) and 70%, 75%, 85%, 93% (steam process) at reactor temperatures of 700, 750, 800, and >850 °C, respectively, using a linear curve fit. Error in functionalizing η_c and gas composition as purely temperature dependent is expected to be only approximately 3% (de Andres et al., 2011) for wet fuel.

Gasifier chemistry was calibrated to fit the literature data by varying $\Delta T_{\mathrm{approach}}$ for Reactions (1)–(3) in an iterative procedure to best fit model predictions to literature data over the temperature range of interest (700–1000 °C). $\Delta T_{\mathrm{approach}}$ for Reactions (1)–(3) was found to be 55, 202, −55 °C for the air-blown process and 0, −290, −45 °C for the steam process respectively. Gasifier calibration was validated by comparing model gas composition to experimental data (de Andres et al., 2011; Nipattummakul et al., 2010) at various gasifier temperatures. Validation runs were conducted using the proximate and ultimate analyses of the sludge used in the cited experiments (Table 3). A sum of error squared method was used to evaluate the simulation results. Error for a set of N data points is defined as:

Table 3
Chemical composition of sludges used in model validation.

Analysis	Air Process[a]	Steam Process[b]
Fixed carbon	7.0	21.8
Volatile matter	53.7	44.3
Ash	41.3	33.9
C	28.7	45.8
H	4.8	2.99
O	19.8	14.7
N	4.5	1.49
S	0.90	1.11

[a] de Andres et al. (2011).
[b] Nipattummakul et al. (2010).

$$RE = \sum_{i=1}^{N} \left(\frac{\chi_{e,i} - \chi_{m,i}}{\chi_{e,i}} \right)^2 \quad (8)$$

where χ is a mole fraction or heating value, e refers to experimental data, and m refers to model data. Mean relative error is given as $\sqrt{RE/N}$.

Calibration results are shown for the air-blown process in Fig. 2a and for the steam process in Fig. 2b. The air-blown gasification model generally validates well, capturing the trend of gas evolution over the temperature range investigated. CH_4 shows the highest relative error, 112%, while the errors of the remaining species were less than 25%. Thermodynamic equilibrium methods tend to over predict CH_4, yet remain the best modeling choice for interrogating system designs at an early stage (Puig-Arnavat et al., 2010). The steam model also reproduces the experimental trends with good accuracy except for CH_4, which had an error of 78%.

In addition to the overall composition, calculated syngas LHV was compared to the experimental data. Syngas LHV represents the chemical potential energy of the syngas stream, and is more important to a thermodynamic analysis than gas composition (Puig-Arnavat et al., 2010). For air-blow gasification, the error in modeled syngas LHV was only 6.7%. The steam process LHV was also in good agreement with the experimental data at 4.3% relative error (Fig. 2c). From this data it can be concluded that the restricted equilibrium model is capable of simulating air and steam blown gasifier chemistry with a degree of accuracy appropriate for a thermodynamic analysis.

While directly applicable data were found to validate the gasifier model calibration, no such data were found for the engine model calibration. The major assumption used in calibrating the engine model is that the calibration holds for any syngas composition. Unfortunately, studies were not found which examine spark-ignition engine efficiency as a function of syngas composition. Instead, a sensitivity analysis was performed to determine the extent to which assumed engine efficiency parameters affect net power output. Three efficiencies were used in the engine model: polytropic compression efficiency, isentropic expansion efficiency, and mechanical efficiency. As compression efficiency was expected to affect shaft power the least, due to the relatively small contribution of compression work compared to expansion work, the compressor efficiency was guessed to be 85%. Variation of compressor efficiency by ±5% results in nearly 11% increased (or decreased) compressor work. However, the engine shaft power changes by only 0.9% due to the lower contribution of compression work to shaft power. Expansion efficiency has a greater effect. A 5% drop in efficiency from the calibrated value of 89% results in 13.8% less shaft power. Conversely, an increase in efficiency by 5% raises shaft power by 11.3%. Mechanical efficiency is linear, a 1% change in mechanical efficiency results in a 1% change in shaft power. These sensitivity data demonstrate that the engine model is not overly sensitive to its calibration parameters.

