Performance evaluation of a small scale digester for achieving decentralised

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*Highlights

Decentralised digestion operation was evaluated considering organic removal and energy efficiency

Small scale digestion units demonstrated technical suitability for biogas valorisation

Electrical self-efficiency coefficient was 0.95 for a yield of 360 L CH_4/kg VS and 93% of VS removal

The auxiliary storage system offered high flexibility to the small-scale plant increasing process stability.

Reactor thermal needs highly compromised the efficiency when operating at low organic loads

Essential title page information

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Title: Performance evaluation of a small-scale digester for achieving decentralised management of waste

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Abstract

The performance of a small-scale prototype digestion plant (7.2 m^3 working volume) intended for decentralised operation was evaluated considering energy efficiency and technical suitability for biogas valorisation in producing electrical and thermal energy. The digester operated in recirculation mode to enhance organic matter conversion and improve volatile solid degradation. An energy assessment of the process assumed the incorporation of a combined heat and power (CHP) unit. The coefficient of overall performance of the plant for electrical energy (COP_{el}) was 0.95 — this values was estimated at an electrical efficiency

of 22.5% and represents the ratio between energy production and consumption — for a methane yield of 360 L/kg VS and an organic loading rate (OLR) of 1.06 g VS/L d. This parameter was slightly lower than the unit thus indicating that the micro-plant was close to attaining self-sufficiency regarding electrical energy use. The temperature increase of the feed to process conditions supposed a significant amount of thermal energy which highly compromised the efficiency when operating at low organic load, thus accounting for more than 80% of the total energy demand of the installation. When the energy assessment of the process was performed at higher OLR of 2.7 g VS/ L d, the resulting COP_{el} value was1.68, demonstrating the feasibility of this configuration for decentralised digestion.

Keywords: micro-plant energy performance, food waste digestion, decentralised waste treatment, prototype evaluation, anaerobic digestion

Nomenclature

CHP: combined heat and power

COP_{th}: Coefficient of overall performance, thermal energy

COP_{el}: Coefficient of overall performance, electrical energy

EG_{elec}: Electrical energy generation

$EG_{thermal}$: Thermal energy generation

HRT: Hydraulic retention time

OLR: organic loading rate

PA: Partial alkalinity

Q_{Biogas}: Energy contained in biogas

$Q_{C.el}$: Energy needed for electrical purposes

$Q_{C.th}$: Energy needed for thermal purposes

Q_{Feed}: Heat needed for increasing the temperature of the feed to process conditions

Q_{Losses}: Heat loss through reactor walls and piping

Q_{Man}: Maintenance heat

Q_{Proc}: Process heat

Q_{Total}: Total energy demand of the plant

TA: Total alkalinity

VFA: Volatile fatty acid

1. Introduction

Decentralised anaerobic digestion is a promising alternative for low population density areas or the treatment of a relatively small amount of wastes produced seasonally. This management alternative may also be appropriate when the distance to a large-scale centralised plant is less attractive on the grounds of sustainability due to transport impacts. Anaerobic digestion is an efficient technology for treating organic substrates capable of increasing the contribution of renewables to the overall energy production matrix (González et al., 2020). The lack of aeration provides a great advantage due to the decrease in the implicit cost associated with energy demand. However, depending on scale, anaerobic digestion may be favoured for centralised treatment; whereas, composting may be preferred in the case of small on-farm for treating animal manure (Lin et al., 2019)

Another parameter of particular significance is the transport of residues. In reality, there are limits based on reasonableness to the distance from where organic wastes are collected and subsequently treated at a centralised plant. Low population density areas impose challenges to the extrapolation of conventional technologies due to the smaller scale of the treatment units, the correspondingly higher operating costs and the practical and sustainability logistics of waste transportation. A study performed by Piñas et al. (2019) for a Brazilian scenario showed that biogas plants using mono-substrates such as cattle manure presented economic viability for electrical power higher than 740 kWe whereas a co-digestion system presented economic viability for electrical power above 1000 kWe. This scale is unsuitable for many applications in rural areas due to seasonal production and the significantly lower amount of available waste.

The development of small-scale digestion plants for treating waste in rural regions is crucial. Decentralised biogas production from manure and crop residues is not currently economical or reliable because gas production usually has a low energy potential. Codigestion systems are not economically viable because of the costs associated with silages and transport of biomass material (Piñas et al., 2019). These factors have delayed development in the implementation of this type of technology. Transportation of feedstock must not exceed 20 – 30 km (two-way), since increasing this distance negatively affects the economics (Rajendran and Murthy, 2019) thus justifying the need for decentralised units. However, the high initial investment of these plants act as the main disincentive requiring support from fiscal subsidies (Win et al., 2017). The design of small biogas reactors also raises fundamental issues concerning the energy demand associated with pretreatment and achieving good mixing in the reactor in order to optimise gas yields when feeding combinations of dry and wet food wastes, or domestic and agricultural wastes (Radu et al., 2016).

Small-scale digestion plants have recently attracted considerable interest to shift towards a more decentralised biowaste management strategy. This approach offers advantages compared to the conventional centralised waste treatment associated with reduced transport requirements and the potential benefit of increased community involvement. The efficiency of the decentralised approach relies on a close integration of the whole treatment supply chain, from the generation of biowastes to the valorisation of biogas and digestate (Thiriet et al., 2020).

Production activities and small communities must meet certain requirements if the decentralised approach for waste treatment is to become feasible. The first requirement is that these activities must generate organic waste streams of a high organic content that is readily biodegradable and with high methane potential. The second requirement is that they must be associated with local electrical and thermal energy demands. These characteristics define a niche of activities that mainly focuses on agri-food industries (agri-food farms, food products industries, livestock farms), sewage treatment plants and food waste managers. Decentralised anaerobic digestion facilities could produce operational benefits such as, greater ease in the management of wastes, the possibility of having energy autonomously, as well as economic benefits linked to thermal energy generation and production of organic fertilisers and amendments (Anyaoku and Baroutian, 2018). Other benefits are the ability to handle and treat wastes using the proximity principle basis, as close to point of origin as possible, which can drastically reduce emissions and impacts associated with transport. In addition, there would be another series of short-term improvements such as the increase in the generation of distributed energy, which in turn would contribute to the stability of the electrical system and reduce the costs of transporting energy and potential pollution from large centres of generation.

The reduction of costs associated with logistics when treating wastes would also be part of the short-term improvement as well as the optimisation of municipal waste treatment (Wang et al., 2014). It is for these previous reasons that the possibility of using "flexible anaerobic digestion micro-plants" are promising at this moment. Here, these plants are defined as "micro" for their small treatment capacity, being limited to amounts of < 1,000 t/year (<15 kWel) (NNFCC, 2017). The term "flexible" refers to a type of facility, despite its small size, that is provided with a process control system with similar benefits to those of large centralised anaerobic digestion plants, being also capable to adapt to any particular need regarding the type of organic material, nutrient content and seasonal production of waste. These characteristics allow this type of decentralised plant to present an operational improvement over traditional centralised installations. Decentralised treatment units are being considered a more sustainable solution because energy requirements are much lower and sophisticated operation is avoided thus being easy to adapt to different geographic contexts (Lourenço and Nunes, 2020).

This research evaluates the suitability of a micro-digestion plant for the decentralised treatment of food wastes in a real environment. The novelty of this prototype is based on its capacity for valorising wastes near the source, reducing transport needs and activating the local economy. Developing a new solid waste management strategy based on small decentralised units offers new opportunities for implementing this model into developing countries and communities with a disperse population. These small treatment plants can promote community participation and avoid an undesirable accumulation of organic materials prone to degrade uncontrollably. Decentralised management of wastes offers several advantages but there is a lack of studies reporting on the electric and thermal performance of small-scale prototypes.

In the present study, the first objective was to evaluate the digestion process using a micro-plant of flexible configuration fed with substrates comparable to those composing the organic fraction derived from catering services. The second objective was to evaluate the energy efficiency of the micro-plant establishing an energy balance to assess thermal and electric performance. This manuscript, thus reports data for evaluating performance of the digestion process considering not only biological yields but also energy demands associated with the operation of small scale units.

2. Materials and methods

2.1. Inoculum and substrates

Food wastes used as substrates were obtained from a hostelry school for the training of cuisine professionals. The school was dedicated to the teaching of Italian, French, Mexican and Spanish cooks. Undesirable materials like packaging, containers, bones, cutlery and other non-degradable components were manually screened out to obtain a food waste fraction easy to handle for grinding machines and free of plastics and any other kind of inert components that would exert a detrimental effect on the quality of the digestate. Food waste was daily transported using 50 L closed steel vessels from the school to the Algodor plant (located in Toledo, Spain) where the prototype was installed. This plant is specialised in the biological treatment of organic wastes from a diversity of sources, including fruit and vegetable wastes from the fourth-range industry, food waste from the hostelry sector and those from the maintenance of green areas. Currently these materials are transformed by static pile composting. The digestion prototype was installed in this treatment centre, with the aim of evaluating the suitability of energy production from wastes received daily.

The waste was subjected to an initial triage to remove contaminants and record daily quantity of waste received. Periodic sampling was carried out for characterisation of in-coming material for quality control and quality assurance purposes. The weighing of food wastes was performed using an industrial floor scale balance with a precision of 0.5 kg (Steinberg

Systems SBS-BW-1T). After weighing, the waste was fed into the pretreatment unit for grinding and then onwards to the reactor (Fig. 1a). The chemical and physical characterisation of food wastes is summarised in Table 1.

Table 1 here

Inoculum used to seed the micro-plant anaerobic digester was a digestate obtained from a co-digestion plant treating a mixture of municipal solid wastes and wastes derived from a meat processing factory. The industrial digester was located at the solid waste treatment centre of South Madrid, Pinto. The total solid content of the inoculum was 41.0 ± 1.9 g/L with a volatile solid content of 18.4 ± 0.7 g/L. Once loaded into the digester, the inoculum was heated at 37 ± 1 ºC for 20 days without any additional material being fed to allow the removal of the "background" biogas production and to achieve maximum degradation of available organic materials before commencing with pilot plant operations.

Figure 1 here

2.1. Micro-plant description and operation

The operational units comprising the pilot plant are presented in Fig. 1b showing the main equipment and auxiliary components. A detailed description of the plant and controller actions are given in the electronic supplementary material (ESM). The digestion unit consists of a tank reactor, the operation of which depends exclusively on the multipurpose pump. The digester has a total volume of 8776 L with a working volume of 7200 L. Mixing is performed by recirculation of the digestate using a multipurpose pump. The digester operated under mesophilic conditions at 37 ± 1 ºC. The average organic loading rate was 0.68 kg VS/m^3 d. The prototype was kept in operation for 106 days.

Actual biogas production was compared with theoretical predictions from a simplified version of Simons and Buswell equation (Møller et al., 2004):

$$B_u\left(\frac{\mathrm{L\ CH_4}}{\mathrm{kg\ VS}}\right) = \frac{\frac{n}{2}+\frac{a}{8}+\frac{b}{4}}{12n+a+16b} * 22.4$$

Methane production was estimated by assuming that all organic material was converted into biogas (methane and carbon dioxide) along with water. The use of carbon for microbial growth and maintenance needs were not considered. This expression establishes as main elements of organic matter: carbon, hydrogen and oxygen using the empirical formula $C_nH_aO_b$. The ultimate methane production (Bu) was calculated based on the stoichiometric Buswell equation and using the gas ideal factor for estimating the volume of a gaseous substance (1 mol) at STP conditions.

Due to instabilities intrinsic to the plant operation, both the frequency and the feeding rate were variable throughout the trial to adapt to substrate availability. The period selected to assess plant performance was from day 55 to 90 included (35-day continuous period) since this period showed process stability in terms of feeding rate and methane production.

The waste received daily was incorporated into the process through the pretreatment unit. For this, the waste material was poured into the feeding hopper for grinding and then into the pretreatment unit. In this tank, the organic material was mixed with digestate from the digestate storage tank to dilute the mixture, accelerate hydrolysis and further reduce the particle size thus facilitating their introduction into the main digester. A dilution ratio was established to attain a volumetric proportion of 200 L food waste/m^3. The feeding procedure involved manual registration of parameters that are not automatically recorded by the control unit, as it is: the amount of ground waste, the levels of the pretreatment unit and digestate storage tank along with readings of energy consumption of electric devices. During the digestion test, samples of the feed and digestate were regularly collected for characterisation at a frequency of once or twice each week. Difficulties associated with the operation of the prototype led to an irregular feeding of the reactor. The start-up of the plant was carried out at low organic loadings and was denoted phase I. Phase II corresponds to an increase in the organic loading which was based on the performance of digestion. Phase III was the last stage and corresponds to a period where feeding to the reactor was not available due to technical problems at the Algodor plant. The experimental period was established based on the time indicated by project activities to test and evaluate energetic performance of the prototype for obtaining a commercial and flexible unit capable of treating a great variety of wastes at small scale. For this reason, the prototype was not tested for a longer period using this type of feeding.

