Comparative Assessment of Hydrothermal Gasification and Anaerobic Digestion using Aspen Plus and SuperPro

David Ballenghien¹, Rajesh S Kempegowda², Wei-Hsin Chen³, and Khanh-Quang Tran⁴*

¹374 water Inc, USA
²Inrigo As, Trondheim Norway.
³Department of Aeronautics and Astronautics, National Cheng Kung University, Tainan 701, Taiwan
⁴Department of Energy and Process Engineering, Norwegian University of Science and Technology, NO7491 Trondheim, Norway.

Abstract. Wet biomass is gaining increasing attention as an energy source globally. Various wet biomass materials like macroalgae, microalgae, sewage sludge, cattle manures, and food waste, are of high moisture contents, typically around 70% or more. There are two routes of high potential for converting these wet biomasses resoures to gaeous fuel, which are biomethanation and supercritical water gasification (SCWG). Biomethanation is conversion of wet-organic residues to biogas using via anaerobic digestion process. SCWG is a thermochemical conversion process, taking place in supercritical water producing enriched CH₄ and H₂ gases. In this study, the two processes are comparatively investigated using different process simulation softwares. Aspen Plus was employed for SCWG, while the biomethanation process was simulated with SuperPro. Various raw materials were used as input. We calculated the energy efficiency of the biomethanation process using experimental yields from the literature. For the SCWG process, we determined the higher heating value based on the simulated composition of the methane-enriched gas. Additionally, we conducted an economic analysis to compare the two processes, taking into consideration specific criteria relevant to the Norwegian context.

1 Introduction

There is serious speedup required in renewable energy particularly in biomass-based fuels. According to IEA data [1] “global biofuel demand is projected to increase by 28% (41 billion liters) from 2021 to 2026”. This is mainly driven by post-Covid-19 recovery and government policies. As well, factors like transport fuel demand, costs, and policy design will influence growth patterns. Asia is expected to surpass Europe in biofuel production. There is a tendency of another oil-embargo expected in coming years. These uncertainties tend government responses to feedstock prices, which may impact the demand. There are serious discussions

* Corresponding author: khanh-quang.tran@ntnu.no

© The Authors, published by EDP Sciences. This is an open access article distributed under the terms of the Creative Commons Attribution License 4.0 (https://creativecommons.org/licenses/by/4.0/).
in major economies like the United States, Europe, India, and China could potentially push for doubling the production of biofuels.

However, achieving this potential presents significant challenges related to sustainability and cost. In the context of Norway, characterized by its cold climate, experiences slower biomass growth compared to tropical regions. In contrast, macroalgae, which thrives in Norwegian waters, holds great promise as a biofuel source. By dedicating Norwegian sea area of 1% [2], report estimated that 60% of the country's fuel needs could be met by biofuel produced. Another report estimated that an area equivalent to Norway's salmon production zones (800 km²) could sustain the production of 470,000 tonnes of ethanol which can cut down a significant portion of the country's gasoline and diesel consumption [3]. Considerable research efforts are already underway, in Norway and internationally, to cultivate macroalgae for biofuels. Notable national initiatives include Macrobiomass, the Norwegian Center for Seaweed Technology, and projects by Seaweed Energy Solution and Statoil.

Lignocellulosic residues from existing processes, such as fermentation residues from lignocellulosic ethanol and biogas plants, represent another potential biomass feedstock to meet future biofuel needs in Norway. This alternative biofuels production pathway is instrumental in enabling Norway to develop cost-competitive biofuel value chains and achieve ambitious targets set by national and international policies and strategies. This article aims to compare two processes for converting wet biomass to gaseous biofuels: biomethanation and supercritical water gasification (SCWG). The goal is to establish reliable figures describing energy efficiency for different types of feedstock based on process simulation and experimental data from the literature.

