Elsevier Editorial System(tm) for Chemical Engineering Research and Design
Manuscript Draft

Manuscript Number: CHERD-D-17-00990R2

Title: Optimal integrated facility for waste processing

Article Type: SI: Energy Systems Eng

Keywords: Biogas; Digestate; Anaerobic digestion; Manure; Power production; Mathematical optimization

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Abstract: A mixed-integer nonlinear programming strategy is proposed to design integrated facilities to simultaneously recover power and nutrients from organic waste. The facilities consider anaerobic digestion of different types of manure (cattle, pig, poultry, and sheep). The products from this step are biogas and a nutrient-rich effluent. The biogas produced is cleaned and used in a gas turbine to produce power while the hot flue gas obtained from combustion produces steam that is fed to a steam turbine to produce additional power. The nutrient-rich effluent is processed to recover the nutrients using different technologies that include filtration, coagulation, centrifugation, and struvite precipitation in stirred and fluidized bed reactors. This processing step provides a mechanism to prevent phosphorus or nitrogen release to the environment and to avoid the development of eutrophication processes. It is found that struvite production in fluidized beds is the technology of choice to recover nutrients from all manure sources. Furthermore, power production depends strongly on manure composition and exhibits high cost variability (from 4000 €/kW in the case of poultry manure to 25000 €/kW in the case of cattle and pig manure).

Prof. M Giorgiadis
Chem. Eng. Res. Des.

Salamanca, 28th November 2017

Dear Editors,

As per invitation, we are pleased to submit the revised version of the manuscript entitled "**Optimal integrated facility for waste processing**" where we evaluate the use of various types of waste towards the production of power and the recovery of nutrients out of the digestate. Following the suggestion of the reviewers we have added the model to the text and the computing time to solve the model. The facility compares the use of 5 different technologies for the recovery of phosphorus and nitrogen, from simple filtration to the production of struvite, and the yield to power of cattle, pig, sheep and poultry manure. In all cases struvite production using a fluidized bed reactor is selected as the most efficient technology for processing the digestate while poultry is the manure with the largest power production.

Sincerely Yours

Mariano

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Optimal integrated facility for waste processing

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We greatly appreciate the comments of the reviewers to improve the quality and the clarity of the manuscript. Please find in **black** the reviewers' comments, in **red** the answers to the points raised by the reviewer and in **blue** the new additions to the paper.

Reviewer #2: The authors have responded to my requests and comments in a satisfactory way. The manuscript may be accepted for publication.

We appreciate the support of the reviewer to the revised version.

Reviewer #3: The revised paper has addressed most of the previous comments. However, it is still not ready for publication. Authors should further improve the structure of the paper to provide better flow of the proposed idea. Based on the current arrangement, readers will be difficult to understand the proposed scope fully.

Besides, authors should also address the following issues.

- Literature review of process synthesis and optimization for bioenergy system should be provided. Highlight the novelty of the proposed framework.

We have provided a mathematical framework for the evaluation of nutrient recovery technologies that is nonexistent in the literature. Most of the work is experimental and the comparisons between technologies scarce

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- Provide the list of component in a table form.

Following the suggestion of the reviewers

Components set							
Number of component	**Component**	**Number of component**	**Component**	**Number of component**	**Component**	**Number of component**	**Component**
1	Wa	**12**	O	**23**	K2O	**34**	Cl
2	CO2	**13**	N	**24**	CaCO3	**35**	Struvite
3	CO	**14**	Norg	**25**	FeCl3	**36**	KStruvite
4	O2	**15**	P	**26**	Antifoam	**37**	MgCl2_CSTR
5	N2	**16**	K	**27**	Fe2SO4_3	**38**	NaOH_CSTR
6	H2S	**17**	S	**28**	Al2SO4_3	**39**	Mg_CSTR
7	NH3	**18**	Rest	**29**	AlCl3	**40**	Cl_CSTR
8	CH4	**19**	Cattle_slurry	**30**	MgCl2	**41**	Struvite_CSTR
9	SO2	**20**	Pig_slurry	**31**	NaOH	**42**	KStruvite_CSTR
10	C	**21**	Poultry_slurry	**32**	Struvite_seeds	**43**	FeCl3_Coag
11	H	**22**	P2O5	**33**	Mg		

- Equations should be provided along with the discussion in Section 3. This will allow the readers to understand the modelling better.

