Chemical Engineering Journal 255 (2014) 71–77

Contents lists available at ScienceDirect

Chemical Engineering Journal

journal homepage: www.elsevier.com/locate/cej

Importance of scale and hydrodynamics for modeling anaerobic digester performance

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HIGHLIGHTS

- A simplified anaerobic digestion model was developed.
- The performance of the reactor at lab-scale and pilot-scale was predicted.
- The influence of mixing was investigated.
- Unmixed conditions resulted at pilot-scale in reduced biogas production and VFA accumulation.

ARTICLE INFO

Article history:
Received 13 February 2014
Received in revised form 4 June 2014
Accepted 10 June 2014
Available online 18 June 2014

Keywords:
Anaerobic digestion
Modeling
Mixing behavior
Biogas production
Methane

ABSTRACT

In this study it is demonstrated that a correct description of the mixing behavior of an anaerobic digester is necessary to accurately model and predict the reactor performance, especially at a larger scale operation. A lab-scale (3.78 l) and a pilot-scale (120 l) anaerobic digester were operated at both mixing and non-mixing conditions. At lab-scale no significant difference in performance was found between these conditions. In each case about 0.25 $lCH_4/l_{reactor}/d$ of biogas was produced. The model predictions were in close agreement with measured values. At pilot-scale, however, the influence of mixing became important. A reduction of about 10% of methane production occurred during unmixed conditions. Furthermore also a build-up of volatile fatty acids (up to 3 gCOD/l) prevailed. This behavior could not be predicted by using a CSTR approach. In contrast, it was accurately predicted with the proposed model, which includes a more rigorous mixing next to common reaction kinetics. Prediction of the daily biogas production assuming a perfectly mixed reactor revealed an overestimation of the biogas production by 10%. It is therefore recommended to use more accurate mixing models in order to predict biogas production from anaerobic digesters.

1. Introduction

The cost of sludge handling, processing and disposal makes up a large proportion of wastewater treatment plant (WWTP) operational cost. By applying digestion, energy can be recovered and the remaining sludge volume can be reduced thereby bringing further financial savings. Digested sludge is also more suitable for fertilizer use in agriculture and reduces the odour problems during transportation and field application.

Nowadays, mathematical models are being used extensively for optimizing and evaluating digester performance at WWTPs. Modeling has been applied in all digester process variations, most often with completely stirred tank reactor (CSTR) and upflow anaerobic sludge blanket (UASB) process designs. In most cases the Anaerobic Digestion Model No. 1 (ADM1) proposed by the IWA task group for Anaerobic Digestion Model development [1,2] was used. Next to this, other, either simpler or more elaborate models exist [3,4]. Ramirez et al. [5] presented a case where ADM1 is modified to take into account microbial diversity maintaining the same, original ADM1 processes. Esposito et al. [6] extended the ADM1 model for co-digestion of sewage sludge and the organic fraction of municipal solid waste (OFMSW). ADM1 has also been complemented with a biofilm model and used in CSTR performance predictions [7]. Borja et al. [8] used a simplified kinetic model for studying

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http://dx.doi.org/10.1016/j.cej.2014.06.041
1385-8947/

the anaerobic digestion (AD) process of olive pomace, while Keshtkar et al. [9] used a kinetic model for the dynamic simulation of the anaerobic digestion of cattle manure.

Most anaerobic digestion models use the presumption that the component distribution in the digester is homogeneous due to stirring or hydraulic mixing. In a recent review by Lauwers et al. [4] this mixing aspect is ignored. Anaerobic models can however be extended to account for mixing and diffusion in biofilms. Batstone et al. [10] modeled the performance of the granules of an UASB reactor with a distributed parameter biofilm model with the ADM1 model. Mu et al. [11] developed a model for a UASB-reactor where axial dispersion in the reactor is taken into account. Finally, Pontes and Pinto [12] demonstrated the use of compartmental models to properly simulate different digester sections: the sludge bed, the blanket and the settler.

Simplification of the mixing behavior can further lead to the necessity to calibrate the kinetic model in order to fit the experimental data while in fact the kinetic model is correct and the deviation from the experiments is related to the incorrect description of the mixing behavior of the reactor. Delrue et al. [13], for example, modeled a full-scale membrane bioreactor (1600 $m^3\ d^{-1}$) based on the ASM1 model [14] and a significantly larger value for the half-saturation coefficient for nitrate was obtained, caused by the non-ideal mixing behavior [15].

This study elaborates further on the importance of accurate description of reactor hydrodynamics, even with simplified (kinetic) models. As such, mixing at lab-scale and pilot-scale anaerobic digesters is compared. Furthermore, the aim is to demonstrate that a more rigorous description of reactor mixing behavior leads to more accurate predictions of the system performance, especially at a larger scale operation, without the need for significant calibration of the model parameters.

2. Materials and methods

2.1. Experimental data collection

The experimental data used in this study was gathered in previous studies [16,17]. More in particular data was used from a lab-scale experiment that lasted 84 days and a pilot-scale experiment of 154 days. As there were some experimental problems for the first 44 days of the lab-scale experiment, only data of the last 40 days of that particular experiment was used.