The prototype control unit has four operating states (Operation, Heating, Recirculation and Feeding) to carry out the digestion process. Based on the values of the process variables and the operating instructions, the control unit was responsible for activating the corresponding operating status, so that the process was carried out following the operating instructions. There was another operating state (Grinding) that is outside the control capacity of the control unit, since this was done manually, activities related to this later state took place simultaneously to any of the other operating states. The recirculation frequency of the mixing pump was 10 min every 6 hours, which for the installed device having a volumetric flow of 14.82 m^3/h, represents a turnover time of 17 h. The operational turnover time was affected by the heating needs, which are met by turning on the multipurpose pump and the heat exchanger, thus reducing this parameter. The heating system was capable of maintaining the temperature at the set value with a dead band of 1 ºC.

2.3. Chemical analyses and data recording

Total alkalinity (TA), partial alkalinity (PA), pH, total and volatile solids (TS and VS) and ammonia nitrogen (measured via selective electron) were quantified following APHA standard methods (2005). Free ammonia concentration was measured using the equation reported by Calli et al., (2005). Volatile fatty acids (VFAs) were measured using a gas chromatographer (Varian CP-3800) fitted with a flame ionisation detector coupled to a Nukol capillary column (30 m × 0.25 mm × 0.25 µm, Supelco). The injector and detector temperatures were 220 and 250 °C, respectively. The temperature was programmed to start at 150 °C for 3 min and increase to 180 °C at 15 °C/min. The use of Supelco Column for VFA measurement gives better results when applying an initial oven temperature of 150 ºC to avoid peak overlapping when high acetic and propionic acid concentration are to be expected. Helium was the carrier gas, and calibration proceeded using a commercial C2-C7 standard mixture of VFAs (Supelco, Germany) up to a detection limit of 5.0 mg/L. Samples preparation proceeded by centrifugation, for 10 min at 3500 g, to separate the supernatant which is filtrated with a 0.45 µm cellulose filter. Methane production was calculated using recorded data of biogas production and composition as detailed in the Micro-plant description section in electronic supplementary material, see Table ESM_1. The calibration of online analytical equipment proceeded as recommended by the manufacturer.

Mass balance calculations were performed using manual data recording, results from sampling of the feeding material and digester liqueur and data obtained from the automatic register of the control unit. A description of the parameters recorded is provided in supplementary information Table ESM_2. These data along with the different parameters of the process were used for calculating mass balances and biological indicators of reactor performance as it is: daily biogas production (L/d), methane yield (mL CH_4/ g VS), methane production performance (mL CH_4/ $m^3_{reactor}$ g VS). Unlike laboratory tests where feeding conditions are completely controlled, during evaluation of the prototype the operating conditions are subject to waste availability, variability in its composition and the degree of dilution at which the pretreatment unit is operating. These conditions directly influence parameters such as hydraulic retention time (HRT), organic loading rate (OLR) and biogas production. To facilitate data evaluation and estimate plant performance, process parameters obtained during seven days were averaged.

2.4. Energy analysis

This analysis evaluated the energy demand for each operating state of the micro-plant (Operation, Recirculation, Heating, Feeding and Grinding) and each state of the different actuators (main pump, submersible pump, heating pump, stirrer and control unit). The energy analysis of the pretreatment unit was based on the daily quantities of crushed waste, which was manually recorded. The energy consumed was based on the demand of the grinder and the operating time (having a capacity of 6 kg/min of waste) and the supply of dilution liquid to the feeding unit which was done by gravity from the digestate storage tank.

The coefficient of overall process performance was evaluated using the thermal energy produced (COP_{th}). This coefficient was calculated as the ratio between the useful thermal energy produced and the thermal energy consumed. The coefficient of overall performance for electrical energy (COP_{el}) was calculated as the ratio between the electrical energy produced and the one consumed, therefore this coefficient represents a self-sufficiency rate.

ESM provides supplementary information on the active time for the different devices (Table ESM_3), which was used to estimate the energy demand of the prototype.

The temperature of the digester was maintained by using a heating system involving electric water heaters having two thermostats responsible for keeping the temperature of the water tank at 60 °C. When there was no demand for heating, the activation of the thermostats only takes place for maintaining the temperature at the established set-point, with a certain frequency, f1. On the contrary, during the heating state (when there was demand for heat) activation of the thermostats takes plays at a different frequency, f2, which was higher to meet the heat demand. The determination of the operating frequencies f1 and f2 allows differentiating between the energy consumed during the heating state (process heat (Q_{Proc})) and the energy consumed during the remaining operating states (maintenance heat (Q_{Man})). The summation of these two quantities accounts for energy needs associated with thermal purposes ($Q_{C.th}$). The electric water heaters had an associated energy meter, IWATION 3680W, which allowed the manual recording of energy consumption over time. This mode of operation implied that Q_{Man} had a permanent electricity consumption baseline. The amount of heat necessary for keeping the temperature of the reactor (Q_{Proc}) comprised two aspects, one for increasing the temperature of the feed (Q_{Feed}) and another regarding the loss of heat through reactor walls and piping (Q_{Losses}).

2.5. Analysis of scenarios

This research work deals with the installation and operation of a prototype thus low OLR issues were associated with initial tests due to acclimation of the anaerobic microflora. The present manuscript shows results obtained from the performance of this unit when treating highly degradable wastes, thus acid build-up limited the treatment capacity of this plant. Operation of this plant was continued beyond the present state here reported but data obtained from subsequent experimental stages were not reported in the present manuscript due to commercial decisions regarding companies investing in this prototype. The performance of the micro-plant was evaluated considering two different scenarios A and B. Scenario A is used to measure the energy demand of the installation without taking into account biogas valorisation. The data associated with the total energy demand of the plant (Q_{Total}) are classified according to their purpose, that is, the energy for electrical purposes ($Q_{C.el}$), and that for thermal purposes ($Q_{C.th}$).

Scenario B considers the inclusion of a hypothetical micro-cogeneration system. The energy balance is then evaluated assuming the production of thermal and electrical energy. The heat needed for the process was assumed to be provided by a micro combined heat and power (CHP) system Ecowill cogenerator (Roselli et al., 2011; Staffell et al., 2015). This unit has an electrical and thermal output of 1 kW and 2.8 kW, respectively with overall energy efficiency of 85% (electrical efficiencies of 22.5% and thermal efficiency of 63.0%). The energy contained in biogas is denoted as Q_{Biogas}, and that derived from the hypothetical valorisation using the CHP unit was denoted as EG_{elec} and $EG_{thermal}$, regarding the electrical and thermal energy produced.

Low OLR directly affects biogas yields, therefore hypothetical performance of this plant was evaluated at a higher value, considering that electric and thermal related parameters were already measured during the first experimental stage and are independent of the OLR applied. The efficiency parameters of the micro-plant were also estimated assuming the application of a theoretical OLR of 2.7 g VS/L d and a content of volatile solids in the reactor of 29 g/L, based on the operating values reported by Banks et al. (2011). The low heating value of methane was 35.7 kJ/m^3 (Demosthenous et al., 2016).

3. Results and discussion

3.1. Reactor performance

The temporal distribution shown in Fig. 2 represents the mass flows of the auxiliary tank. These flows correspond to the ground waste streams, dilution liquid and the flow of feeding substrate to the reactor. Since the flows are represented for the pre-treatment storage tank, the feeding flow in the diagram is represented as a negative value. Therefore, in this diagram incoming materials to the pretreatment unit have positive values, and the feeding volume into the anaerobic reactor withdrawn from the pretreatment tank has negative values. The tank acts as a buffer system for the daily variations of the amount of waste received. The temporal distribution of the feeding substrate into the reactor is different from that of the crushed waste. However, in the long term, the accumulated values are obviously equivalent. Difficulties associated with the operation of the prototype and the performance of the digestion process led to irregular feeding of the reactor. Fig. 2 also shows the start-up of the plant where low organic loadings are applied (represented as phase I). The increase in the organic loading was based on the performance of digestion and it is represented in the diagram as phase II. Finally phase III corresponds to a period where feeding to the reactor was not available due to technical problems at the Algodor plant.

During the total period of 106 days of pilot plant operation, the digestate had average TS concentrations of 3.05 ± 0.34% with a value of 48.1 ± 3.14% VS/TS, accounting for an average content of 14.7 g VS/L, which is within the range between 2.9 and 40 g VS/L reported by Banks et al. (2011) and Walker et al. (2017). The pilot plant operated in a stable form throughout the evaluation with the values of solid content measured for the digested material (TS and VS) reporting low variability. The pilot plant operated in a stable form throughout the evaluation with values of solid content (TS and VS) reporting low variability for the digested material.

Figure 2 here

Figure 3 shows the mass flow expressed as a mass loading rate for the different streams involved in the operation of the reactor. Fig. 3a shows the average values for the whole evaluation time, whereas Fig. 3b represents the period corresponding to days 55 to 90 inclusive. The operation of the prototype was characterised by a high recirculation rate accounting for an average value of 73.7% for the whole experimental term, whilst this value was slightly higher during the 55 – 90 day period, accounting for about 80% of the total mass flow entering into the reactor.

Figure 3 here

The mass flow expressed in terms of volatile solids is represented in Fig 3c and 3d for the average values obtained during the whole experimental period and also for days 55 – 90. The contrast between the two figures is clearly observed by the percentage associated with the recycling streams. This stream was characterised by a high value when the total mass flow is considered, but it only represents approximately 18% of the VS flow (see Fig 3c). Due to difficulties associated with manual data handling and plant operation, there was a disagreement of about 5% in closing the mass balance and this value was also observed when evaluating the period from days 55 to 90. This disagreement in volatile solid balance was however considered reasonable given the scale of operation and the heterogeneity of food wastes. Other authors report an acceptable mass balance disagreement of up to 9.4% for similar studies (Banks et al., 2011).

The removal of volatile solids attained during the digestion process was on average 93.1%, with these values decreasing slightly when the feeding was regularly available to 91.5% (days 55 - 90). This high value of solid destruction was associated with the high rate of recycling digestate back into the reactor, therefore affecting the residence time of microorganisms in a favourable way and increasing the capability of the reactor for degrading OLR supplied.

The parameters evaluating the performance of the digestion process are reported in Fig. 4, showing the daily biogas production, evolution of solid content in the reactor, pH, ammonia, VFA and alkalinity. Biogas production presented and increasing trend which was explained by the performance of the different digestion parameters. The volume of feeding to the reactor was based on the evolution of acid intermediaries and alkalinity, thus the increasing biogas production is in consonance with varying OLR.

The content of VS in the reactor was particularly low, being associated with the high degradation attained (Fig. 4b). This value was on average slightly higher than 40% not reaching values greater than 50% after day 30 of operation demonstrating the high stabilisation achieved during the degradation process. Organic loading was restricted based on the trend of the parameters being monitored during digestion. During the total period of 106 days, the pH was on average 8.10 ± 0.17 and remained stable until the end of operation (Fig. 4c). This parameter was not at all useful to establish feeding periods for the digester.

Figure 4 here

Ammonia nitrogen concentrations also remained stable at an average concentration of 5.10 ± 0.50 g NH_3-N/L. Although this value is considered high and may even be a cause of inhibition (Cabbai et al., 2016), in the present case, inhibition was avoided by restricting organic loading to the reactor attaining regular production of methane and high removal of volatile solids. The stable behaviour of pH was explained by the high alkalinity provided by ammonium levels in the digester. In this experimental work, high ammonia concentration was attributed to the inoculum used for the start-up process which was rich in ammoniacal nitrogen as it came from a co-digestion reactor treating a substrate from the meat industry. In addition, food waste had a C/N ratio of 18.7 ± 1.98 so that its nitrogen content did not represent a risk on the deficit of this nutrient.

There was a high recycling rate of digestate. During the plant daily operation, about 30% of supernatant was withdrawn from the digestate storage tank. The recycling of digestate was intended for solubilising the food waste and aid in further particle size reduction. Ammoniacal nitrogen was thus minimally depleted from the digestate since the content of feed material plus that from the recycling streams supports the retention of high nitrogen values inside the digester. Coupled to the high pH and ammonia values, was also alkalinity which presented an average value of 12.38 ± 1.74 g $CaCO_3$/L. The behaviour of this parameter was explained by ammoniacal nitrogen levels.