Wet biomass, which does not typically compete with food production, includes feedstocks such as macro- and microalgae, wastewater-treatment-plant sludge, cattle manure, and high-moisture content waste from the food industry. These feedstock, often waste streams or by-products, can lead to significant pollution if mishandled due to their high organic contents. Current treatment methods involve biomethanation or aerobic digestion processes, but these processes have limitations, including the inability to digest lignin-rich materials and the challenges of storing the resulting digestate sludges, especially in densely populated areas.

SCWG is pushed as a solution to these issues. It achieves nearly complete conversion on a pilot plant scale, leaving virtually no organic matter after the reactor. SCWG offers several advantages over biomethanation process, including higher throughputs, energy efficiency, the ability to handle mixed feedstocks, and no need for specialized microbial cultures or enzymes.

In contrast to biomethanation, SCWG requires harsh conditions, including high pressure and temperature (above 22.1 MPa and 374°C). Heat losses at such temperatures are higher, and substantial energy is needed to reach the required pressure. While SCWG claims higher overall energy efficiency than biomethanation, when considering the investment, operation, and maintenance costs, the preference between the two processes is not straightforward. To help in determining the best process design, a detailed Pinch analysis is performed to optimize heat integration and implement heat exchangers in the SCWG process. This analysis evaluates the overall energy efficiency of methane-rich gas production. Additionally, the SCWG process is compared with conventional anaerobic processes in terms of technical and economic viability under Norwegian conditions.
2 Process description of two process under compare

2.1 Hydrothermal gasification

Fig. 1 depicts the SCWG process, accommodating a biomass input rate of 20 tons per hour with an initial moisture content of 80%. The process initiates with the ambient-temperature biomass and undergoes a series of pre-treatment steps, including chipping, milling, weak acid hydrolysis to reduce alkalinity, and coarse filtration based on the characteristics of the feedstock. Subsequently, the high-pressure pump elevates the pressure to 34.5 MPa, while the temperature is increased to approximately 377°C. This temperature increase is achieved through a two-step process: first, via heat exchange, and second, by an additional heat supply from a combustor fuelled by a fraction of the product gas. The final temperature is selected to align with the desired gasifier outlet temperature of 600°C.

After leaving the gasifier, the ash and supercritical mixture are separated. The ash is removed from the process, while the supercritical stream passes through multiple heat exchangers to preheat the incoming stream. It then undergoes the first flashing tank, maintaining a pressure of 34.5 MPa. The gas stream is enriched with hydrogen, with a hydrogen content as high as 95%, whereas the liquid stream is extracted and routed through a second flashing tank, where the pressure is finally reduced to 1.0 atm. This step aims to recover gases dissolved in the water. A portion of this stream is directed to a combustor, which supplies the additional heat needed in the preheating section, especially when heat exchange alone cannot suffice. The remainder of the stream is available for further use, similar to the hydrogen-enriched gas stream. The liquid stream following the second flashing tank is removed from the process. It predominantly consists of water, with some inorganic and unconverted organic compounds.

![Diagram of the process](image)

Fig. 1. Process configuration of wet-biomass processing using hydrothermal gasification under super-critical conditions for CHP

2.2 Anaerobic digestion

Biomethanation is an anaerobic process involving four steps: hydrolysis, acidogenesis, acetogenesis, and methanation. Biomethanation is applied in various wastewater, slurry, and solid waste treatment processes, utilizing different reactor types and conditions to maximize energy output and process efficiency. In this example, data gathered from Inrigo Biosystem data for three-stage anaerobic reactor after slurry pre-treatment are proposed to enhance the biomethane yield. Stage thermal hydrolysis after slurry prepared by milling, pumped to thermal hydrolysis which take place at 70°C to meet the PAS rules as per pasteurization concern. Then pasteurized slurry supplied to anaerobic digester and temperature of the slurry
cooled down by mixing with recycled sludge. The main criteria to avoid inhibition is by controlling total organic nitrogen below 5%.