A lab-scale digester (15.2 cm in diameter) with an overall volume of 5 l was used as well as a pilot-scale digester (45.7 cm in diameter) with an overall volume of 127 l. Both digesters were geometrically similar. The working volume of the lab-scale unit was 3.78 l, whereas for pilot-scale it was 96 l. The digesters were operated inside a temperature-controlled (35 °C) cabinet.

The digesters were operated in both mixed and unmixed conditions. To achieve mixing by recirculation of biogas, the digesters were equipped with a draft tube and a sparger. The biogas recirculation rate was 1 L/min in the lab-scale reactor and 9 L/min in the pilot-scale digester. This resulted in gas hold-up of respectively 3.78 min and 10.67 min. As such, the gas hold-up was similar, although somewhat larger for the pilot-scale reactor. For operation under unmixed conditions, there was no recirculation of gas at all. The only potential mixing that occurred under unmixed conditions was due to reactor feeding and evolution of biogas.

The digesters were operated using cow manure collected from a dairy farm. The cow manure was obtained fresh (less than 7 days old) from cows kept in a pasture (i.e. grass-fed) under no antibiotic treatment (this limiting the viability of methane generating microorganisms in the cow manure). The cow manure was collected once at the beginning of the experiment and was stored at 4 °C until use.

Before feeding, the wet manure was blended with tap water (in 1:3 ratio, to adjust total volatile solids content) for 2 min with an impeller mixer and placed into a large bucket for the heavy solids (sand, etc.) to settle out. Then the slurry was passed through a sieve with 9.5 mm openings. Finally, the slurry was diluted with water to obtain 6.6% total volatile solids concentration (total solids of about 12–13% with very low sand content). The feeding rate (or effluent removal rate) was adjusted to maintain a hydraulic retention time of 16 days. The hydraulic residence times applied are in line with hydraulic residence times used in literature (see e.g. [18]). A known amount of reactor content (470 mL for lab-scale and 12 l for pilot-scale) was removed every 2 days and triplicate samples were collected for total solids and total volatile solids determination. At the same time, 470 mL or 12 l of feed slurry was added to the top of the lab-scale and pilot-scale digesters, respectively.

Every 2 days, gas composition and cumulative gas production volume were determined. The cumulative gas volume was determined by a wet gas test meter. Gas samples were collected using a gas-tight syringe from a sampling port in the gas production line to analyze the methane and carbon dioxide content. Effluent samples were analyzed for volatile suspended solids (VSS) and volatile fatty acids (VFA).

VSS were determined by volatilization of a known weight of dried slurry at 540 °C for a minimum of 15–20 min until constant weight. VFA was determined using High Performance Liquid Chromatography (HPLC).

2.2. Kinetic model

A simplified kinetic model for the anaerobic digestion of manure was formulated, inspired by the models proposed by e.g. Zaher et al. [19], Keshtkar et al. [9], Garcia-Ochoa et al. [20] and Borja et al. [8]. The base unit for the model was Chemical Oxygen Demand (COD), similar to ADM1 [10,21,22]. For the kinetic model it was assumed that the insoluble organic matter or volatile suspended solids (VSS) is first transformed to soluble organic matter or volatile dissolved solids (VDS) respectively according to first-order kinetics. The VDS is transformed by acetogenic bacteria (X_1) to volatile (fatty) acids (VFA) following Monod kinetics. Subsequently, the resulting VFAs are transformed by methanogenic bacteria (X_2) to methane again following Monod kinetics. In contrast to Borja et al. [8], the biomass concentration was not assumed to be practically constant and as such biomass growth and decay were explicitly incorporated into the model. For both types of bacteria the yield was assumed to be equal to 0.05 gCOD/gCOD [10], meaning that for every gram COD of substrate removed, 0.05 gCOD of biomass is produced. The resulting Gujer Matrix summarizing processes, components, process rates and stoichiometry is presented in Table 1.

The parameter values for the kinetic model were derived from literature and are presented in Table 2. The main difference with the ADM1 model default parameters [10] was based on Lübken et al. [21], studying the digestion of cattle manure and concluding that for accurate model simulations the first order hydrolysis constant (k_1) needed to be decreased significantly. Indeed, hydrolysis is generally considered the rate-limiting step during the AD of particulate organic matter [23]. Furthermore, this hydrolysis constant is also indicated as the most sensitive parameter to determine when complex substrates are used [24,25].

Kinetic constants for this hydrolysis of particulate organic matter are typically in the order of 0.003–0.6 d^{-1} [19,21,26–28]. As in this study a similar model as presented by Zaher et al. [19] ($k_1 = 0.0036\ d^{-1}$) and Borja et al. [8] ($k_1 = 0.054\ d^{-1}$) was used, also a similar hydrolysis rate constant was selected ($k_1 = 0.02\ d^{-1}$). The kinetic constants k_2 and k_4 were assumed to be equal to the ADM1