The reactor was initially fully loaded with inoculum, the daily addition of feed caused a slow increase in VS content and also in the amount of organics to be degraded by microorganisms. During the first days of operation the concentration of VFAs in the reactor was lower than 0.50 g/L as shown in Fig. 4d. From day 8 onwards, the content of total VFAs (TVFAs) presented a steep rise reaching values of 5.63 g/L, probably due to the high ammonia content. Feeding to the reactor was performed at low organic loadings, carefully increasing the incoming flow of this material to give enough time to anaerobic microflora to adapt to inhibitory conditions. VFA accumulation is likely to be a negative outcome due to toxic and inhibitory changes to the harmonious global interplay between living and non-living matter in the sludge. The increase in organic loading associated with phase II led to an accumulation in VFA.

The main acids measured were acetic and propionic, with the latter having values close to 3.0 g/L between days 10 and 40 of the experiment. Inhibitory values of 3.5 g/L of propionic acid have been reported by Ahring et al. (1995) whereas Fierro and co-workers (2016) evaluated a digester with propionic acid levels as high as 4.0 and 5.9 g/L when treating a mixture of swine manure and glycerine without reporting digester failure problems. Thus in the present study, high values of VFA were not believed to be responsible for irreversible inhibitory conditions during digester performance, which was demonstrated by their subsequent decrease. Although phase III was a period presenting low feeding, the decrease in VFA was associated mainly to the acclimation of microbial biomass since it took place earlier, just at the beginning of phase II.

Ammoniacal nitrogen has been linked to VFA imbalances in similar experiments. For instance, values in the range between 4.05 and 5.73 g NH_3-N/L strongly affected acidogenic microbes and methanogens causing activity loss of 56.5% for these later species (Chen et al., 2008). Basic pH is also linked to VFA imbalances. At pH > 8, methanogenic activity is likely to be damaged (Fisgativa et al., 2016). After the lapse of the first 50 days of the experiment, the concentration of propionic acid rapidly decreased but that of acetic acid increased. This is linked to a predictable acclimation of fast propionic acid utilisers to produce acetic acid that is slowly converted into methane by methanogens. Around day 65, propionic acid concentration was almost zero whilst acetic acid kept decreasing until reaching a concentration below detectable limits on day 100. At this point, the process could be considered stable as suggested by Chen et al., (2008) after observing similar patterns, and by

Yenigün and Demirel (2013) who ran a digester with ammonium concentrations > 5 g/L after an initial adaptation period.

Partial and total alkalinity (PA and TA) followed a similar trend (Fig. 4e). The high values reported for these two parameters were explained by the high ammonia concentration in the reactor. During the first 10 days, alkalinity decreased associated with the evolution of VFA increments. The high alkalinity observed in this reactor resulted in values of TVFA/alkalinity ratio of 0.33 as maximum (attained in day 23). Once VFA concentration was significantly reduced, alkalinity values experienced an increasing trend and the TVFA/alkalinity ratio was reduced to 0.22 on day 42 and continued to decrease thereafter. In the present work, alkalinity ratio (PA/TA) was stable throughout the whole process, always below 0.37. From day 80 onwards a decrease was observed along with a decrease in PA, indicating the process is reaching stable conditions as suggested by Ripley et al. (1986).

Table 2 shows the results derived from the calculated parameters for evaluating biological performance. Methane yield was lower than that reported by Fisgativa et al. (2016) for wastes of similar composition (460 mL/g VS). However, this value was obtained from biochemical methane potential tests that were performed under batch conditions and therefore the hydraulic dynamic of the reactor was different. The value obtained in this experimental work was within the expected range considering the negative effect exerted by high ammonia concentration. Thus, the production obtained in this case was slightly lower than that reported by other authors under similar experimental conditions. Banks et al. (2011) reported a value of 402 mL/g VS for an industrial anaerobic digestion plant of 900 m^3 of reactor volume. Algapani et al (2019) obtained a value of 510 mL g/VS in a two-phase configuration using a recycling stream for producing H_2 in the first reactor and CH_4 in the second one (with working volumes of 2 L and 4.5 L and also treating food waste as substrate). Considering the composition of wastes used in the present research, the theoretical value calculated using the Buswell equation was 494 mL/g VS. The methane yield obtained from the reactor was 37% lower than that of the theoretical one. When taking as reference data obtained by Fisgativa et al. (2016) and Banks et al. (2011), the reduction accounts for 16.5% on average.

Table 2 here

The average OLR for the period between days 55 and 90 was 1.06 g VS/L d, comparable to the lowest range used by Cabbai et al. (2016) and less than the one evaluated by Bolzonella et al. (2019) with a value of 3.5 g VS/L d. During the operation of the unit, the irregularity in the supply of the feed and the recirculation rate applied caused that only for days 55 - 90 feeding of the digester could be performed on a regular basis with a low OLR and an average HRT of 55 days. However, the average HRT for the whole experiment (106 d) was about 80 days.

3.2. Energy assessment

The consumption of energy for the period analysed reached a value (Q_{Total}) of 9,155 kJ/kg VS, which is equivalent to a specific power of 100 W/m^3. This consumption is slightly higher than that obtained by Walker et al. (2017) reporting a value of 75.1 W/m^3 for a digester with size 3.6 times smaller. In the present research, the thermal energy needed for maintaining the temperature of water tanks was obtained by the use of electric heaters (Q_{Man}). If this energy value was subtracted, so that both results became comparable, the value of specific power would be 68.8 W/m^3 which was slightly lower. The scale factor is an important parameter to be considered because the increase in reactor scale causes a decrease in energy demand. This is due to the fact that some energy consumption is associated with auxiliary equipment and remains approximately constant regardless of the size of the installation, such as the control unit. Others do not experience a linear increase proportional to the size of the plant, as it is the case of the heating system, because the surface of thermal leaks does not evolve proportionally to the volume of the reactor. The relationship between surface and volume decreases as the size of the digester increases.

The energy demand was analysed considering the different operating states of the prototype. The summation of categories in Fig. 5a accounts for the total energy demand of the prototype (9155 kJ/kg VS). The Heating state reports the highest energy consumption with a value of 58.4% of the total energy demand, followed by the Operation state that represents 35.9% (see Fig. 5). The energy consumption associated with the states of Recirculation, Feeding and Grinding was reduced, not exceeding 5.8% of the total consumption between the three operating states. This low energy demand was due to the fact that, although several actuators are involved in these states, the time needed for their activity is only 4.1%.

The plant is 73.3% of the time in the Operation state, for which the control unit monitors the process variables, controls the evacuation of biogas and records data. Although these tasks do not imply a high energy demand, the long-time associated with this state along with the use of thermal energy based on electric heaters (having high energy consumption) causes an exacerbated demand for energy. The availability of a different thermal source would reduce the energy needs of this operating state from 3,283 kJ/kg VS to 401 kJ/kg VS, which was the value associated with the control unit. This reduction would translate into the global context of the plant in 87.8% reduction in energy demand.

Within the heating state, the energy demand for producing hot water accounts for 33.5% (3,065 kJ/kg VS) of the total energy needs, whereas the pumping of digestate through the heating exchange system accounts for 22.3% (2,041 kJ/kg VS). These results show the high amount of energy necessary for keeping the temperature of the digester. The process was set to be in the Recirculation state 2.8% of the time (10 min ON and 5h and 50 min OFF) (Meroney and Colorado, 2009), which was equivalent to a renewal rate of 1.37 renewals/d, and a time of renovation equivalent to 17.5 h.

Fig. 5a shows that the system had a recirculation rate apparently lower than the set value. This is explained by the fact that heating and recirculation tasks were performed simultaneously but the system gives priority to heating over recirculation. In this prototype heating also implies recirculating. If the periods associated with heating state are to be considered as recirculating time, an effective recirculation regime is obtained with an average renewal time very close to 2 h, exceeding the values of 4 h used by Cabbai et al. (2016). Despite this, a lower renewal rate could be used since high recirculation regimes do not show remarkable improvements in the amount of biogas produced (Lindmark et al., 2014), contributing to decreasing energy needs of the prototype.

Figure 5 here

Regarding the energy consumption by the different devices, it should be noted that the electric water heaters are responsible for 66.6% of the energy consumed. This is mainly due to the energy associated with the Operation state and the high amount of energy required for maintaining process temperature, which takes place during the Heating state. The energy requirement of the main pump accounts for 24.0% of total devices. Most of this demand

(22.3%) was associated with the state of Heating. It should also be noted, that the energy associated with the main pump during the Heating state also has an additional effect on the demand of the recirculation. Energy consumption of the rest of the actuators was considered negligible compared to the total energy needs.

The consumption of energy for electrical uses ($Q_{C.el}$) accounts for 3,061 kJ/kg VS which was equivalent to an average power of 31.88 W/m^3. This result is considerably lower than that obtained by Walker et al. (2017) with a value of 75.1 W/m^3. The scale factor is obviously behind the lower demand of the present prototype. These authors used a digester size 3.6 times smaller than the present one. 54% of its consumption was due to data recording, which can be considered to have an energy demand approximately constant regardless of the size of the plant. The thermal energy demand to keep process temperature (Q_{Proc}) corresponds to a power of 31.90 W/m^3 (3065 kJ/kg VS), with this value being also less than the obtained by Walker et al. (2017), due to the scale factor previously discussed, 40 W/m^3.
Q_{Proc} can be divided in two categories, one is the energy necessary for heating the feed (Q_{Feed}) and the other is the heat associated with intrinsic losses (Q_{Losses}). The plant demanded 1,934 kJ/kg VS for heating the feed to the process temperature and had thermal losses of 1,131 kJ/kg VS associated with inefficiencies of insulation, representing 36.9% of the thermal demand of the unit.

The biogas yield of the prototype was 0.56 m^3/kg VS with an average methane richness of 64.3% (equivalent to an energy value (Q_{Biogas}) of 12,864 kJ/kg VS). If the valorisation of biogas is assumed by means of a CHP unit (conditions stated in scenario B), an electrical energy generation (EG_{elec}) of 2,894 kJ/kg VS and thermal energy ($EG_{thermal}$) of 8,104 kJ/kg VS would be expected. Taking into account that 3,065 kJ/kg VS are needed for meeting the thermal demand of the plant, then 5,039 kJ/kg VS would be available for other applications. Energy flows are represented by a Sankey diagram shown in Fig. 6.

Figure 6 here

For the calculated energy flows, a COP_{th} value of 2.64 and a COP_{el} value of 0.95 were obtained, thus in the case of electricity, the micro-plant produces less energy than that needed for its operation when an efficiency coefficient of 22.5% was assumed. However, when results are recalculated considering the use of a CHP system with efficiencies of η_{el} = 25% and η_{th} = 50%, for electricity and heat respectively, the new coefficients are COP_{th} = 2.09 and a COP_{el} = 1.05. Comparing these values with those obtained by Walker et al. (2017) (COP_{th} = 5.55 and COP_{el} = 1.47) the prototype reported in this study gives a lower performance explained by the 33.8% lower OLR of operation.

The assessment of the reactor was also performed considering a theoretical increase in OLR to a value of 2.7 g VS/L d. Results for this scenario are shown in blue colour in Fig. 8 affecting some of the energy quantities estimated. These results considered the same methane yield. The increase in OLR caused a proportional increase in the electricity produced and thermal energy available. Regarding the demand for thermal energy, a decrease in the theoretical operating point relative to the one tested was observed, from 3,065 to 2,407 kJ/kg VS. This decrease is explained by the fact that thermal losses of the digester remain constant regardless of the OLR applied. The demand of electrical energy has a similar trend, resulting in a decrease when evaluating the theoretical operating point, from 3,061 (experimental value) to 1,725 kJ/kg VS. In this case, the decrease is greater because the consumption of the different electric actuators remains about the same in spite of the increase in OLR, but for the main pump and heating pump which have higher energy demands. Values of COP_{th} and COP_{el} for this theoretical point were also calculated. The COP_{th} is increased from 2.64 to 3.37, because heat losses remain constant causing a better thermal efficiency per unit of volatile solid. The COP_{el} has a behaviour similar to that of COP_{th}, due to the same reason, in this case it rises from 0.95 to 1.68, because much of the electricity demand of the plant presents minimum variations regarding the OLR applied.

4. Conclusions

The assessment of the prototype was successfully carried out obtaining a methane yield of 360 L/kg VS at an OLR of 1.06 g VS/L d (calculated using data from days 55 to 90 of operation) in spite of the high concentrations of ammoniacal nitrogen (5100 mg/L). Most of the thermal energy requirements were associated with the raise of temperature of the feed to process conditions. The efficiency of the heating system was crucial, since a large amount of both electrical and thermal energy was necessary for operation, accounting for more than 80% of the total energy demand of the installation.