Gasifier operating temperature and fuel moisture were determined by parametric studies maximizing net electrical output. Both systems show the greatest net electrical output at 850 °C, the temperature for which η_c is maximized. Sludge moisture optimizations revealed the expected result that output is inversely proportional to sludge moisture. Net electrical output of the air process peaks at 10 wt% water content, the lowest value studied; however, net electrical output is nearly flat from 10–15 wt%. Optimized sludge moisture for the steam process is found to be 28%, corresponding to a molar carbon to oxygen (C/O) ratio of unity for the sludge studied. Operation of a steam gasifier at C/O > 1 may reduce η_c, while operation at C/O < 1 reduces gasification efficiency by over-oxidizing the fuel.

3.2. System performance

Net electrical output as a function of dry sludge flow rate is linear for both processes (Fig. 2d). The maximum net electrical output for air-blown gasification of 153 kW occurs at a dry sludge feed rate of 0.052 kg/s (5 dry metric tons/day), which corresponds to a WWTP capacity of 5.3 MGD. System performance was calculated

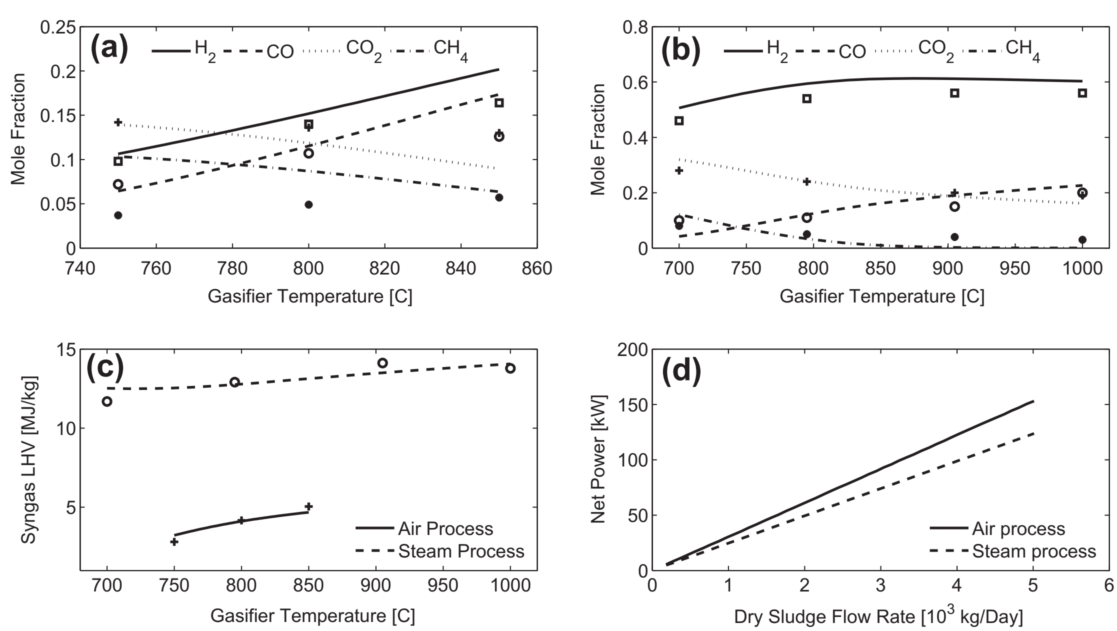

Fig. 2. Model validation and net power output. Gas composition and LHV predicted by the models is compared to experimental literature data for the indicated gasification temperature. Experimental composition data are shown as markers; □ H_2, ○ CO, + CO_2, ● CH_4. (a) air-blown process composition. (b) steam process composition. (c) Dry mass basis LHV (including N_2, CO_2). All validations conducted using the operating conditions (T, P, η_c , ER, steam/sludge ratio) and sludge composition (Table 3) of the experimental investigation (de Andres et al., 2011; Nipattummakul et al., 2010). (d) Net electrical power output of the air-blown and steam-blown processes as a function of dry sludge flow rate (0 % moisture).

as a function of fully dry, 0 % moisture, sludge flow rate in order to provide results which are independent of the level of sludge drying. This approach also allowed conversion from dry sludge flow rate to WWTP plant flow rate, using the assumed solids concentration of 0.25 kg/m^3 of raw sewage (Metcalf et al., 2010).