![Block flow diagram of biomethanation process](image)

**Fig. 2.** Block flow diagram of biomethanation process.

Fig. 2 depicts the outline of the biomethanation process. The pretreatment involves chopping and milling waste feedstocks. These pretreated materials are then heated to 70°C during thermal hydrolysis and subsequently transferred to a mesophilic anaerobic reactor maintained at 35°C. The hydraulic retention time for anaerobic digestion considered here ranges from 15 to 20 days, resulting in the production of a CH₄ and CO₂ mixture. The typical conversion of chemical oxygen demand (COD) in the reactor is approximately 60%. Unconverted effluents from the first-stage anaerobic digester are directed to the mechanical dewatering system, known as the "mechanical dewatering system," where the cake can yield around 30% dry mass. Further drying is carried out using a flue gas dryer. The dried biosolid sludge is incinerated in the boiler/combustor to provide heat for the preheating section. The effluent from the reactor is then sent to an anaerobic biofilm reactor to reduce unconverted COD and enhance biomethane production. Subsequent effluent can undergo aerobic treatment in a packed bed/trickling bed aerobic reactor to reduce COD and meet water recycling objectives. The biomass input is adjusted to 20 tons per hour with a moisture content of 10%, ensuring that an equivalent amount of dry matter is utilized in the SCWG process, enabling a fair and meaningful comparison.

### 3 Methodology of process system design

<table>
<thead>
<tr>
<th>Proximate</th>
<th>Kelp</th>
<th>Birch</th>
<th>Water Hyacinth</th>
</tr>
</thead>
<tbody>
<tr>
<td>Moisture</td>
<td>0.00</td>
<td>0.00</td>
<td>0.0</td>
</tr>
<tr>
<td>Ash</td>
<td>45.63</td>
<td>0.22</td>
<td>22.1</td>
</tr>
<tr>
<td>Volatile</td>
<td>49.36</td>
<td>89.43</td>
<td>62.0</td>
</tr>
<tr>
<td>Fixed Carbon</td>
<td>5.01</td>
<td>10.35</td>
<td>15.90</td>
</tr>
</tbody>
</table>

**Ultimate**

<table>
<thead>
<tr>
<th>Ultimate</th>
<th>Kelp</th>
<th>Birch</th>
<th>Water Hyacinth</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon</td>
<td>24.19</td>
<td>48.62</td>
<td>36.4</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>3.08</td>
<td>6.34</td>
<td>4.77</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>2.62</td>
<td>0.09</td>
<td>1.82</td>
</tr>
<tr>
<td>Sulphur</td>
<td>0.70</td>
<td>0.05</td>
<td>0.4</td>
</tr>
<tr>
<td>Ash</td>
<td>30.00</td>
<td>0.022</td>
<td>22.1</td>
</tr>
<tr>
<td>Oxygen</td>
<td>23.78</td>
<td>44.90</td>
<td>32.6</td>
</tr>
<tr>
<td>Chlorine</td>
<td>15.63</td>
<td>0.00</td>
<td>1.9</td>
</tr>
<tr>
<td>HHV (MJ/kg)</td>
<td>9.33</td>
<td>19.80</td>
<td>14.81</td>
</tr>
</tbody>
</table>
Fig. 3 illustrates the comprehensive process simulation of SCWG using Aspen Plus. However, for the biomethanation process, the requirement for well-established and validated data on anaerobic digestion necessitates the utilization of Superpro Designer [5], a bioprocess simulation software. The methane yield and, consequently, the efficiency of the biomethanation process are determined for various feedstocks based on experimental data sourced from the literature. Additionally, a heat integration strategy were based on the Aspen Plus simulation screened temperatures. Then, the techno-economic viability of the two process were compared based on an annuity analysis. The attributes of the feedstocks examined in the context of Norwegian conditions are presented in Table 1.

Based on the process flow diagram shown in Fig. 3, mass and energy balances for each process and unit were performed using the Aspen Plus. This package comprises numerous sub-models that simulate individual units and components. Calculations were conducted for each of the sub-models to determine thermodynamic properties and describe the behaviour of every single component and mixture. Equations and data required for the calculations were obtained from Aspen Plus databases. Input data were gathered from various sources, including experimental and literature data. Due to the high pressure and temperature of the process, the behavior of compounds deviates significantly from ideality. Typically, for petrochemical reactions, Peng-Robinson or RK-Soave equations are employed [4–6]. In our case, we adopted the Peng-Robinson equation, which is widely accepted for solids handling.