Actually we removed them because the reviewers said it was too long. We preferred them in the text of course. We have added them again

- Reaction equations should also be numbered.

Following the suggestion of the reviewer we have number all, equations and reactions.

- What is the computational time and model size for the case study?

Following the suggestion of the reviewer we have added to section 3

The superstructure consists of an NLP of approximately 4000 equations and 5000 variables solved using a multistart procedure with CONOPT 3.0 as the preferred solver. The computational time is around 60 min, although it varies for each problem as a consequence of the different data used in each case.

- Is the process selection only focusing on recovery of N and P? How about the selection technology for production and purification of biogas, power generation via biogas?

This work focuses on nutrient recovery since we have already addressed optimal power generation in a previous paper. Therefore, building on those results we focus on N and P recovery.

León, E., Martín, M., 2016. Optimal production of power in a combined cycle from manure based biogas. Energy Conv. Manag. 114, 89-99.

- How to generate the presented results?

We solve the optimization model as presented in section 3.

Highlights

-Cattle, pig, poultry and sheep manure are evaluated for power and digestate processing

-5 technologies are compared for the recovery of P and N

-FBR technology producing struvite is the selected technology

-Poultry shows larger yield to biogas, but smaller biomass availability

-Power production is competitive as long as nutrients are sold as fertilizers.

*Manuscript
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Optimal integrated facility for waste processing

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Abstract.

A mixed-integer nonlinear programming strategy is proposed to design integrated facilities to simultaneously recover power and nutrients from organic waste. The facilities consider anaerobic digestion of different types of manure (cattle, pig, poultry, and sheep). The products from this step are biogas and a nutrient-rich effluent. The biogas produced is cleaned and used in a gas turbine to produce power while the hot flue gas obtained from combustion produces steam that is fed to a steam turbine to produce additional power. The nutrient-rich effluent is processed to recover the nutrients using different technologies that include filtration, coagulation, centrifugation, and struvite precipitation in stirred and fluidized bed reactors. This processing step provides a mechanism to prevent phosphorus or nitrogen release to the environment and to avoid the development of eutrophication processes. It is found that struvite production in fluidized beds is the technology of choice to recover nutrients from all manure sources. Furthermore, power production depends strongly on manure composition and exhibits high cost variability (from 4000 €/kW in the case of poultry manure to 25000 €/kW in the case of cattle and pig manure).

Keywords: Biogas, Digestate, Anaerobic digestion, Manure, Power production, Mathematical optimization

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1.- Introduction

Countries across the globe generate large amounts of organic waste that include urban residues and sludge and manure from livestock activities. While many of these waste streams can be used as a source for power and chemical products, identifying suitable cost-effective technologies is challenging. Anaerobic digestion (AD) is a promising technology to treat these residues to produce biogas, which can be used as a source for thermal energy and electrical power (León and Martín, 2016) or chemicals (Hernández and Martín, 2016). However, AD technologies also generate a nutrient-rich residual stream called digestate, that must be further processed to prevent waste and soil contamination. In particular, nutrient management is needed to prevent losses of phosphorous and nitrogen to surface and underground water bodies which leads to eutrophication processes (Sampat et al., 2017, Garcia Serrano et al., 2009). There are a number of technologies that can be used to process the digestate that range from simple mechanical separations such as filters (Gustafsson et al., 2008) and centrifugation units (Meixner et al., 2015) to chemical processing such as struvite precipitation (Bhuiyan et al., 2008). Recent studies have analyzed the production of highly concentrated nutrient products such as struvite (Lin et al., 2015). The variability in the recovered product quality, selling price, and production cost present complex trade-offs for the optimal use of the digestate. Existing studies have only addressed the performance of various treatment mechanisms and lack a systematic design perspective that evaluates the performance of coupled biogas and nutrient recovery technologies (Drosg et al., 2015). This is necessary, for instance, to assess economic performance of nutrient recovery in the face of strong variations in the digestate content obtained from AD (Al Seadi et al., 2008).