From the perspective of the overall WWTP, the air blown gasification-generation process reported here can satisfy about one-third of the electrical demand of an aerated activated-sludge process according to the typical process energy consumption figures reported in (Metcalf et al., 2010). Dryer exhaust temperatures are found to be 64 °C for the air-blown process and 97 °C for the steam process; both too low to export useful heat. However, the dryer energy balance is nearly satisfied by the recycled exhaust streams.Hot recycled exhaust gas provides sufficient heat to the dryer that the air-blown process burns only 2% of the produced syngas for additional drying heat; the steam process drying heat demand is entirely satisfied by recycled hot gas streams.

The air-blown process is seen to outperform the steam process by 26% in terms of net power output. This difference is greater than the modeling error in syngas LHV, suggesting that this conclusion is robust. System efficiencies, calculated as $\dot{W}_{\text{electrical}}/(\dot{m}_{\text{sludge}}\text{LHV}_{\text{sludge}})$, are 17.5% for the air-blown process and 12.3% for the steam process. The air-blown process efficiency is greater than small-scale steam Rankine cycles (Bridgwater, 2003) and about triple the performance of electricity generation from anaerobic digester gas.

The lower efficiency of the steam process is partially due to the thermodynamics of indirect heating. The model assumed heat from a syngas burner was transferred at 20 °C above reactor temperature. This process is inherently inefficient as heat exchange with a body at 850 °C produces a flue gas stream of temperature higher than 850 °C. Calculating a gasifier heating efficiency as $\eta_{\text{burner}} = \dot{Q}_{\text{burner}}/(\dot{m}_{\text{syngas}}\text{LHV}_{\text{syngas}})$, where $\dot{Q}_{\text{burner}}$ is the heat transferred from hot combustion gases to the gasifier and $\dot{m}_{\text{syngas}}$ refers to the mass flow of syngas diverted to the burner, reveals a heating efficiency of only about 56%. The air process generates heat internally by combusting a portion of the fuel. The simulated equivalence ratio was found to be 34%; or about 34% of the sludge must be burned to heat the gasifier.

Cold gas efficiency (CGE), Eq. (9), is a figure of merit which relates the ability of the gasifier to convert chemical energy of the fuel to chemical energy in the syngas. Because the air process introduces a nitrogen diluent, which reduces the mass basis LHV of syngas, it is reasonable that its CGE of 72% is less than that of the steam process, 88%.

$$\text{CGE} = \frac{\dot{m}_{\text{syngas}}\text{LHV}_{\text{syngas}}}{\dot{m}_{\text{sludge}}\text{LHV}_{\text{sludge}}} \tag{9}$$

Syngas composition predicted by the restricted equilibrium model is presented in Table 4 in mole percent. Air process gas is diluted by 47% N_2, a consequence of admitting air into the reactor. The fuel species, H_2, CO, and CH_4, contribute only 15.1%, 17.8%, and 1.2% of the syngas, respectively, giving a LHV of 4 MJ/kg. In contrast, the steam process syngas was found to be 49.1% H_2, 29.7% CO, and 7.0% CH_4 with a LHV of 17 MJ/kg. Air process syngas flowrate, on a mass basis, is about 2.5 times greater than that of the steam process (mostly due to the N_2 diluent). Modeling data confirmed that from a systems perspective, the air-blown system is substantially superior. A comparison of system performance indicators is given in Table 4.

Table 4
System performance indicators, simulated gas composition (mole%), and economic indicators (*economic results in thousand USD).

Technical results			
Indicator	Air process	Steam process	
Max. Net Power (kW)			
(5.3 MGD)	153.0	123.7	
η_{system} (%)	17.5	12.3	
CGE (%)	72	88	
$\text{LHV}_{\text{syngas}}$ (MJ/kg)	4.0	17.0	
ER (Air process) (%)	34	N/A	
Burner gas (%)	2 (Dryer)	54.2 (Gasifier)	
Composition			
H_2	15.1	49.1	
CO	17.8	29.7	
CO_2	7.0	7.4	
CH_4	1.2	7.0	
H_2O	11.2	3.6	
N_2	47.7	2.7	
Economic results			
WWTP capacity (MGD)	NPW*	Cost*	Cost per kW*
1.0	−575.5	703.6	24.0
2.1	1344	1136	18.4
3.2	3540	1398	15.5
4.2	5540	1654	13.5
5.3	7580	1885	12.4

3.3. Technical feasibility

The analysis has determined that electrical generation based on the air-blown gasification of sludge is energy positive. In order to support the thermodynamic conclusions, a technical evaluation was conducted to determine if such a process could be feasible to construct and operate. Wastewater sludge properties are variable, with organic and inorganic compositions that depend on numerous factors, including WWTP process design, geography, time of day and year, upstream plant conditions, etc. An additional complication comes from limited commercial experience with TCC of sludge, especially at small scale. Where sludge-specific components were not available, the analysis used data from general biomass sources with the assumption that this technology is likely adaptable for use with sludge.