3.1 Hydrothermal gasifier modelling in Aspen Plus

The synthesis of the base configuration flowsheet is developed using Aspen Plus, as illustrated in Fig. 3. In Aspen Plus, the defining of properties for biomass is done according to Aspen Plus solid modelling methodology. Gasifier modelled by combination of decomposition reactor R-Yield where biomass decomposed into its basic proximate and ultimate fractional yields. Carbon is transformed into solid carbon, oxygen into O\textsubscript{2}, nitrogen into N\textsubscript{2}, hydrogen into H\textsubscript{2}, sulfur into solid sulfur (S), and chlorine into Cl\textsubscript{2}. Water remains as water, and ash remains as ash, with ash being a pre-tabulated compound in Aspen Plus. Gasification reactor modelled based on non-stoichiometric model by the well-known Gibbs approach. It is worth noting that ash is assumed not to undergo any chemical interactions within the process, which is a significant simplification. It is also important to mention that the catalytic effects of mineral compounds, such as KOH, Na\textsubscript{2}CO\textsubscript{3}, and K\textsubscript{2}CO\textsubscript{3}, have been reported [7, 8] and leaching and demineralization effect of inorganics ignored in the mass and energy balance. These compounds can readily form from ash in supercritical water, especially since the ion dissociation constant (\(K_w\)) of water decreases significantly after the critical point, leading to the precipitation of most ionic compounds [9]. However, a comprehensive modelling effort would be necessary to fully understand how all components
of ash interact with other compounds, but this level of detail falls outside the scope of this work. The gasifier is represented in the model by a Gibbs reactor.

### 3.2 Gas upgrading and separation process

K-values predicted with the thermodynamics property Peng-Robinson were used to calculate the mole fraction of each chemical in the vapour and liquid phase shown below equation 3.1.

\[
y_i = K_i x_i
\]

(3.1)

Where \( y_i \) is the molar fraction of compound \( i \) in the gas phase, and \( x_i \) the molar fraction of the compound in the liquid phase.

Boiler: The boiler is simulated as a Gibbs reactor, aiming to minimize the Gibbs free energy of the mixture in the presence of air (with oxygen at a 1:2 molar ratio). The reactions for boiler combustion are shown below:

\[
H_2 + \frac{1}{2} O_2 \rightarrow H_2O
\]

(3.2)

\[
CO + \frac{1}{2} O_2 \rightarrow CO_2
\]

(3.3)

\[
CH_4 + 2O_2 \rightarrow 2H_2O + CO_2
\]

(3.4)

It receives a fraction of the generated gas to meet the preheating requirements. The hot exhaust gas aids in the final preheating step before the gasifier through heat exchange with the incoming stream. The usable temperature range is limited to around 400°C, given that initial preheating steps rely on heat recovery from the outgoing gas.

### 3.3 Simulation of the biomethanation process

This section provides details on the simulation of the biomethanation process using the commercial software Superpro Designer. The simulated sub-processes include anaerobic digestion, aerobic reactor, mechanical dewatering, and sludge combustion for steam generation, as illustrated Fig. 4. The mass yields of each compound depend on the characteristics of the biomass feedstock type. Experimental yields validated under the anaerobic reactor model of Superpro for is valid only for the model compound of Water Hyacinth C_{6}H_{12}O_{6}.8 as methane 14.8%, CO2 40.8%, sludge 34.6%, and lignin 10%. Volatile Solid (VS) is the fraction of total solid that has been volatilized after combustion at 550°C.