In this work we propose a systematic design framework to optimize the simultaneous production of energy from the biogas obtained by anaerobic digestion of cattle, sheep, poultry and pig manure, along with the recovery of nitrogen and phosphorous from the digestate. The proposed framework determines the optimal technology configuration, equipment sizing, and operational conditions for various compositions of manure and digestate and revenues for biogas, electricity, and fertilizer.

The paper is organized as follows. In section 2 we present a brief description of the process and the flowsheet. In section 3 we focus on the modelling of the digestate processing technologies and costing. Section 4 presents the results for various feedstocks and section 5 draws conclusions.

2.- Process description

The proposed process consists of four sections: biogas production, biogas purification (biogas generation), electrical power generation, and nutrient recovery from digestate. This is illustrated in Figure 1.

The biomass together with water and nutrients (manure slurry) are fed to a bioreactor through stream 1, where the mixture is anaerobically digested to produce biogas and a decomposed substrate (digestate). The biogas, composed of methane, carbon dioxide, nitrogen, hydrogen sulfide, ammonia and moisture leaves the bioreactor through stream 2, and it is then sent to the purification section to remove H_2S in a fixed-bed reactor and to eliminate CO_2 and traces of NH_3 in a second step by using a Pressure Swing Adsorption (PSA) system. The purified biogas (stream 3) is used in a Brayton cycle, modelled as a furnace and an expansion, producing power. Air is fed via stream 4 and the exhaust gases (stream 7) are fed to a regenerative Rankine cycle, where it produces high pressure overheated steam extracted in stream 8. This overheated steam is fed to a steam turbine, where it is expanded to produce power. The exhaust steam from the turbine is recovered in stream 9 and reused in the Rankine cycle through stream 10. Between streams 9 and 10 hot flue gases from the gas turbine reheat and produced overheated steam from the recycled water (León and Martín, 2016).

The digestate is released from the digester through stream 12, and it can be processed through a number of technologies to remove nitrogen and phosphorous. We consider filtration, centrifugation, coagulation, and struvite production using either a fluidized bed reactor (FBR) or a continuous stirred tank reactor (CSTR). These technologies are described in detail in Section 3.4.

Four manure types have been considered as raw material for the process: cattle, pig, poultry and sheep manure. Table 1 shows the composition and properties of each type of manure.

Table 1: Manure composition and properties (Kowalski et., 2013; Lorimor et al., 2004, Al Seadi et al., 2008; Martins das Neves et al., 2009).

Figure 1.-Flowsheet for the production of power and fertilizers

3.- Modelling issues

We evaluate the performance of the different unit operations in the process by using detailed models that comprise mass and energy balances, thermodynamics, chemical and vapor-liquid equilibria, and product yield calculations. The global process model comprises total mass flows, component mass flows, component mass fractions, temperatures and pressures of the streams in the process network. The components that are considered in our calculations belong to the set shown in Table 2.

Table 2: Set of components

In the following subsections, we briefly present the main equations used to characterize the operation of the different units. For the sake of brevity, simpler balances based on removal efficiency or stoichiometry and equations connecting units are omitted. The power production system is described in detail in previous work (León and Martín, 2016) and we thus only provide a brief description.

The cost estimation for the alternatives and for the entire process is based on the estimation of the unit costs from different sources using the factorial method. From the units cost, the facility cost is estimated using the coefficients in Sinnot (1999), so that the total physical plant cost involving equipment erection, piping instrumentation, electrical, buildings, utilities, storages, site development, and ancillary buildings is 3.15 times the total equipment cost for processes which use fluids and solids. On the other hand, the fixed cost, which includes design and engineering, contractor's fees, and contingency items is determined as 1.4 times the total physical plant cost for the fluid and solid processes. In the subsequent cost estimation procedures these parameters are denoted as f_i for the total physical plant parameter and f_j for the fixed cost parameter.