The energy assessment of the process was carried out assuming the incorporation of a combined heat and power (CHP) unit for valorising biogas and avoiding the use of electrical heaters for the supply of heat. The Sankey diagram showed that the process was thermally sustainable, since only 37.8% of the useful thermal energy generated by the CHP system would be used to meet the heat demand of the reactor. However, it should be pointed out that 36.9% of the energy used in heating was lost due to thermal losses associated with the reactor external surface and piping along with inefficiencies of the heating exchange system. The COP_{el} parameter was slightly lower than the unit indicating that the micro-plant was close to reaching self-sufficiency. The OLR of the process was decisive for the overall performance directly influencing the energy consumption per unit of volatile solid and therefore the energy efficiency indicators. The evaluation of operating parameters when estimated at an OLR of 2.7 g VS/ L d would result in values of 3.37 for the COP_{th} and 1.68 for the COP_{el}

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Supplementary Material Caption

ESM. Details of Micro plant description and operation. Parameters recorded during the operation of the plant. Operating time of the different equipment constituting the micro-plant

Essential title page information

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Title: Performance evaluation of a small-scale digester for achieving decentralised management of waste

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Abstract

The performance of a small-scale prototype digestion plant (7.2 m^3 working volume) intended for decentralised operation was evaluated considering energy efficiency and technical suitability for biogas valorisation in producing electrical and thermal energy. The digester operated in recirculation mode to enhance organic matter conversion and improve volatile solid degradation. An energy assessment of the process assumed the incorporation of a combined heat and power (CHP) unit. The coefficient of overall performance of the plant for electrical energy (COP_{el}) was 0.95 — this values was estimated at an electrical efficiency of 22.5% and represents the ratio between energy production and consumption — for a methane yield of 360 L/kg VS and an organic loading rate (OLR) of 1.06 g VS/L d. This parameter was slightly lower than the unit thus indicating that the micro-plant was close to attaining self-sufficiency regarding electrical energy use. The temperature increase of the feed to process conditions supposed a significant amount of thermal energy which highly compromised the efficiency when operating at low organic load, thus accounting for more than 80% of the total energy demand of the installation. When the energy assessment of the process was performed at higher OLR of 2.7 g VS/ L d, the resulting COP_{el} value was1.68, demonstrating the feasibility of this configuration for decentralised digestion.

Keywords: micro-plant energy performance, food waste digestion, decentralised waste treatment, prototype evaluation, anaerobic digestion

Nomenclature

CHP: combined heat and power

COP_{th}: Coefficient of overall performance, thermal energy

COP_{el}: Coefficient of overall performance, electrical energy

EG_{elec}: Electrical energy generation

$EG_{thermal}$: Thermal energy generation

HRT: Hydraulic retention time

OLR: organic loading rate

PA: Partial alkalinity

Q_{Biogas}: Energy contained in biogas

$Q_{C.el}$: Energy needed for electrical purposes

$Q_{C.th}$: Energy needed for thermal purposes

Q_{Feed}: Heat needed for increasing the temperature of the feed to process conditions

Q_{Losses}: Heat loss through reactor walls and piping

Q_{Man}: Maintenance heat

Q_{Proc}: Process heat

Q_{Total}: Total energy demand of the plant

TA: Total alkalinity

VFA: Volatile fatty acid

1. Introduction

Decentralised anaerobic digestion is a promising alternative for low population density areas or the treatment of a relatively small amount of wastes produced seasonally. This management alternative may also be appropriate when the distance to a large-scale centralised plant is less attractive on the grounds of sustainability due to transport impacts. Anaerobic digestion is an efficient technology for treating organic substrates capable of increasing the contribution of renewables to the overall energy production matrix (González et al., 2020). The lack of aeration provides a great advantage due to the decrease in the implicit cost associated with energy demand. However, depending on scale, anaerobic digestion may be favoured for centralised treatment; whereas, composting may be preferred in the case of small on-farm for treating animal manure (Lin et al., 2019)

Another parameter of particular significance is the transport of residues. In reality, there are limits based on reasonableness to the distance from where organic wastes are collected and subsequently treated at a centralised plant. Low population density areas impose challenges to the extrapolation of conventional technologies due to the smaller scale of the treatment units, the correspondingly higher operating costs and the practical and sustainability logistics of waste transportation. A study performed by Piñas et al. (2019) for a Brazilian scenario showed that biogas plants using mono-substrates such as cattle manure presented economic viability for electrical power higher than 740 kWe whereas a co-digestion system presented economic viability for electrical power above 1000 kWe. This scale is unsuitable for many applications in rural areas due to seasonal production and the significantly lower amount of available waste.

The development of small-scale digestion plants for treating waste in rural regions is crucial. Decentralised biogas production from manure and crop residues is not currently economical or reliable because gas production usually has a low energy potential. Codigestion systems are not economically viable because of the costs associated with silages and transport of biomass material (Piñas et al., 2019). These factors have delayed development in the implementation of this type of technology. Transportation of feedstock must not exceed 20 – 30 km (two-way), since increasing this distance negatively affects the economics (Rajendran and Murthy, 2019) thus justifying the need for decentralised units. However, the high initial investment of these plants act as the main disincentive requiring support from fiscal subsidies (Win et al., 2017). The design of small biogas reactors also raises fundamental issues concerning the energy demand associated with pretreatment and achieving good mixing in the reactor in order to optimise gas yields when feeding combinations of dry and wet food wastes, or domestic and agricultural wastes (Radu et al., 2016).

Small-scale digestion plants have recently attracted considerable interest to shift towards a more decentralised biowaste management strategy. This approach offers advantages compared to the conventional centralised waste treatment associated with reduced transport requirements and the potential benefit of increased community involvement. The efficiency of the decentralised approach relies on a close integration of the whole treatment supply chain, from the generation of biowastes to the valorisation of biogas and digestate (Thiriet et al., 2020).

Production activities and small communities must meet certain requirements if the decentralised approach for waste treatment is to become feasible. The first requirement is that these activities must generate organic waste streams of a high organic content that is readily biodegradable and with high methane potential. The second requirement is that they must be associated with local electrical and thermal energy demands. These characteristics define a niche of activities that mainly focuses on agri-food industries (agri-food farms, food products industries, livestock farms), sewage treatment plants and food waste managers. Decentralised anaerobic digestion facilities could produce operational benefits such as, greater ease in the management of wastes, the possibility of having energy autonomously, as well as economic benefits linked to thermal energy generation and production of organic fertilisers and amendments (Anyaoku and Baroutian, 2018). Other benefits are the ability to handle and treat wastes using the proximity principle basis, as close to point of origin as possible, which can drastically reduce emissions and impacts associated with transport. In addition, there would be another series of short-term improvements such as the increase in the generation of distributed energy, which in turn would contribute to the stability of the electrical system and reduce the costs of transporting energy and potential pollution from large centres of generation.

The reduction of costs associated with logistics when treating wastes would also be part of the short-term improvement as well as the optimisation of municipal waste treatment (Wang et al., 2014). It is for these previous reasons that the possibility of using "flexible anaerobic digestion micro-plants" are promising at this moment. Here, these plants are defined as "micro" for their small treatment capacity, being limited to amounts of < 1,000 t/year (<15 kWel) (NNFCC, 2017). The term "flexible" refers to a type of facility, despite its small size, that is provided with a process control system with similar benefits to those of large centralised anaerobic digestion plants, being also capable to adapt to any particular need regarding the type of organic material, nutrient content and seasonal production of waste. These characteristics allow this type of decentralised plant to present an operational improvement over traditional centralised installations. Decentralised treatment units are being considered a more sustainable solution because energy requirements are much lower and sophisticated operation is avoided thus being easy to adapt to different geographic contexts (Lourenço and Nunes, 2020).

This research evaluates the suitability of a micro-digestion plant for the decentralised treatment of food wastes in a real environment. The novelty of this prototype is based on its capacity for valorising wastes near the source, reducing transport needs and activating the local economy. Developing a new solid waste management strategy based on small decentralised units offers new opportunities for implementing this model into developing countries and communities with a disperse population. These small treatment plants can promote community participation and avoid an undesirable accumulation of organic materials prone to degrade uncontrollably. Decentralised management of wastes offers several advantages but there is a lack of studies reporting on the electric and thermal performance of small-scale prototypes.

In the present study, the first objective was to evaluate the digestion process using a micro-plant of flexible configuration fed with substrates comparable to those composing the organic fraction derived from catering services. The second objective was to evaluate the energy efficiency of the micro-plant establishing an energy balance to assess thermal and electric performance. This manuscript, thus reports data for evaluating performance of the digestion process considering not only biological yields but also energy demands associated with the operation of small scale units.

2. Materials and methods

2.1. Inoculum and substrates

Food wastes used as substrates were obtained from a hostelry school for the training of cuisine professionals. The school was dedicated to the teaching of Italian, French, Mexican and Spanish cooks. Undesirable materials like packaging, containers, bones, cutlery and other non-degradable components were manually screened out to obtain a food waste fraction easy to handle for grinding machines and free of plastics and any other kind of inert components that would exert a detrimental effect on the quality of the digestate. Food waste was daily transported using 50 L closed steel vessels from the school to the Algodor plant (located in Toledo, Spain) where the prototype was installed. This plant is specialised in the biological treatment of organic wastes from a diversity of sources, including fruit and vegetable wastes from the fourth-range industry, food waste from the hostelry sector and those from the maintenance of green areas. Currently these materials are transformed by static pile composting. The digestion prototype was installed in this treatment centre, with the aim of evaluating the suitability of energy production from wastes received daily.

The waste was subjected to an initial triage to remove contaminants and record daily quantity of waste received. Periodic sampling was carried out for characterisation of in-coming material for quality control and quality assurance purposes. The weighing of food wastes was performed using an industrial floor scale balance with a precision of 0.5 kg (Steinberg

Systems SBS-BW-1T). After weighing, the waste was fed into the pretreatment unit for grinding and then onwards to the reactor (Fig. 1a). The chemical and physical characterisation of food wastes is summarised in Table 1.

Table 1 here

Inoculum used to seed the micro-plant anaerobic digester was a digestate obtained from a co-digestion plant treating a mixture of municipal solid wastes and wastes derived from a meat processing factory. The industrial digester was located at the solid waste treatment centre of South Madrid, Pinto. The total solid content of the inoculum was 41.0 ± 1.9 g/L with a volatile solid content of 18.4 ± 0.7 g/L. Once loaded into the digester, the inoculum was heated at 37 ± 1 ºC for 20 days without any additional material being fed to allow the removal of the "background" biogas production and to achieve maximum degradation of available organic materials before commencing with pilot plant operations.

Figure 1 here

2.1. Micro-plant description and operation

The operational units comprising the pilot plant are presented in Fig. 1b showing the main equipment and auxiliary components. A detailed description of the plant and controller actions are given in the electronic supplementary material (ESM). The digestion unit consists of a tank reactor, the operation of which depends exclusively on the multipurpose pump. The digester has a total volume of 8776 L with a working volume of 7200 L. Mixing is performed by recirculation of the digestate using a multipurpose pump. The digester operated under mesophilic conditions at 37 ± 1 ºC. The average organic loading rate was 0.68 kg VS/m^3 d. The prototype was kept in operation for 106 days.

Actual biogas production was compared with theoretical predictions from a simplified version of Simons and Buswell equation (Møller et al., 2004):

$$B_u\left(\frac{\mathrm{L\ CH_4}}{\mathrm{kg\ VS}}\right) = \frac{\frac{n}{2}+\frac{a}{8}+\frac{b}{4}}{12n+a+16b} * 22.4$$

Methane production was estimated by assuming that all organic material was converted into biogas (methane and carbon dioxide) along with water. The use of carbon for microbial growth and maintenance needs were not considered. This expression establishes as main elements of organic matter: carbon, hydrogen and oxygen using the empirical formula $C_nH_aO_b$. The ultimate methane production (Bu) was calculated based on the stoichiometric Buswell equation and using the gas ideal factor for estimating the volume of a gaseous substance (1 mol) at STP conditions.

Due to instabilities intrinsic to the plant operation, both the frequency and the feeding rate were variable throughout the trial to adapt to substrate availability. The period selected to assess plant performance was from day 55 to 90 included (35-day continuous period) since this period showed process stability in terms of feeding rate and methane production.