Aerobic Reactor: The reactions in the aerobic reactor, designed to treat water by oxidizing the remaining organic compounds, primarily produce CO_2. Some ammonium radicals may also be released based on substrate N_2 concentration, which is subsequently oxidized into nitrates. The denitrification reaction has not been included here, as our focus is on the energy balance of the process. Methane produced from the anaerobic reactor and the organic fraction of the sludge are considered energy products. To prevent the excessive accumulation of sludge, it is further sent to the belt filtration system.

Belt Filtration: The sludge water content after belt filtration is set at 70% [10]. Mechanical water removal can only concentrate the sludge to around 30% solid content. However, for it to be usable as a fuel, the sludge needs to be dried to a moisture content of around 25%. Therefore, for every 1 kg of sludge, approximately 2.0 kg of water must be thermally removed, consuming 4.9 MJ of energy. The dried sludge is then fed into the boiler process for heat production.
Results and discussions

4.1 Sensitivity analysis

Sensitive analysis were carried to know the influence of gasification temperature on the composition of the product gas, of which the results are presented in the Fig. 5. This sensitivity analysis shows that H$_2$ is favoured at high temperature, whereas CH$_4$ gets under 10% molar fraction beyond 1000 K or 727°C. The flashing pressure sensitivity analysis shows that the composition of the flows can be controlled through a pressure control in the flashing tanks. Methane molar fraction is the highest around 10 bar, whereas H$_2$ keeps increasing with the pressure. For different types of feedstock, the yields were obtained from experimental data found in the literature[11,12] and supported by in-house industrial data from Inrigo.
4.2 Heat recovery process and design modification

The heat integration is essential in the SCWG process and impacts dramatically the energy efficiency. The methodology is detailed in [8]. The high temperature and pressure of the process and the large quantity of water in biomass makes it certain challenges with selection of heat exchanger due to corrosion after the gasifier. In our case where the purpose is to use wet biomass, recycling of water is not interesting since the biomass is contained it and there is already an excess of water. However, if SCWG is to be applied to dry feedstock such as straw or wood, a stream of high pressure and high temperature water is mixed with the stream of biomass. Most of the previous studies in that field based their efficiency analysis on assumptions of the heat exchange efficiency, or energy recovery efficiency. Pinch analysis was carried out to propose a relevant heat integration and implementation of heat exchangers. Due to uncommon conditions of temperature and pressure, an average value of the thermal capacity $C_p$ of the different fluids would be irrelevant, therefore the specific enthalpy of each stream has been calculated numerically by steps of 0.5 K with Aspen Plus, and average values of $C_p$ have been calculated on the same ranges of temperature. Fig. 6 shows the heat integration strategies for hydrothermal gasifier.

![Fig. 6](image)

Fig. 6. (a) Heat integration strategy and (b) implemented in Aspen plus
4.3 Composite diagrams for different types of feedstock

Fig. 7 shows the composite diagrams of the flows to be cooled (upper curve) and those to be heated. The impact of the minimum temperature approach on the quantity of heat that has to be supplied by the external source with minimum temperature approach of 15 K and 30 K are illustrated. It is clear from Fig. 7 that the available energy from the hot streams is larger than the energy requirement to heat the cold streams. For instance for Kelp, in the formerly given conditions, the total enthalpy rate available at the hot streams is 9847 kJ/s, which needs to be given away, whereas the enthalpy rate required at the cold streams is 9622 kJ/s. However, for a minimum temperature difference of 30 K in the heat exchangers, part of the heat must be supplied by an external source due to insufficient high temperature heat. This amount of heat depends of course on the temperature difference, which will influence both the heating rate and the size of the heat exchangers, and therefore the investment costs. The same pattern occurs with Birch and Kelp as shown in figure. From an energy efficiency point of view, efficient heat exchangers should be preferred however, the larger investment cost may hinder the solution.

![Composite diagrams for WATER HYACINTH, BIRCH and KELP at 15K and 30K](image)

**Fig. 7.** Composite diagrams for WATER HYACINTH, BIRCH and KELP at 15K and 30K

4.4 Energy efficiency of SCWG

The results of the simulation are given in **Table 2**, consisting of syngas composition and energy efficiency for different types of feedstocks with two different heat integration strategies.
Table 2 Energy efficiency and composition of the produced gas with SCWG for different fuels.