Feedstock drying and conveyance may pose the greatest mechanical challenges to system design. Sludge enters the system boundary at 80 wt% water content as pumpable slurry. The air-blown and steam gasification processes described above require fuel moisture of 10 and 28 wt% respectively. Mechanical dewatering approaches such as centrifuges and belt presses are generally incapable of reducing moisture below 75 wt% (Werther and Ogada, 1999). Some filter presses are available which can reduce moisture below 75 wt% but they are nonetheless unable to achieve the moisture levels needed for the system specifications. Thus, even where mechanical dewatering is incorporated as a first step, full drying must be accomplished by thermal means.

Two general classes of thermal dryers are available, which can readily dry sludge to less than 10% water content. Direct, or convective, dryers contact sludge with hot drying gas. Direct dryers can accept drying gas streams of relatively low temperatures, even below 200 °C (Kemp, 2005), allowing for large mass flows of drying gas. This is useful where low temperature waste heat sources are available. Indirect dryers operate in the 300–400 °C or greater temperature range by circulating a heat transfer fluid through a jacketed chamber. Conduction from the chamber wall raises the temperature of sludge causing water to evaporate. Regardless of the drying technology used, a saturated, malodorous exhaust gas stream exits the dryer and dryer-gas scrubbing will likely be necessary. Water scrubbers offer an effective, simple solution to condense water and organics from the dryer exhaust stream before stack discharge. It is anticipated that the water demanded by the dryer exhaust scrubber would not pose a problem to a WWTP.

Sludge conveyance after the dryer is influenced by final moisture. Sludges with water content less than 15 wt% can be considered granular solids and are easily handled (Arlabosse et al., 2005). Discussions with sludge conveyance equipment suppliers including Komline–Sanderson and RUF suggest that sludges above 15 wt% water content may not be conveyable. Thus, at a fuel moisture of 10 wt%, sludge conveyance for the air-blown process is found to be feasible. However, sludge conveyance for the steam process fuel moisture of 28 wt% may not be possible. Redesigning the steam process to dry sludge to 15 wt% moisture and supply the necessary steam from a utility steam generator reduces net electrical output further below the air-blown process.

The final components involved in sludge handling are the briquetter (for fixed bed gasifiers) and gasifier charging (feeding) equipment. The dry, granular solids required by the air-blown process are readily briquetted by available equipment. Depending on sludge characteristics, a binder may be required for briquetting. Gasifier charging equipment design depends on the type of reactor. Downdraft gasifiers may be gravity fed by metering hoppers or conveyors. Charging downdraft gasifiers should be simple and robust for briquettes of dry fuel (Cummer and Brown, 2002). Fluidized bed gasifiers operate on small particulate fuel; usually no briquetting is necessary. However, fluidized beds require more sophisticated charging apparatus (Cummer and Brown, 2002). The technical challenges and mechanical complexity of charging fuel to fluidized bed reactors suggests that these systems will be more expensive and require more operator attention than mechanically simpler fixed bed reactors.

Numerous gasifier designs have been developed for biomass processing (Bridgwater, 2003). This analysis considers the fixed-bed downdraft gasifier to be the most economical option. The fixed-bed downdraft gasifier supports a continuously replenished fuel pile on top of an ash grate. Fresh fuel is charged from the top of the reactor and proceeds through drying, pyrolysis, and combustion zones as it travels downward toward the ash grate. Combustion air is drawn through the bed and usually also injected via controlled tuyéres (air injectors) in the combustion zone. Syngas contact with charcoal on the ash grate acts to filter many contaminants, producing a low tar syngas with limited heavy metal entrainment. Additionally, hot ash serves to catalyze tar cracking reactions (Fonts et al., 2012), resulting in a relatively low tar syngas. Syngas is drawn from the bottom of the reactor, which also holds ash withdrawal machinery. Fixed-bed reactors are however limited in scale by the potential to develop gas channels and/or hot spots.