The waste received daily was incorporated into the process through the pretreatment unit. For this, the waste material was poured into the feeding hopper for grinding and then into the pretreatment unit. In this tank, the organic material was mixed with digestate from the digestate storage tank to dilute the mixture, accelerate hydrolysis and further reduce the particle size thus facilitating their introduction into the main digester. A dilution ratio was established to attain a volumetric proportion of 200 L food waste/m^3. The feeding procedure involved manual registration of parameters that are not automatically recorded by the control unit, as it is: the amount of ground waste, the levels of the pretreatment unit and digestate storage tank along with readings of energy consumption of electric devices. During the digestion test, samples of the feed and digestate were regularly collected for characterisation at a frequency of once or twice each week. Difficulties associated with the operation of the prototype led to an irregular feeding of the reactor. The start-up of the plant was carried out at low organic loadings and was denoted phase I. Phase II corresponds to an increase in the organic loading which was based on the performance of digestion. Phase III was the last stage and corresponds to a period where feeding to the reactor was not available due to technical problems at the Algodor plant. The experimental period was established based on the time indicated by project activities to test and evaluate energetic performance of the prototype for obtaining a commercial and flexible unit capable of treating a great variety of wastes at small scale. For this reason, the prototype was not tested for a longer period using this type of feeding.

The prototype control unit has four operating states (Operation, Heating, Recirculation and Feeding) to carry out the digestion process. Based on the values of the process variables and the operating instructions, the control unit was responsible for activating the corresponding operating status, so that the process was carried out following the operating instructions. There was another operating state (Grinding) that is outside the control capacity of the control unit, since this was done manually, activities related to this later state took place simultaneously to any of the other operating states. The recirculation frequency of the mixing pump was 10 min every 6 hours, which for the installed device having a volumetric flow of 14.82 m^3/h, represents a turnover time of 17 h. The operational turnover time was affected by the heating needs, which are met by turning on the multipurpose pump and the heat exchanger, thus reducing this parameter. The heating system was capable of maintaining the temperature at the set value with a dead band of 1 ºC.

2.3. Chemical analyses and data recording

Total alkalinity (TA), partial alkalinity (PA), pH, total and volatile solids (TS and VS) and ammonia nitrogen (measured via selective electron) were quantified following APHA standard methods (2005). Free ammonia concentration was measured using the equation reported by Calli et al., (2005). Volatile fatty acids (VFAs) were measured using a gas chromatographer (Varian CP-3800) fitted with a flame ionisation detector coupled to a Nukol capillary column (30 m × 0.25 mm × 0.25 µm, Supelco). The injector and detector temperatures were 220 and 250 °C, respectively. The temperature was programmed to start at 150 °C for 3 min and increase to 180 °C at 15 °C/min. The use of Supelco Column for VFA measurement gives better results when applying an initial oven temperature of 150 ºC to avoid peak overlapping when high acetic and propionic acid concentration are to be expected. Helium was the carrier gas, and calibration proceeded using a commercial C2-C7 standard mixture of VFAs (Supelco, Germany) up to a detection limit of 5.0 mg/L. Samples preparation proceeded by centrifugation, for 10 min at 3500 g, to separate the supernatant which is filtrated with a 0.45 µm cellulose filter. Methane production was calculated using recorded data of biogas production and composition as detailed in the Micro-plant description section in electronic supplementary material, see Table ESM_1. The calibration of online analytical equipment proceeded as recommended by the manufacturer.

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Mass balance calculations were performed using manual data recording, results from sampling of the feeding material and digester liqueur and data obtained from the automatic register of the control unit. A description of the parameters recorded is provided in supplementary information Table ESM_2. These data along with the different parameters of the process were used for calculating mass balances and biological indicators of reactor performance as it is: daily biogas production (L/d), methane yield (mL CH_4/ g VS), methane production performance (mL CH_4/ $m^3_{reactor}$ g VS). Unlike laboratory tests where feeding conditions are completely controlled, during evaluation of the prototype the operating conditions are subject to waste availability, variability in its composition and the degree of dilution at which the pretreatment unit is operating. These conditions directly influence parameters such as hydraulic retention time (HRT), organic loading rate (OLR) and biogas production. To facilitate data evaluation and estimate plant performance, process parameters obtained during seven days were averaged.

2.4. Energy analysis

This analysis evaluated the energy demand for each operating state of the micro-plant (Operation, Recirculation, Heating, Feeding and Grinding) and each state of the different actuators (main pump, submersible pump, heating pump, stirrer and control unit). The energy analysis of the pretreatment unit was based on the daily quantities of crushed waste, which was manually recorded. The energy consumed was based on the demand of the grinder and the operating time (having a capacity of 6 kg/min of waste) and the supply of dilution liquid to the feeding unit which was done by gravity from the digestate storage tank.

The coefficient of overall process performance was evaluated using the thermal energy produced (COP_{th}). This coefficient was calculated as the ratio between the useful thermal energy produced and the thermal energy consumed. The coefficient of overall performance for electrical energy (COP_{el}) was calculated as the ratio between the electrical energy produced and the one consumed, therefore this coefficient represents a self-sufficiency rate.

ESM provides supplementary information on the active time for the different devices (Table ESM_3), which was used to estimate the energy demand of the prototype.

The temperature of the digester was maintained by using a heating system involving electric water heaters having two thermostats responsible for keeping the temperature of the water tank at 60 °C. When there was no demand for heating, the activation of the thermostats only takes place for maintaining the temperature at the established set-point, with a certain frequency, f1. On the contrary, during the heating state (when there was demand for heat) activation of the thermostats takes plays at a different frequency, f2, which was higher to meet the heat demand. The determination of the operating frequencies f1 and f2 allows differentiating between the energy consumed during the heating state (process heat (Q_{Proc})) and the energy consumed during the remaining operating states (maintenance heat (Q_{Man})). The summation of these two quantities accounts for energy needs associated with thermal purposes ($Q_{C.th}$). The electric water heaters had an associated energy meter, IWATION 3680W, which allowed the manual recording of energy consumption over time. This mode of operation implied that Q_{Man} had a permanent electricity consumption baseline. The amount of heat necessary for keeping the temperature of the reactor (Q_{Proc}) comprised two aspects, one for increasing the temperature of the feed (Q_{Feed}) and another regarding the loss of heat through reactor walls and piping (Q_{Losses}).

2.5. Analysis of scenarios

This research work deals with the installation and operation of a prototype thus low OLR issues were associated with initial tests due to acclimation of the anaerobic microflora. The present manuscript shows results obtained from the performance of this unit when treating highly degradable wastes, thus acid build-up limited the treatment capacity of this plant. Operation of this plant was continued beyond the present state here reported but data obtained from subsequent experimental stages were not reported in the present manuscript due to commercial decisions regarding companies investing in this prototype. The performance of the micro-plant was evaluated considering two different scenarios A and B. Scenario A is used to measure the energy demand of the installation without taking into account biogas valorisation. The data associated with the total energy demand of the plant (Q_{Total}) are classified according to their purpose, that is, the energy for electrical purposes ($Q_{C.el}$), and that for thermal purposes ($Q_{C.th}$).

Scenario B considers the inclusion of a hypothetical micro-cogeneration system. The energy balance is then evaluated assuming the production of thermal and electrical energy. The heat needed for the process was assumed to be provided by a micro combined heat and power (CHP) system Ecowill cogenerator (Roselli et al., 2011; Staffell et al., 2015). This unit has an electrical and thermal output of 1 kW and 2.8 kW, respectively with overall energy efficiency of 85% (electrical efficiencies of 22.5% and thermal efficiency of 63.0%). The energy contained in biogas is denoted as Q_{Biogas}, and that derived from the hypothetical valorisation using the CHP unit was denoted as EG_{elec} and $EG_{thermal}$, regarding the electrical and thermal energy produced.

Low OLR directly affects biogas yields, therefore hypothetical performance of this plant was evaluated at a higher value, considering that electric and thermal related parameters were already measured during the first experimental stage and are independent of the OLR applied. The efficiency parameters of the micro-plant were also estimated assuming the application of a theoretical OLR of 2.7 g VS/L d and a content of volatile solids in the reactor of 29 g/L, based on the operating values reported by Banks et al. (2011). The low heating value of methane was 35.7 kJ/m^3 (Demosthenous et al., 2016).

3. Results and discussion

3.1. Reactor performance

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The temporal distribution shown in Fig. 2 represents the mass flows of the auxiliary tank. These flows correspond to the ground waste streams, dilution liquid and the flow of feeding substrate to the reactor. Since the flows are represented for the pre-treatment storage tank, the feeding flow in the diagram is represented as a negative value. Therefore, in this diagram incoming materials to the pretreatment unit have positive values, and the feeding volume into the anaerobic reactor withdrawn from the pretreatment tank has negative values. The tank acts as a buffer system for the daily variations of the amount of waste received. The temporal distribution of the feeding substrate into the reactor is different from that of the crushed waste. However, in the long term, the accumulated values are obviously equivalent. Difficulties associated with the operation of the prototype and the performance of the digestion process led to irregular feeding of the reactor. Fig. 2 also shows the start-up of the plant where low organic loadings are applied (represented as phase I). The increase in the organic loading was based on the performance of digestion and it is represented in the diagram as phase II. Finally phase III corresponds to a period where feeding to the reactor was not available due to technical problems at the Algodor plant.

During the total period of 106 days of pilot plant operation, the digestate had average TS concentrations of 3.05 ± 0.34% with a value of 48.1 ± 3.14% VS/TS, accounting for an average content of 14.7 g VS/L, which is within the range between 2.9 and 40 g VS/L reported by Banks et al. (2011) and Walker et al. (2017). The pilot plant operated in a stable form throughout the evaluation with the values of solid content measured for the digested material (TS and VS) reporting low variability. The pilot plant operated in a stable form throughout the evaluation with values of solid content (TS and VS) reporting low variability for the digested material.

Figure 2 here

Figure 3 shows the mass flow expressed as a mass loading rate for the different streams involved in the operation of the reactor. Fig. 3a shows the average values for the whole evaluation time, whereas Fig. 3b represents the period corresponding to days 55 to 90 inclusive. The operation of the prototype was characterised by a high recirculation rate accounting for an average value of 73.7% for the whole experimental term, whilst this value was slightly higher during the 55 – 90 day period, accounting for about 80% of the total mass flow entering into the reactor.

Figure 3 here

The mass flow expressed in terms of volatile solids is represented in Fig 3c and 3d for the average values obtained during the whole experimental period and also for days 55 – 90. The contrast between the two figures is clearly observed by the percentage associated with the recycling streams. This stream was characterised by a high value when the total mass flow is considered, but it only represents approximately 18% of the VS flow (see Fig 3c). Due to difficulties associated with manual data handling and plant operation, there was a disagreement of about 5% in closing the mass balance and this value was also observed when evaluating the period from days 55 to 90. This disagreement in volatile solid balance was however considered reasonable given the scale of operation and the heterogeneity of food wastes. Other authors report an acceptable mass balance disagreement of up to 9.4% for similar studies (Banks et al., 2011).

The removal of volatile solids attained during the digestion process was on average 93.1%, with these values decreasing slightly when the feeding was regularly available to 91.5% (days 55 - 90). This high value of solid destruction was associated with the high rate of recycling digestate back into the reactor, therefore affecting the residence time of microorganisms in a favourable way and increasing the capability of the reactor for degrading OLR supplied.

The parameters evaluating the performance of the digestion process are reported in Fig. 4, showing the daily biogas production, evolution of solid content in the reactor, pH, ammonia, VFA and alkalinity. Biogas production presented and increasing trend which was explained by the performance of the different digestion parameters. The volume of feeding to the reactor was based on the evolution of acid intermediaries and alkalinity, thus the increasing biogas production is in consonance with varying OLR.

The content of VS in the reactor was particularly low, being associated with the high degradation attained (Fig. 4b). This value was on average slightly higher than 40% not reaching values greater than 50% after day 30 of operation demonstrating the high stabilisation achieved during the degradation process. Organic loading was restricted based on the trend of the parameters being monitored during digestion. During the total period of 106 days, the pH was on average 8.10 ± 0.17 and remained stable until the end of operation (Fig. 4c). This parameter was not at all useful to establish feeding periods for the digester.

Figure 4 here

Ammonia nitrogen concentrations also remained stable at an average concentration of 5.10 ± 0.50 g NH_3-N/L. Although this value is considered high and may even be a cause of inhibition (Cabbai et al., 2016), in the present case, inhibition was avoided by restricting organic loading to the reactor attaining regular production of methane and high removal of volatile solids. The stable behaviour of pH was explained by the high alkalinity provided by ammonium levels in the digester. In this experimental work, high ammonia concentration was attributed to the inoculum used for the start-up process which was rich in ammoniacal nitrogen as it came from a co-digestion reactor treating a substrate from the meat industry. In addition, food waste had a C/N ratio of 18.7 ± 1.98 so that its nitrogen content did not represent a risk on the deficit of this nutrient.