3.1.- Biogas production

AD is a complex microbiological process that decomposes organic matter in the absence of oxygen. It produces a gas mixture following hydrolysis, acidogenesis, acetogenesis, and methanogenesis steps, consisting mainly of methane and carbon dioxide (biogas), and decomposed substrate (digestate). The anaerobic reactor is modeled using mass balances of the species involved in the production of biogas and digestate. Inorganic nitrogen, organic nitrogen, sulfur, carbon, and phosphorus balances are formulated by using the composition of volatile solids in manure, see Table 1 (Al Seadi et al., 2008; Martins das Neves et al., 2009). Typical bounds for the biogas composition are provided. The reactor operates at 55 ºC. We refer the reader to the supplementary material and León and Martín (2016) for details on the modelling of the digester.

3.2.- Biogas purification

This system consists of a number of stages to remove H_2S, CO_2 and NH_3. Here we highlight some basics about the operation of these stages. For further details we refer the reader to previous work (León and Martín, 2016).

The removal of H_2S is carried out in a bed of Fe_2O_3, that operates at 25-50 ºC producing Fe_2S_3. The regeneration of the bed uses oxygen to produce elemental sulfur and Fe_2O_3

CO_2 is adsorbed using a packed bed of zeolite 5A. The typical operating conditions for PSA systems are low temperature (25 ºC) and moderate pressure (4.5 bar). The recovery of the PSA system is assumed to be 100% for NH_3 and H_2O (because of their low total quantities in the biogas, in general), 95 % for CO_2, and 0% for all other gas of the mixture.

In both cases the system is modelled as two beds in parallel so that one bed is in adsorption mode while the second one is in regeneration mode, to allow for continuous operation of the plant. Further details can be found in the supplementary material.

3.3.- Electricity generation

We consider two stages for the generation of power. The initial one consists of the use of a gas turbine, a common alternative for using any gas fuel. However, the flue gas that exits the gas turbine is at high temperature. We can either produce steam as a utility or use that steam within a regenerative Rankine cycle to enhance the production of power. The details for the process appear in León and Martín (2016) or in the supplementary material.

3.3.1.- Brayton cycle

We model the Brayton cycle as a double-stage compression system (one for the air and one for the fuel) with intercooling with variable operating pressure for the gas turbine. The compression is assumed to be polytropic with a coefficient equal to 1.4 and an efficiency of 85% (Moran and Shapiro, 2003).

The combustion of methane from the biogas is assumed to be adiabatic, heating up the mixture. We consider the combustion chamber as an adiabatic furnace. We use an excess of 20% of air with respect to the stoichiometry and 100 % conversion of the reaction:

$$CH_4 + \ 2O_2 \rightarrow CO_2 + 2H_2O \tag{1}$$

The hot flue gas is expanded in the gas turbine to generate power and the expansion is assumed polytropic. In this case, a value of 1.3 is used based on an offline simulation using CHEMCAD®, with an efficiency of 85 % (Moran and Shapiro, 2003). Finally, the exhaust gas is cooled down and used to generate high-pressure steam to be fed to the Rankine cycle.

3.3.2.- Rankine cycle

We use the hot flue gas from the turbine to generate steam following a scheme that consists of using the hot gas in the order that follows. First, the hot flue gas is used for the superheating stage of the steam that is to be fed to the turbine. Next, the hot gas is used in the regenerative stage of the Rankine cycle, reheating the steam from the expansion of the high pressure turbine. Subsequently, the flue gas is used in the evaporation and preheating of the condensed water, see Figure 1. The details of the modelling of the Rankine cycle can be seen in Martín and Martín (2013). We assume an isentropic efficiency of 0.9 for each expansion.

3.4.- Digestate conditioning

Four different alternatives are considered to process the digestate including filtration, centrifugation, coagulation, and struvite production. For struvite production, the performance of fluidized bed reactors (FBR) and stirred tanks reactors (CSTR) systems is compared. For filtration, centrifugation, and coagulation technologies, nutrients output is a cake composed of different solids and nutrients, with a complex composition. The credit that we can get from the cake has been estimated based on the amount of nutrients contained. The prices for the nutrients (N, P and K) are assumed as follows: 0.45€/kg for N, 0.24€/kg for K and 0.32€/kg for P (Hernández et al., 2017).