<table>
<thead>
<tr>
<th>Feedstock Component</th>
<th>KELP Mol Fraction</th>
<th>Birch Mol Fraction</th>
<th>Water Hyacinth Mol Fraction</th>
<th>Literature (I)</th>
</tr>
</thead>
<tbody>
<tr>
<td>H₂</td>
<td>0.31</td>
<td>0.24</td>
<td>0.28</td>
<td>0.29</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.18</td>
<td>0.33</td>
<td>0.29</td>
<td>0.25</td>
</tr>
<tr>
<td>CO</td>
<td>0.01</td>
<td>0.01</td>
<td>0.01</td>
<td>0.01</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.35</td>
<td>0.41</td>
<td>0.39</td>
<td>0.43</td>
</tr>
<tr>
<td>N₂</td>
<td>0.02</td>
<td>0.00</td>
<td>0.01</td>
<td>0.03</td>
</tr>
<tr>
<td>SO₂</td>
<td>0.01</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
</tr>
<tr>
<td>HCl</td>
<td>0.12</td>
<td>0.00</td>
<td>0.01</td>
<td>0.00</td>
</tr>
</tbody>
</table>

Temperature approach \( \Delta T_{\text{min}} [K] = 30 \text{ K} \)
- Energy efficiency \( \eta \) 55% \(^a\) 77% 70% 64.8% \(^b\)
- Fraction Methane-enriched gas sent to boiler 0.54 0.24 0.34 -
- H₂ rich flow [kg/h] 41.5 64.5 55.9 -
- Average HHV [MJ/kg] 90 83 85.8 -
- Output Methane-enriched gas sent to boiler [kg/h] 894 1855 1248 -
- Average HHV [MJ/kg] 7.9 13.9 13.1 -

Temperature approach \( \Delta T_{\text{min}} [K] = 15 \text{ K} \)
- Energy efficiency \( \eta \) 69% \(^c\) 84% 80% -
- Fraction Methane-enriched gas sent to boiler 0.36 0.15 0.22 -
- Boiler after HX temperature [K] 680 680 680 -

4.5 Anaerobic digestion process

Yields are taken from experimental data found in the literature [9] for different types of feedstock. They are usually expressed as Nm³ methane per kg of volatile solid (VS). The methane yields and the simulated energy efficiency are tabulated shown below as a comparison to hydrothermal gasifier-based methane rich gas production as shown in Table 4.

Table 3 Biomethanation energy efficiency for different feedstocks (Anaerobic digestion)

<table>
<thead>
<tr>
<th>Feedstock</th>
<th>Water Hyacinth, (theoretical)</th>
<th>Water Hyacinth Pretreatment</th>
<th>Laminaria Hyperborea</th>
<th>Kelp</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water [kg]</td>
<td>9000</td>
<td>9000</td>
<td>9000</td>
<td>9000</td>
</tr>
<tr>
<td>Biomass [kg]</td>
<td>1000</td>
<td>1000</td>
<td>1000</td>
<td>1000</td>
</tr>
<tr>
<td>VS(^a) added [kg]</td>
<td>1000</td>
<td>779</td>
<td>779</td>
<td>820</td>
</tr>
<tr>
<td>Ash content [wt %]</td>
<td>0</td>
<td>22.1</td>
<td>22.1</td>
<td>18</td>
</tr>
<tr>
<td>VS reduction</td>
<td>0.554</td>
<td>0.498</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>CH₄ produced [kg]</td>
<td>148</td>
<td>92</td>
<td>155</td>
<td>131</td>
</tr>
<tr>
<td>HHV Biogas [MJ]</td>
<td>8234</td>
<td>5145</td>
<td>8639</td>
<td>7300</td>
</tr>
<tr>
<td>Preheating duty [MJ]</td>
<td>680</td>
<td>680</td>
<td>680</td>
<td>680</td>
</tr>
<tr>
<td>Evaporation of water in sludge</td>
<td>3499</td>
<td>2549</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Heat loss from the reactor [MJ]</td>
<td>20</td>
<td>13</td>
<td>13</td>
<td>17</td>
</tr>
<tr>
<td>Electric demand [MJ]</td>
<td>520</td>
<td>520</td>
<td>520</td>
<td>520</td>
</tr>
<tr>
<td>Dry sludge produced [kg]</td>
<td>446</td>
<td>612</td>
<td>-</td>
<td>-</td>
</tr>
</tbody>
</table>
4.6 Techno-Economic analysis