There was a high recycling rate of digestate. During the plant daily operation, about 30% of supernatant was withdrawn from the digestate storage tank. The recycling of digestate was intended for solubilising the food waste and aid in further particle size reduction. Ammoniacal nitrogen was thus minimally depleted from the digestate since the content of feed material plus that from the recycling streams supports the retention of high nitrogen values inside the digester. Coupled to the high pH and ammonia values, was also alkalinity which presented an average value of 12.38 ± 1.74 g $CaCO_3$/L. The behaviour of this parameter was explained by ammoniacal nitrogen levels.

The reactor was initially fully loaded with inoculum, the daily addition of feed caused a slow increase in VS content and also in the amount of organics to be degraded by microorganisms. During the first days of operation the concentration of VFAs in the reactor was lower than 0.50 g/L as shown in Fig. 4d. From day 8 onwards, the content of total VFAs (TVFAs) presented a steep rise reaching values of 5.63 g/L, probably due to the high ammonia content. Feeding to the reactor was performed at low organic loadings, carefully increasing the incoming flow of this material to give enough time to anaerobic microflora to adapt to inhibitory conditions. VFA accumulation is likely to be a negative outcome due to toxic and inhibitory changes to the harmonious global interplay between living and non-living matter in the sludge. The increase in organic loading associated with phase II led to an accumulation in VFA.

The main acids measured were acetic and propionic, with the latter having values close to 3.0 g/L between days 10 and 40 of the experiment. Inhibitory values of 3.5 g/L of propionic acid have been reported by Ahring et al. (1995) whereas Fierro and co-workers (2016) evaluated a digester with propionic acid levels as high as 4.0 and 5.9 g/L when treating a mixture of swine manure and glycerine without reporting digester failure problems. Thus in the present study, high values of VFA were not believed to be responsible for irreversible inhibitory conditions during digester performance, which was demonstrated by their subsequent decrease. Although phase III was a period presenting low feeding, the decrease in VFA was associated mainly to the acclimation of microbial biomass since it took place earlier, just at the beginning of phase II.

Ammoniacal nitrogen has been linked to VFA imbalances in similar experiments. For instance, values in the range between 4.05 and 5.73 g NH_3-N/L strongly affected acidogenic microbes and methanogens causing activity loss of 56.5% for these later species (Chen et al., 2008). Basic pH is also linked to VFA imbalances. At pH > 8, methanogenic activity is likely to be damaged (Fisgativa et al., 2016). After the lapse of the first 50 days of the experiment, the concentration of propionic acid rapidly decreased but that of acetic acid increased. This is linked to a predictable acclimation of fast propionic acid utilisers to produce acetic acid that is slowly converted into methane by methanogens. Around day 65, propionic acid concentration was almost zero whilst acetic acid kept decreasing until reaching a concentration below detectable limits on day 100. At this point, the process could be considered stable as suggested by Chen et al., (2008) after observing similar patterns, and by

Yenigün and Demirel (2013) who ran a digester with ammonium concentrations > 5 g/L after an initial adaptation period.

Partial and total alkalinity (PA and TA) followed a similar trend (Fig. 4e). The high values reported for these two parameters were explained by the high ammonia concentration in the reactor. During the first 10 days, alkalinity decreased associated with the evolution of VFA increments. The high alkalinity observed in this reactor resulted in values of TVFA/alkalinity ratio of 0.33 as maximum (attained in day 23). Once VFA concentration was significantly reduced, alkalinity values experienced an increasing trend and the TVFA/alkalinity ratio was reduced to 0.22 on day 42 and continued to decrease thereafter. In the present work, alkalinity ratio (PA/TA) was stable throughout the whole process, always below 0.37. From day 80 onwards a decrease was observed along with a decrease in PA, indicating the process is reaching stable conditions as suggested by Ripley et al. (1986).

Table 2 shows the results derived from the calculated parameters for evaluating biological performance. Methane yield was lower than that reported by Fisgativa et al. (2016) for wastes of similar composition (460 mL/g VS). However, this value was obtained from biochemical methane potential tests that were performed under batch conditions and therefore the hydraulic dynamic of the reactor was different. The value obtained in this experimental work was within the expected range considering the negative effect exerted by high ammonia concentration. Thus, the production obtained in this case was slightly lower than that reported by other authors under similar experimental conditions. Banks et al. (2011) reported a value of 402 mL/g VS for an industrial anaerobic digestion plant of 900 m^3 of reactor volume. Algapani et al (2019) obtained a value of 510 mL g/VS in a two-phase configuration using a recycling stream for producing H_2 in the first reactor and CH_4 in the second one (with working volumes of 2 L and 4.5 L and also treating food waste as substrate). Considering the composition of wastes used in the present research, the theoretical value calculated using the Buswell equation was 494 mL/g VS. The methane yield obtained from the reactor was 37% lower than that of the theoretical one. When taking as reference data obtained by Fisgativa et al. (2016) and Banks et al. (2011), the reduction accounts for 16.5% on average.

Table 2 here

The average OLR for the period between days 55 and 90 was 1.06 g VS/L d, comparable to the lowest range used by Cabbai et al. (2016) and less than the one evaluated by Bolzonella et al. (2019) with a value of 3.5 g VS/L d. During the operation of the unit, the irregularity in the supply of the feed and the recirculation rate applied caused that only for days 55 - 90 feeding of the digester could be performed on a regular basis with a low OLR and an average HRT of 55 days. However, the average HRT for the whole experiment (106 d) was about 80 days.

3.2. Energy assessment

The consumption of energy for the period analysed reached a value (Q_{Total}) of 9,155 kJ/kg VS, which is equivalent to a specific power of 100 W/m^3. This consumption is slightly higher than that obtained by Walker et al. (2017) reporting a value of 75.1 W/m^3 for a digester with size 3.6 times smaller. In the present research, the thermal energy needed for maintaining the temperature of water tanks was obtained by the use of electric heaters (Q_{Man}). If this energy value was subtracted, so that both results became comparable, the value of specific power would be 68.8 W/m^3 which was slightly lower. The scale factor is an important parameter to be considered because the increase in reactor scale causes a decrease in energy demand. This is due to the fact that some energy consumption is associated with auxiliary equipment and remains approximately constant regardless of the size of the installation, such as the control unit. Others do not experience a linear increase proportional to the size of the plant, as it is the case of the heating system, because the surface of thermal leaks does not evolve proportionally to the volume of the reactor. The relationship between surface and volume decreases as the size of the digester increases.

The energy demand was analysed considering the different operating states of the prototype. The summation of categories in Fig. 5a accounts for the total energy demand of the prototype (9155 kJ/kg VS). The Heating state reports the highest energy consumption with a value of 58.4% of the total energy demand, followed by the Operation state that represents 35.9% (see Fig. 5). The energy consumption associated with the states of Recirculation, Feeding and Grinding was reduced, not exceeding 5.8% of the total consumption between the three operating states. This low energy demand was due to the fact that, although several actuators are involved in these states, the time needed for their activity is only 4.1%.

The plant is 73.3% of the time in the Operation state, for which the control unit monitors the process variables, controls the evacuation of biogas and records data. Although these tasks do not imply a high energy demand, the long-time associated with this state along with the use of thermal energy based on electric heaters (having high energy consumption) causes an exacerbated demand for energy. The availability of a different thermal source would reduce the energy needs of this operating state from 3,283 kJ/kg VS to 401 kJ/kg VS, which was the value associated with the control unit. This reduction would translate into the global context of the plant in 87.8% reduction in energy demand.

Within the heating state, the energy demand for producing hot water accounts for 33.5% (3,065 kJ/kg VS) of the total energy needs, whereas the pumping of digestate through the heating exchange system accounts for 22.3% (2,041 kJ/kg VS). These results show the high amount of energy necessary for keeping the temperature of the digester. The process was set to be in the Recirculation state 2.8% of the time (10 min ON and 5h and 50 min OFF) (Meroney and Colorado, 2009), which was equivalent to a renewal rate of 1.37 renewals/d, and a time of renovation equivalent to 17.5 h.

Fig. 5a shows that the system had a recirculation rate apparently lower than the set value. This is explained by the fact that heating and recirculation tasks were performed simultaneously but the system gives priority to heating over recirculation. In this prototype heating also implies recirculating. If the periods associated with heating state are to be considered as recirculating time, an effective recirculation regime is obtained with an average renewal time very close to 2 h, exceeding the values of 4 h used by Cabbai et al. (2016). Despite this, a lower renewal rate could be used since high recirculation regimes do not show remarkable improvements in the amount of biogas produced (Lindmark et al., 2014), contributing to decreasing energy needs of the prototype.

Figure 5 here

Regarding the energy consumption by the different devices, it should be noted that the electric water heaters are responsible for 66.6% of the energy consumed. This is mainly due to the energy associated with the Operation state and the high amount of energy required for maintaining process temperature, which takes place during the Heating state. The energy requirement of the main pump accounts for 24.0% of total devices. Most of this demand

(22.3%) was associated with the state of Heating. It should also be noted, that the energy associated with the main pump during the Heating state also has an additional effect on the demand of the recirculation. Energy consumption of the rest of the actuators was considered negligible compared to the total energy needs.

The consumption of energy for electrical uses ($Q_{C.el}$) accounts for 3,061 kJ/kg VS which was equivalent to an average power of 31.88 W/m^3. This result is considerably lower than that obtained by Walker et al. (2017) with a value of 75.1 W/m^3. The scale factor is obviously behind the lower demand of the present prototype. These authors used a digester size 3.6 times smaller than the present one. 54% of its consumption was due to data recording, which can be considered to have an energy demand approximately constant regardless of the size of the plant. The thermal energy demand to keep process temperature (Q_{Proc}) corresponds to a power of 31.90 W/m^3 (3065 kJ/kg VS), with this value being also less than the obtained by Walker et al. (2017), due to the scale factor previously discussed, 40 W/m^3.
Q_{Proc} can be divided in two categories, one is the energy necessary for heating the feed (Q_{Feed}) and the other is the heat associated with intrinsic losses (Q_{Losses}). The plant demanded 1,934 kJ/kg VS for heating the feed to the process temperature and had thermal losses of 1,131 kJ/kg VS associated with inefficiencies of insulation, representing 36.9% of the thermal demand of the unit.

The biogas yield of the prototype was 0.56 m^3/kg VS with an average methane richness of 64.3% (equivalent to an energy value (Q_{Biogas}) of 12,864 kJ/kg VS). If the valorisation of biogas is assumed by means of a CHP unit (conditions stated in scenario B), an electrical energy generation (EG_{elec}) of 2,894 kJ/kg VS and thermal energy ($EG_{thermal}$) of 8,104 kJ/kg VS would be expected. Taking into account that 3,065 kJ/kg VS are needed for meeting the thermal demand of the plant, then 5,039 kJ/kg VS would be available for other applications. Energy flows are represented by a Sankey diagram shown in Fig. 6.

Figure 6 here

For the calculated energy flows, a COP_{th} value of 2.64 and a COP_{el} value of 0.95 were obtained, thus in the case of electricity, the micro-plant produces less energy than that needed for its operation when an efficiency coefficient of 22.5% was assumed. However, when results are recalculated considering the use of a CHP system with efficiencies of η_{el} = 25% and η_{th} = 50%, for electricity and heat respectively, the new coefficients are COP_{th} = 2.09 and a COP_{el} = 1.05. Comparing these values with those obtained by Walker et al. (2017) (COP_{th} = 5.55 and COP_{el} = 1.47) the prototype reported in this study gives a lower performance explained by the 33.8% lower OLR of operation.

The assessment of the reactor was also performed considering a theoretical increase in OLR to a value of 2.7 g VS/L d. Results for this scenario are shown in blue colour in Fig. 8 affecting some of the energy quantities estimated. These results considered the same methane yield. The increase in OLR caused a proportional increase in the electricity produced and thermal energy available. Regarding the demand for thermal energy, a decrease in the theoretical operating point relative to the one tested was observed, from 3,065 to 2,407 kJ/kg VS. This decrease is explained by the fact that thermal losses of the digester remain constant regardless of the OLR applied. The demand of electrical energy has a similar trend, resulting in a decrease when evaluating the theoretical operating point, from 3,061 (experimental value) to 1,725 kJ/kg VS. In this case, the decrease is greater because the consumption of the different electric actuators remains about the same in spite of the increase in OLR, but for the main pump and heating pump which have higher energy demands. Values of COP_{th} and COP_{el} for this theoretical point were also calculated. The COP_{th} is increased from 2.64 to 3.37, because heat losses remain constant causing a better thermal efficiency per unit of volatile solid. The COP_{el} has a behaviour similar to that of COP_{th}, due to the same reason, in this case it rises from 0.95 to 1.68, because much of the electricity demand of the plant presents minimum variations regarding the OLR applied.