3.4.1.-Filtration

Filtration is a low-cost technology that is appropriate for small installations where the amount of P to be removed is moderate. This technology consists of a filter that contains a reactive medium to help remove phosphorus. P removal using reactive filtration takes place through various mechanisms depending on the characteristics of the filter media. For instance, filter media made of compounds rich in cations under basic environments (usually containing calcium silicates at pH values above 9) form orthophosphate precipitates in the form of calcium phosphates, principally as hydroxyapatite (Pratt et al., 2012). Metallurgical slag captures P by adsorption over metal at pH close to 7 (Pratt et al., 2012). In this work we consider the use of five different types of filter media. Among them, we have studied wollastonite as a filter media rich in alkaline calcium silicates, dolomite Polonite® as calcium carbonate based components, and Filtra P® as calcium hydroxide based product (Österberg, 2012; Vohla et al., 2011). For the metallurgical slag, we have considered the blast furnace slag described by Cucarella et al. (2008). These filters are used in wastewater treatment facilities (Gustafson et al., 2008) and further analysis can be found in Shilton et al. (2006). Details on Ca-rich filters can be found in Koiv et al. (2010). The removal yield of P and N for the different filter media is shown in Table 3. It is possible to combine this filter medium with nitrogen-philic filters to simultaneously remove nitrogen and phosphorous. An advantage of this technology is that the cake produced can be used as soil fertilizer (Hylander et al., 2006). The removal yield of nitrogen for Filtra P has been considered similar to the limestone nitrogen removal yield, as Filtra P is a limestone derived product.

Table 3. Recovered P and N yield for different filter media

The model for the filtration is based on the removal efficiency per filter media, see Figure 2. It has been considered that materials such as total solids, carbon and potassium are forming solid compounds, so they will be retained by the filter media, Eqs. (4)-(7).

$$\begin{gathered} F_i^{cake} \geq F_i^{in} \cdot \eta_i^j - M \cdot (1 - y^j) \\ i \in \{P, N\} \\ j \in \{ filter\ media \} \end{gathered} \qquad (2)$$

$$\sum y^j = 1 \qquad (3)$$

$$F_i^{liquid\ effluent} = F_i^{in} - F_i^{cake} \tag{4}$$

$$F_k^{cake} = F_k^{in} \quad k \in \{TS, C, K\} \tag{5}$$

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} \right) \cdot \frac{C_{Wa}^{cake}}{1 - C_{Wa}^{cake}} \tag{6}$$

$$F_{Wa}^{liquid\ effluent} = F_{Wa}^{in} - F_{Wa}^{cake} \tag{7}$$

As we have considered five filter media, we have used a Big M formulation to select one of them assigning a binary variable $y^{filter\ media}$ for each filter media, Eqs. (2)-(3). This variable takes a value of 1 for the selected filter media and 0 for the rest, so that we are able to evaluate one filter media per time.

It is assumed that the cake obtained contains moisture with a value of 55% in weight basis (C_{Wa}^{cake}).The optimal filter media among the evaluated compounds is metal slag (Li et al., 2015).

Figure 2.- Scheme of the filter

The cost of each alternative has been estimated according to the number of filters, which depends on the maximum flow they can process. The maximum flow per filter unit, $F_{\max}^{filter}$, is 1300 ft³/min (Loh et al., 2002). To design the filter units we have taken the lower value between the flow provided by mass balances and the maximum flow allowed per filter, Eqs. (8)-(10).

$$F\left(\frac{ft^3}{min}\right) = \frac{F_{in}}{\rho_{digestate}} \tag{8}$$

$$n_{filters} \geq \frac{F_{total}^{filter}}{F_{max}^{filter}} \tag{9}$$

$$F_{design}^{filter}\left(\frac{ft^3}{\min}\right) = \min\left(F_{\max}^{filter}, F_{total}^{filter}\right) \tag{10}$$

In fact, since the maximum flow for a cartridge filter is 1300 ft³/min, for this facility the number of filters considered in this work will always be one and the design flow is equal to the flow provided by mass balances.