Biomass gasification in downdraft reactors is established in practice (Bridgwater, 2003). It must be stressed, however, that wastewater sludge inorganic content is quite different from common biomass. Further research is needed to determine the effects of sewage sludge ash at high temperature on reactor materials. Ash fusion may also present operational challenges, although it is unlikely to pose a problem for operating temperatures around 850 °C because sludge ash typically does not soften until 1100 °C (Kupka et al., 2008).

Processes downstream of the gasifier are not expected to be unusually challenged by sludge fuel in comparison to common biomass fuels. However, such equipment may necessarily be specialized and costly. For example, the syngas heat exchanger and associated piping must be constructed of specialized stainless steels to accommodate high temperature hydrogen and the possibility of water condensation during startup and shutdown. Also, special consideration must be given to the internal combustion engine to ensure its compatibility with low LHV syngas. Nevertheless, equipment for all processes downstream of the gasifier was found to be commercially available from process equipment suppliers.

Final disposal of ash and char is not expected to be challenging. Landfilling tariffs and transportation costs for gasification ash and char are likely to be much less than those for dried sludge due to the reduced mass requiring disposal, and absence of organic content which reduces odor and vector concerns at the landfill. Additional economic benefit may be gained by selling (or donating) ash and char as an aggregate amendment to the concrete industry. However, detailed analysis of solid material disposal was outside the scope of this research.

3.4. Economic feasibility

This analysis shows that it is technically and energetically possible to produce net electrical power from the air-blown gasification of wastewater sludge. Whether this technology can be successfully applied in decentralized wastewater treatment plants depends on economic considerations. In order for a gasification plant to be economically feasible, the savings from generated electricity and reduced disposal cost must offset the capital investment and manufacturing costs incurred by the plant.

Present worth of the gasification plant (Eq. (5)), in comparison to the base case, becomes economically feasible at a plant capacity of about 0.093 m^3/s (2.1 MGD). Over a plant lifetime of 20 years, the 2.1 MGD plant will earn about $1,344,000 over the base case. The plant cost (Eq. (6)) is estimated to be $1,136,000, or on a unit basis, $18,400/kW$_e$. An increase in plant capacity to 0.134 m^3/s (3.2 MGD) is expected to cost $1,398,000 and earn $3,540,000 over the base case with unit cost of $15,500/kW$_e$. Profit margins at this level allow for some confidence in covering unanticipated costs. Recall that the cost estimate used is only accurate to ±30%. Increasing the cost estimate by 30% results in a reduction to $1,052,000 in net profit for the 2.1 MGD system. Nonetheless, the 2.1 MGD system remains economically feasible, suggesting that electrical revenues and saved natural gas cost drive economic feasibility more than plant cost. Economic performance indicators for plants from 1.0 to 5.3 MGD are summarized in Table 4.

To benchmark the cost analysis, the TCI of the modeled gasifier plant was compared to a highly automated, turnkey biomass gasification platform produced by Community Power Corporation (CPC) (Community Power Corporation, 2012). The CPC system delivers 100 kW$_e$ from wood fuel at a cost of $1.2 million commissioned. Small-scale gasifier plants are also available from other companies, including Gasek (Reisjärvi, Finland) and All Power Labs (Berkely, CA); however, the sophisticated CPC system better resembles the system design in this work. The modeled sludge gasification system at a capacity equivalent to the CPC system is expected to cost $1.1 million. The close agreement between modeled costs and the commercial system supports the economic modeling approach used in this study. Further engineering effort will likely increase the range of economic feasibility for decentralized WWTPs. A concern in this analysis is the effect of historically low natural gas prices in the United States at the time of this study. Further decrease in natural gas cost could render gasification at this scale economically infeasible. However, economic feasibility may be maintained by co-fueling the engine with syngas and natural gas. This approach would also serve to increase engine efficiency and combustion stability (Szwaja et al., 2013). As in any waste-to-energy scheme, good predictive models of local utility costing should be included in a detailed economic analysis.

4. Conclusion

The results from this study suggest that decentralized, urban WWTPs with plant flows of about 0.093 m^3/s (2.1 MGD) and greater can successfully recover value and energy from sludge using air-blown gasification. Air-blown gasification was found to convert sludge to electricity with an efficiency greater 17%. Over a 20-year plant lifetime, up to $3.5 million may be earned in comparison to a base case of thermal drying and landfill disposal. Application of this technology promises to reduce operating costs of wastewater treatment plants, carbon emissions from fossil-fired electricity, and the quantity of sludge requiring land disposal.

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