4. Conclusions

The assessment of the prototype was successfully carried out obtaining a methane yield of 360 L/kg VS at an OLR of 1.06 g VS/L d (calculated using data from days 55 to 90 of operation) in spite of the high concentrations of ammoniacal nitrogen (5100 mg/L). Most of the thermal energy requirements were associated with the raise of temperature of the feed to process conditions. The efficiency of the heating system was crucial, since a large amount of both electrical and thermal energy was necessary for operation, accounting for more than 80% of the total energy demand of the installation.

The energy assessment of the process was carried out assuming the incorporation of a combined heat and power (CHP) unit for valorising biogas and avoiding the use of electrical heaters for the supply of heat. The Sankey diagram showed that the process was thermally sustainable, since only 37.8% of the useful thermal energy generated by the CHP system would be used to meet the heat demand of the reactor. However, it should be pointed out that 36.9% of the energy used in heating was lost due to thermal losses associated with the reactor external surface and piping along with inefficiencies of the heating exchange system. The COP_{el} parameter was slightly lower than the unit indicating that the micro-plant was close to reaching self-sufficiency. The OLR of the process was decisive for the overall performance directly influencing the energy consumption per unit of volatile solid and therefore the energy efficiency indicators. The evaluation of operating parameters when estimated at an OLR of 2.7 g VS/ L d would result in values of 3.37 for the COP_{th} and 1.68 for the COP_{el}

Acknowledgments

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Supplementary Material Caption

ESM. Details of Micro plant description and operation. Parameters recorded during the operation of the plant. Operating time of the different equipment constituting the micro-plant

Graphical Abstract

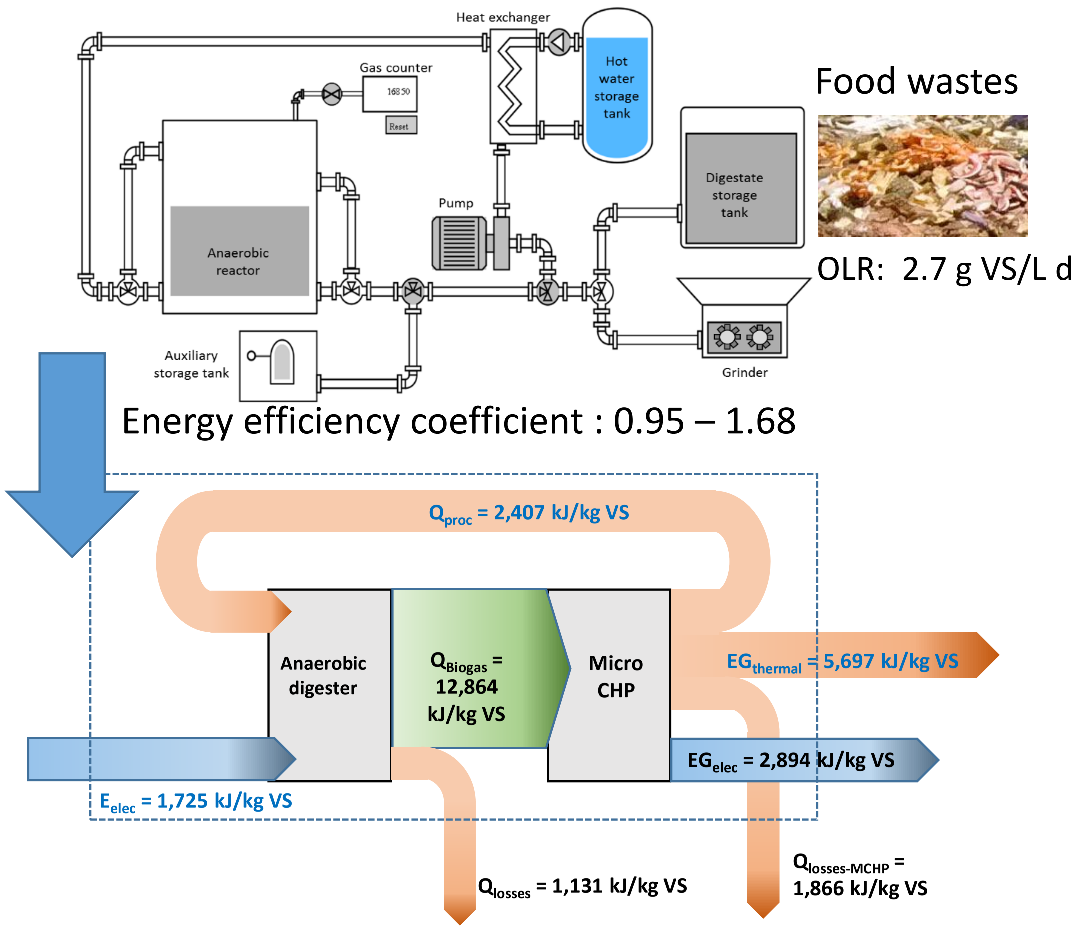

Table 1. Chemical characterisation of food wastes obtained from the cattering school of Toledo

Parameter	Value
Total solids (%)[1]	20.6 ± 0.8
Volatile solids (%)[1]	16.0 ± 1.0
Organic matter (%)[2]	59.2 ± 4.6
Moisture (%)	79.4 ± 2.3
Ash (%)[2]	22.3 ± 1.2
TN (%)[2]	1.80 ± 0.13
C/N ratio	18.7 ± 1.9
pH	5.70 ± 0.2
Conductivity (μS/cm)[1]	285.3 ± 3.5
$N\text{-}NH_3$ (ppm)[1]	2.71 ± 0.11
$N\text{-}NO^{3-}$ (ppm)[2]	1.48 ± 0.10
Total P (ppm)[2]	2.64 ± 0.13
Ca (ppm)[2]	57.21 ± 0.31
Mg (ppm)[2]	2.38 ± 0.14
K (ppm)[2]	23.35 ± 0.10

Na (ppm)[2]	1.03 ± 0.06
Mn (ppm)[2]	71.90 ± 2.11
Fe (ppm)[2]	1.84 ± 0.11
Cu (ppm)[2]	11.40 ± 0.45
Zn (ppm)[2]	1.48 ± 0.07

[1] Wet basis, [2] Dry basis, TN: Total nitrogen

Table 2

Table 2. Parameters of biological performance obtained from the digestion test for the evaluation period from day 55 to day 90

Parameter	Value
Substrate, TS (%)	27.5 ± 3.7
Substrate, VS (%),	23.1 ± 4.5
Biogas production (L/d)	2.746 ± 469
Methane volumetric production (mL CH_4/L d)	378 ± 65
Methane yield (mL CH_4/g VS_{added})	360 ± 67
Methane concentration (%)	64.3 ± 0.6
OLR (g VS/L d)	1.06 ± 0.15
HRT (d)	55.3 ± 11.0
TS (%)	3.11 ± 0.62
VS (%)	1.53 ± 0.41
VS removal (%)	93.1 ± 1.2

OLR: Organic loading rate, HRT: Hydraulic retention time.

Figure 1

Figure 1. a) Selection and transport of food wastes to the pre-treatment unit for grinding. b) Main components of the micro-plant for decentralised digestion of food wastes

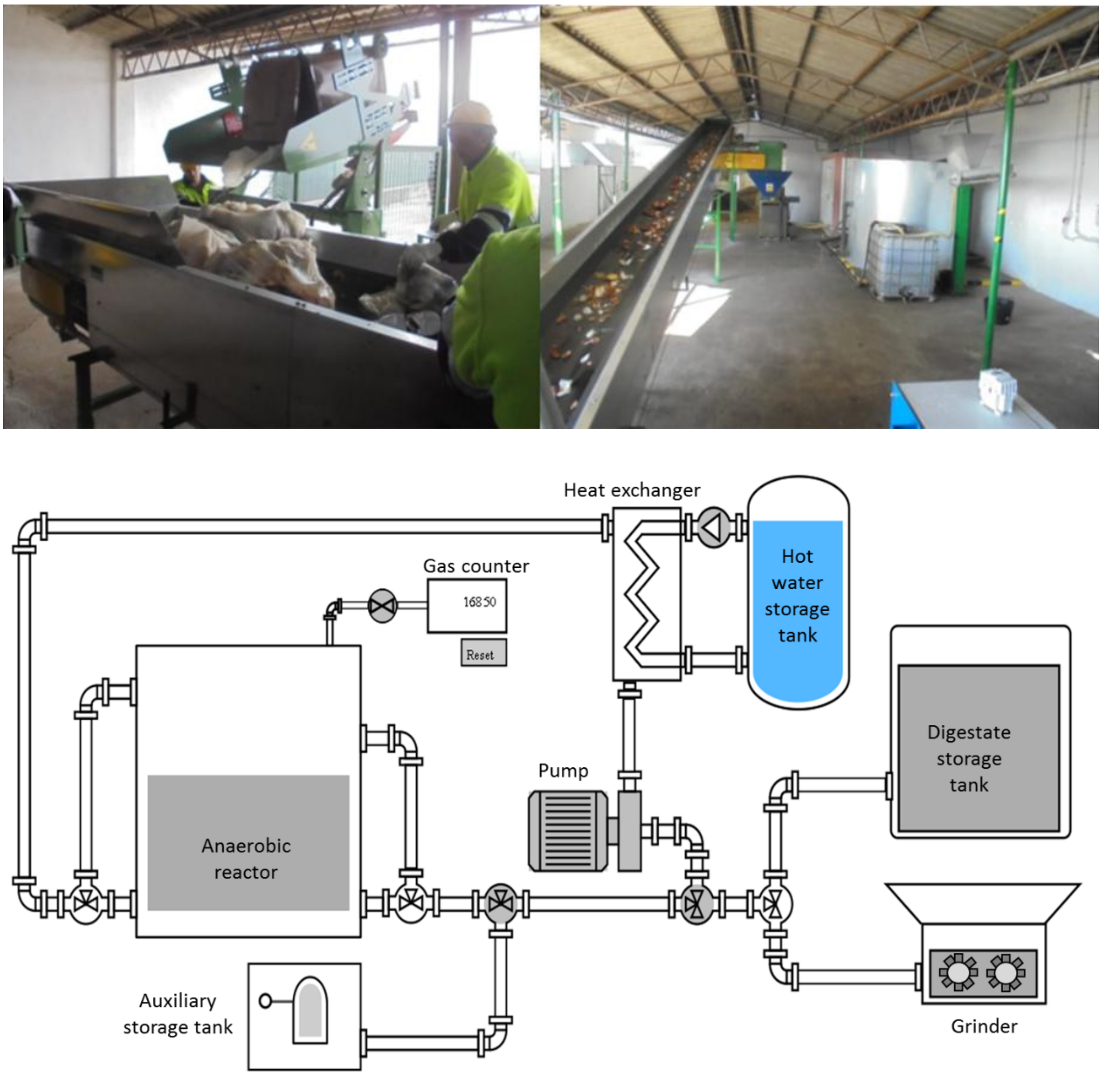

Figure 2

Figure 2. Mass flow evolution in the pre-treatment tank for the 106-day evaluation period

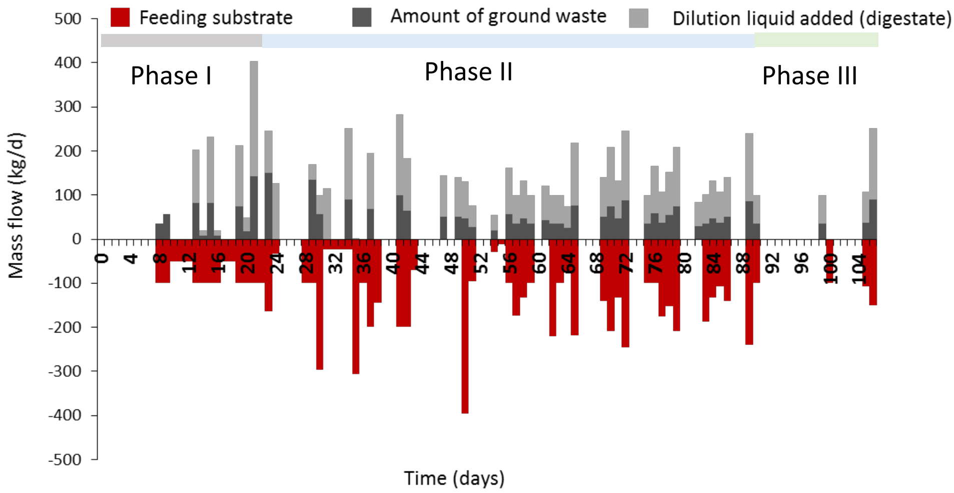

Figure 3

Figure 3. Mass balances of the AD system a) Mass flow of the different streams expressed in kg/m^3 d for the digester unit, b) and for the experimental period between days 55 and 90. VS balances c) organic loading of the different streams expressed in g VS/m^3 d, d) and for the experimental period between days 55 and 90