The correlation used to calculate the filter cost, Eq. (11), is obtained from data reported in Loh et al. (2002). This correlation provides the price in 1998 dollars, so we use the Chemical Engineering Index to update it.

$$FC_{filtration}(\$) = 4.7436 \cdot F_{design}^{filter} + 807.6923 \tag{11}$$

The operating cost is estimated using a simple correlation, Eq. (14), where we assume that the utilities contribute 20% of the total (Vian Ortuño, 1991). The other economical contributions considered are chemicals, estimated as in Eq. (12), labour, as per Eq. (13) and the contribution of the investment cost of the units given by Eqs. (10)-(11). The filter media are considered as chemicals that will be replaces annually.

$$ChemC_{filtration}\left(\frac{€}{year}\right) = \frac{F_P^{in} \cdot 3600 \cdot h \cdot d}{\dfrac{kg_P}{kg_{filter\ media}}} \cdot Price_{filter\ media} \tag{12}$$

In Eq. (12) $kg_{filter\ media}$ are calculated as the P content in the inlet stream divided by the filter media P adsorption capacity.

$$Labour\ cost\left(\frac{€}{year}\right) = \left(61.33 \cdot F_P^{recovered} \cdot 3.6 \cdot h^{(-0.82)}\right) \cdot \left(F_P^{recovered} \cdot 3.6 \cdot h \cdot d\right) \cdot \left(\frac{Salary}{h \cdot d}\right) \cdot n_{OP} \tag{13}$$

The number of operations considered, n_{OP}, is equal to 1.

$$Operating\ cost\left(\frac{€}{year}\right) = \frac{ChemC + 1.5 \cdot Labour\ cost + 0.3 \cdot FixedCost \cdot f_i \cdot f_j}{(1 - Utilities)} \tag{14}$$

Finally the credit obtained from the cake is computed as the weighted sum of each nutrient value, Eq.(15), (Hernandez et al., 2017), and the benefits (or losses) are computed as the difference between the credit obtained from the cake and the operating costs of the facility, Eq. (16).

$$Cost_{cake}\left(\frac{€}{year}\right) = \left(F_P^{recovered} \cdot Price_P + F_N^{recovered} \cdot Price_N + F_K^{recovered} \cdot Price_K\right) \cdot 3600 \cdot h \cdot d \tag{15}$$

$$Benefits_{Filtration}\left(\frac{€}{year}\right) = Cost_{cake} - Operating\ cost \tag{16}$$

3.4.2.-Coagulation

Coagulation is a chemical treatment to process the digestate. The goal of this process is to destabilize colloidal suspensions by reducing the attractive forces, followed by a flocculation process to form flocs from the previously destabilized colloids and to subsequently precipitate them. The nutrients are then recovered with other sedimented solids by clarification. Both N and P can be removed from the influent through coagulation-flocculation, where phosphorus is removed primarily in the form of metal hydroxides, which is the dominant process at typical plant pH values (Szabó et al., 2008). Nitrogen elimination is related to the removal of the colloidal matter (Aguilar et al., 2002). Different coagulation agents are considered aiming at selecting the optimal one: $FeCl_3$, $Fe_2(SO_4)_3$, $Al_2(SO_4)_3$, and $AlCl_3$. The flowsheet for the process of coagulation is presented in Figure 3.

Figure 3.- Scheme of the coagulation process.

The removal efficiency achieved is similar for the different coagulant agents, with values up to 99% for phosphorus and 57% for nitrogen (Aguilar et al., 2002). The main variables which influence the coagulation-flocculation process are the initial ratio of metal to phosphorus, pH, and Chemical Oxygen Demand (COD). The initial metal-phosphorus molar ratio must be between 1.5 and 2.0, and the recommended pH range is from 5.5 to 7. COD has a negative impact on the removal efficiency when its value is increased (Szabó et al., 2008).

To determine the amount of coagulant agent to be added to the system, it has been considered that a metal/phosphorus molar ratio of 1.75 must be achieved (Szabó et al., 2008). Given the relationship between the P in the raw material stream, the metal added, and the metal concentration in the commercial presentation of the coagulant agent, we are able to compute the coagulant agent amount that should be added. In the coagulation and flocculation tanks the flocs are formed and nutrients are recovered in the sediment together with coagulation agents and organic solids contained in the raw material. In the decanter, it has been considered that the stream with solids has a water content of 50% ($C_{Wa}^{sedimentator}$) (Williams and Esteves, 2011) and the water content of the centrifuge outlet solids stream is 60% ($C_{Wa}^{centrifuge}$) (Wakeman, 2007).