Cost estimations of equipment are based on the principles engineering design involving various process design principles and cost of equipment modified from various literatures [13,14]. A scaling factor of 0.6 was applied to SCWG plant costs, and the same scaling factor was used to assess biomethanation plant costs (CAPEX). Direct and indirect costs were drawn from Peter et.al [15], Gasif et.al [14], and our previous publications [16–18]. Costs are expressed in US dollars, and both capital expenditures (CAPEX) and operating expenditures (OPEX) were provided for both SCWG and Biomethanation plants. Cost modelling approach were well documented by us in various sponsored research projects and publications [16–20]. Estimated annualized cost for SCWG and biomethanation process are shown in Table 4.

<table>
<thead>
<tr>
<th>Table 4 Summary of the annuitized cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>SCWG process</td>
</tr>
<tr>
<td>Plant in Million $</td>
</tr>
<tr>
<td>Annual cost of capital</td>
</tr>
<tr>
<td>Annual operation and maintenance cost</td>
</tr>
<tr>
<td>TOTAL</td>
</tr>
</tbody>
</table>

The comparison involves plotting, as shown in Fig. 8, the cost of producing product gases ($/GJ) against the biomass purchase cost. We conducted this comparison without considering combined heat and power (CHP) for a fair assessment. Different biomass purchase prices were considered, ranging from $20/ton (the current price for macroalgae in the Norwegian alginate industry) to as low as negative prices, like -$60/ton, often applicable to materials such as sewage sludge, where disposal costs can be significant. The results show that SCWG generally incurs higher production costs due to its greater investment expenses. However, the choice of feedstock significantly influences production costs. Notably, SCWG competes effectively with biomethanation when a scaling factor of 0.6 is applied for water hyacinth feedstock across a wide range of purchase prices. Conversely, for Kelp, SCWG consistently yields higher production costs compared to biomethanation, regardless of the scaling factor.
In this study, we have examined both biomethanation and supercritical water gasification processes with a focus on energy efficiency, heat integration and cost-economics. We employed two different minimal temperature approaches for the SCWG process, namely 30°C and 15°C, which reflect the performance of heat exchangers and have a significant impact on overall energy efficiency. At a 30°C minimal temperature approach, the energy efficiency for Kelp in the SCWG plant was 55%, which increased to 69% at a 15°C minimum temperature approach, while biomethanation achieved 51% energy efficiency. This difference becomes even more pronounced with feedstocks containing lignin, where biomethanation struggles without pretreatment. For water hyacinth, the SCWG plant achieved 70% energy efficiency at a 30°C minimal temperature approach, while biomethanation with steam pretreatment reached 55% and without pretreatment, only 31%. The key advantage of SCWG is its exceptionally high energy efficiency, particularly for feedstocks containing lignin when compared to biomethanation. Additionally, SCWG offers near-total biomass conversion and faster operations within a smaller footprint.

Acknowledgements

This publication has received funding from the Norway Grants 2014–2021 via the National Center for Research and Development. The article has been prepared within the frame of the project: “Negative CO₂ emission gas power plant” - NOR/POLNORCCS/NEGATIVE–CO₂–PP/0009/2019–00 which is co-financed by programme “Applied research” under the Norwegian Financial Mechanisms 2014–2021 POLNOR CCS 2019 - Development of CO₂ capture solutions integrated in power and industry processes.

References