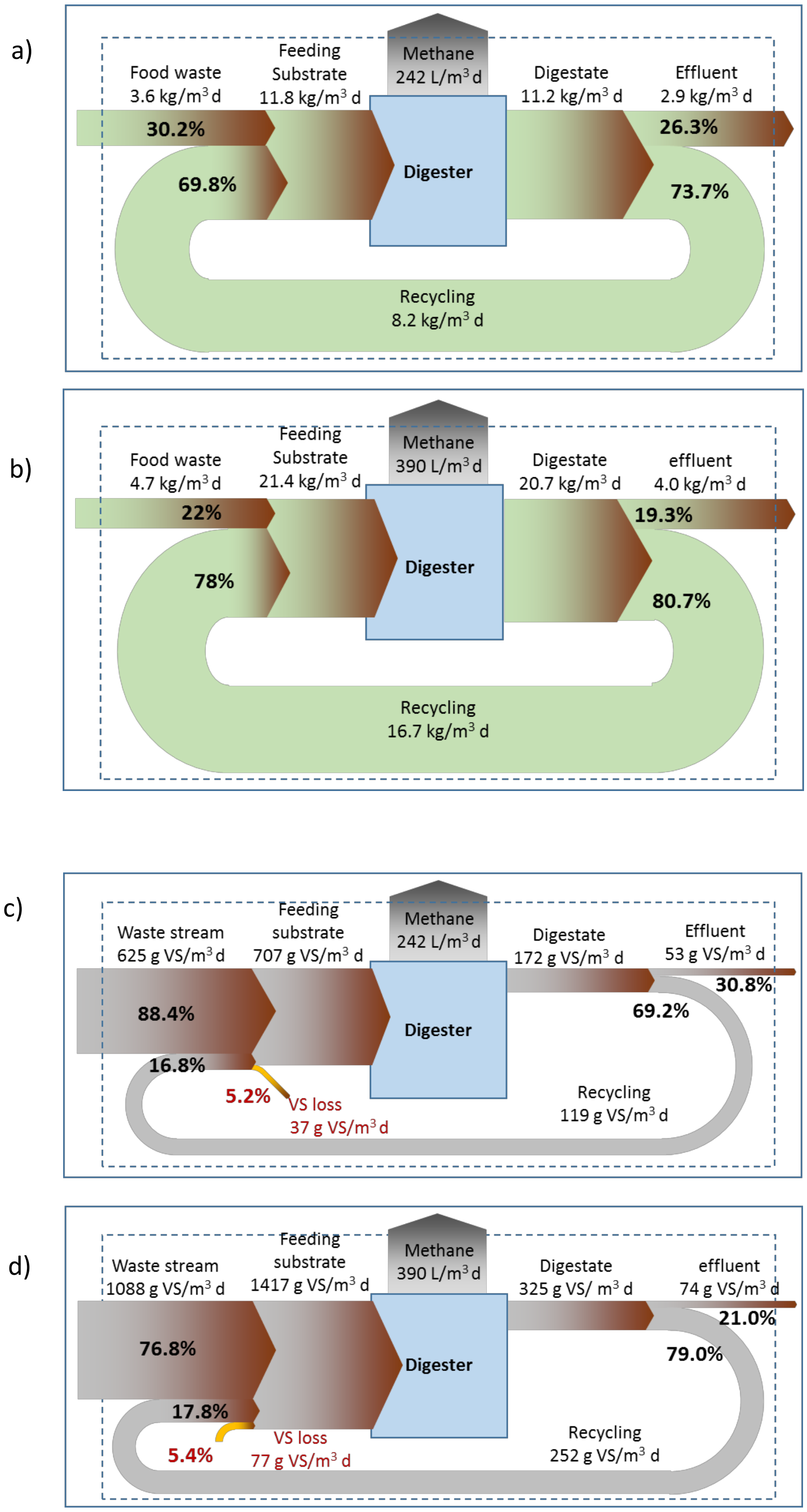

Figure 4

Figure 4. Chemical parameters obtained from digester performance: a) Daily biogas production, b) TS and VS content expressed as percentage of TS, c) pH, ammonia and free ammonia concentration, d) Volatile fatty acids, e) alkalinity represented as total alkalinity (TA), partial alkalinity (PA) and alkalinity ratio (AR).

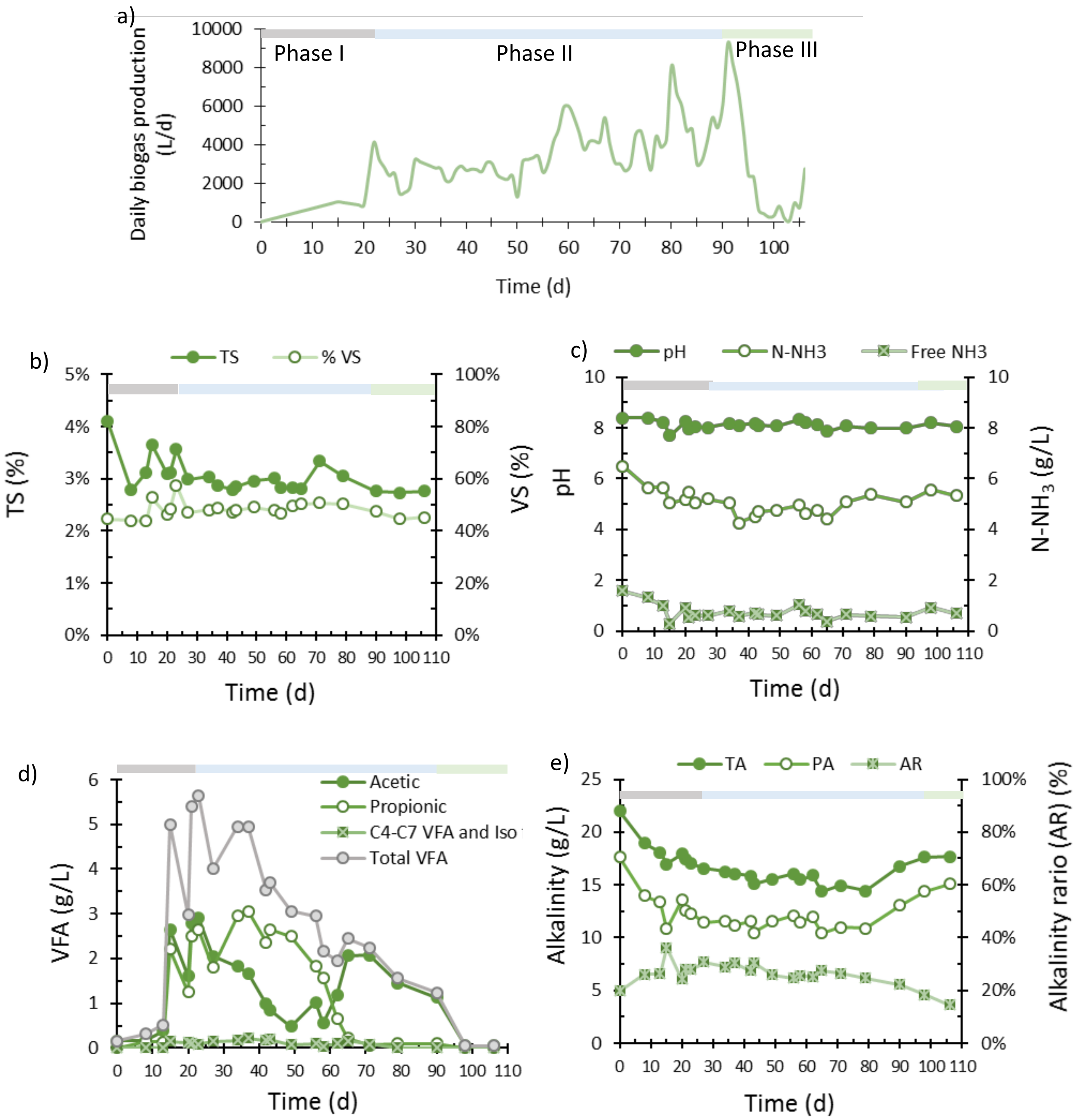

Figure 5

Figure 5. Pie chart representing energy demand of the prototype for the different operating states and circular graphics representing the energy consumption by devices in each state during the period 55 – 90 days. Values are expressed per unit of VS treated

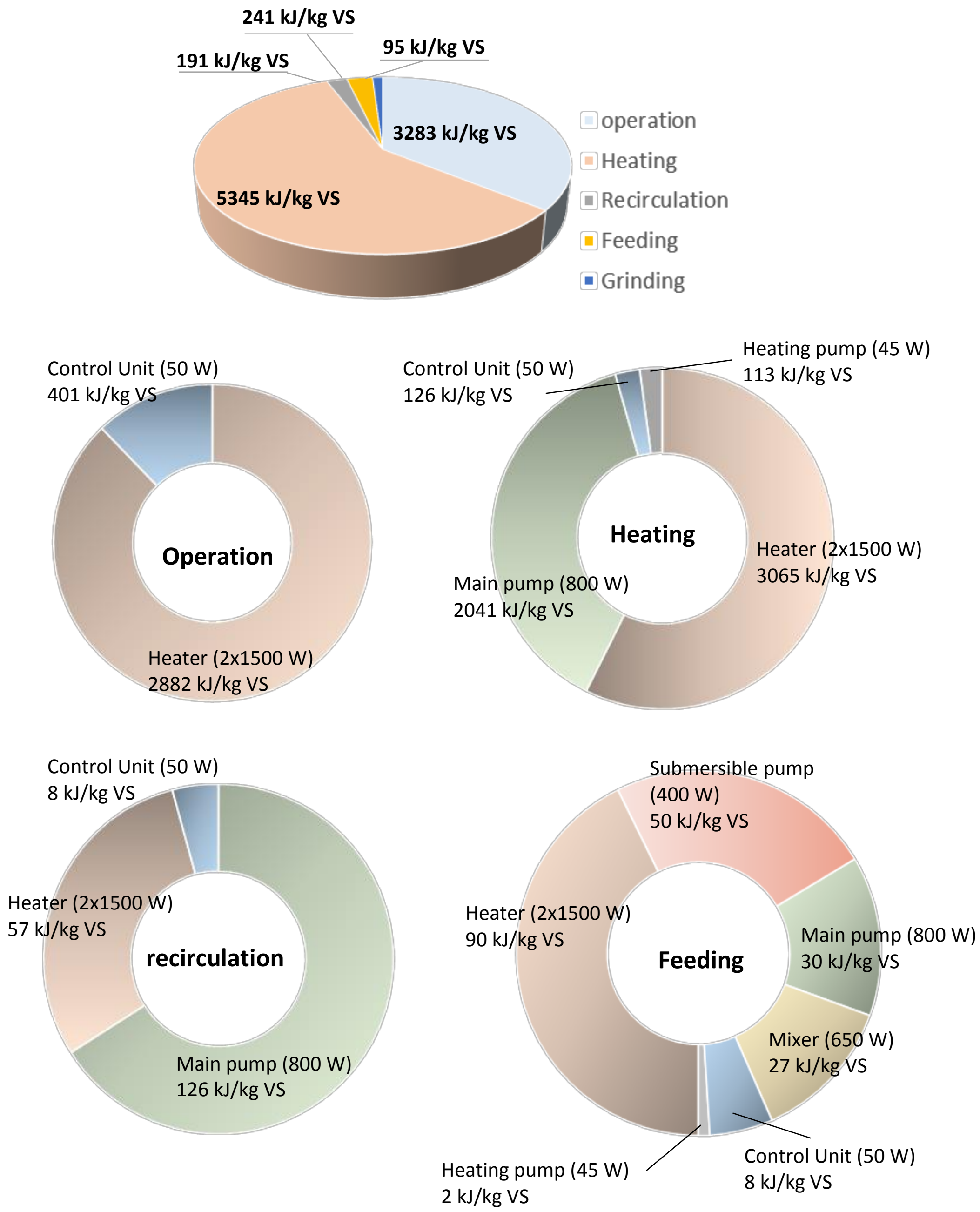

Figure 6

Figure 6. Sankey diagram representing energy flow of the pilot-plant (Scenario B) for the treatment of food waste using an anaerobic reactor and a pre-treatment unit based on the solubilisation of particulate material by the addition of digestate. Tested conditions at an OLR of 1.06 g VS/L d. Data reported in blue correspond to results obtained when the hypothetical OLR applied is 2.7 g VS/L d.