Other elements present in the digestate, such as total solids, carbon, and potassium are assumed to be present in the solid forming compounds that sediment. Thus, they are among species that constitute the cake. Taking into account the elements mentioned above mass balances have been formulated with the corresponding removal ratios. To select and evaluate the different coagulant agents, the problem has been modelled using a mixed-integer nonlinear programming (MINLP) formulation with Big-M constraints, Eqs. (17)-(18)

$$F_j^{coag\ tank} \geq \frac{F_P^{in}}{MW_P} \cdot MeP_{ratio} \cdot \frac{MW_j}{C_{Me}} - M \cdot (1 - y^j) \quad j \in \{coagulant\ agents\} \tag{17}$$

$$\sum y^j = 1 \tag{18}$$

Where MeP is the metal/phosphorus ratio and M is a number larger to formulate the Big-M disjunction to select and evaluate the different coagulant agents. Mass balances are computed using Eqs. (19)-(28).

$$F_j^{coag\ tank} = F_j^{floc\ tank} = F_j^{sedimentator} = F_j^{centrifuge} = F_j^{cake}\ j \in \{coagulants\} \tag{19}$$

$$F_i^{in} = F_i^{coag\ tank} = F_i^{floc\ tank} = F_i^{sedimentator} \quad i \in \{P, N\} \tag{20}$$

$$F_i^{cake} = F_i^{centrifuge} = F_i^{sedimentator} \cdot \eta_i^j \tag{21}$$

$$F_i^{sink1} = F_i^{sedimentator} - F_i^{cntrifuge} \tag{22}$$

$$F_k^{in} = F_k^{coag\ tank} = F_k^{floc\ tank} = F_k^{sedimentator} = F_k^{centrifuge} = F_k^{cake} \quad k \in \{TS, C, K\} \tag{23}$$

$$F_{Wa}^{in} = F_{Wa}^{coag\ tank} = F_{Wa}^{floc\ tank} = F_{Wa}^{sedimentator} \tag{24}$$

$$F_{Wa}^{centrifuge} = \left(F_{TS}^{centrifuge} + \sum_i F_i^{centrifuge} + \sum_j F_j^{centrifuge} \right) \cdot \frac{C_{Wa}^{sedimentator}}{1 - C_{Wa}^{centrifuge}} \tag{25}$$

$$F_{Wa}^{sink1} = F_{Wa}^{sedimentator} - F_{Wa}^{centrifuge} \tag{26}$$

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} + \sum_j F_j^{cake} \right) \cdot \frac{C_{Wa}^{centrifuge}}{1 - C_{Wa}^{centrifuge}} \tag{27}$$

$$F_{Wa}^{sink2} = F_{Wa}^{centrifuge} - F_{Wa}^{cake} \tag{28}$$

The estimation of the size and cost of both the coagulation and flocculation tanks has been carried out using a correlation developed by Almena and Martin (2016) as a function of the weight of the vessels. To simplify the mass balances it is considered that the volume provided by the coagulant agents is negligible with respect to the processed stream of the digestate. The vessel size is computed from the residence time. The hydraulic retention time considered in the coagulation tank is 4 min (Zhou et al., 2008). The vessel size is computed from the residence time, Eq. (29). Using this data, the diameter and length are computed using rules of thumb, Eqs. (30)-(31). Finally, a correlation for the thickness as a function of the diameters allows determining the mass of metal required for the vessel and its weight, Eqs. (32)-(33). Vessel cost estimation is provided by Eq. (34).

$$V_{Coag\ tank}\left(m^3\right) = HRT_{Coag\ tank} \cdot \frac{F_{digestate}^{in}}{\rho_{digestate}} \tag{29}$$

$$D_{Coag\ tank}\left(m\right) = \left(\frac{6 \cdot V_{Coag\ tank}}{7 \cdot \pi}\right)^{1/3} \tag{30}$$

$$L_{Coag\ tank}\left(m\right) = 4 \cdot D_{Coag\ tank} \tag{31}$$

$$e_{Coag\ tank}\left(m\right) = 0.0023 + 0.003 \cdot D_{Coag\ tank} \tag{32}$$

$$W_{Coag\ tank}\left(kg\right) = \rho_{SS316} \cdot \left[\pi \cdot \left(\left(\frac{D_{Coag\ tank}}{2} + e_{Coag\ tank}\right)^2 - \left(\frac{D_{Prec\ tank}}{2}\right)^2\right) \cdot L_{Coag\ tank} + \frac{4}{3} \cdot \pi \cdot \left(\left(\frac{D_{Coag\ tank}}{2} + e_{Coag\ tank}\right)^3 - \left(\frac{D_{Coag\ tank}}{2}\right)^3\right)\right] \tag{33}$$

$$Cost_{Vessel} = 6839.8 \cdot V_{Coag\ tank}\left(m^3\right)^{0.65} \tag{34}$$

To estimate the power consumed by the agitator, Eq. (35), the rules of thumb have been used where the specific power consumed, $\kappa_{agitator}$, is tabulated in Walas (1990). For our slurries a value of $\kappa_{agitator}$ equal to 10 HP per 1000 US gallons is the most appropriate.

$$P_{agitator}\left(HP\right) = V_{Coag\ tank}\left(US\ gallon\right) \cdot \frac{\kappa_{agitator}}{1000} \tag{35}$$

The agitator cost is also estimated using a correlation from Walas (1990), Eq. (36). For cost estimation purposes we have considered stainless steel 316 as construction material and a dual impeller operating at speed between 56 and 100 rpm depending on the tanks size. With this considerations the values for a, b and c are 8.8200, 0.1235 and 0.0818 respectively (Walas, 1990). This correlation provides the cost in 1985 dollars, so it is necessary to update the result using the Chemical Engineering Index as before.

$$Cost_{agitator\ 1985}\left(\$\right) = e^{a + b \cdot ln\left(P_{agitator}\left(HP\right)\right) + c \cdot \left[ln\left(P_{agitator}\left(HP\right)\right)\right]^2} \tag{36}$$

The total cost of the coagulation tank is equal to the sum of the vessel cost and the agitator cost, Eq. (37).

$$Cost_{Coag\ tank} = Cost_{Vessel} + Cost_{agitator\ 2016} \tag{37}$$

The flocculation tank is designed similarly to that of the coagulation, using Eqs. (29)-(37). For this step the hydraulic retention time is 25 min (Zhou et al., 2008).

The decanter is assumed to be circular because of its lower operating and maintenance costs. The area, Eq. (38), is computed using the parameter $A_{specific}$, which is the specific clarifier area in m^2 per ton of inlet flow per day (WEF, 2005). The typical value, 10 m^2/(t/day), is taken from Perry and Green (2008). The diameter of the clarifier, $D_{clarifier}$, is computed from the area value, Eq. (39).

$$A_{clarifier} = \frac{A_{specific} \cdot F_{digestate}^{in}\left(\frac{m^3}{day}\right)}{1000} \tag{38}$$

$$D_{clarifier} = \left(\frac{4 \cdot A_{clarifier}}{\pi}\right)^{1/2} \tag{39}$$

The number of clarifiers is an integer value that has been computed rounding up the ratio between the clarifier diameter calculated before and the maximum clarifier diameter, $D_{max}^{clarifier}$, Eq. (40). The maximum clarifier diameter value considered is 40 m (Perry and Green, 2008).

$$n_{clarifiers} \geq \frac{D_{total}^{clarifier}}{D_{max}^{clarifier}} \tag{40}$$

The diameter used in the final design will be the smallest between $D_{clarifier}$ and $D_{max}^{clarifier}$, Eq. (41).

$$D_{design}^{clarifier} = min(D_{max}^{clarifier}, D_{total}^{clarifier}) \tag{41}$$

To model the minimization function and compute $D_{clarifier\ design}$, the following smooth function approximation, given by Eq. (42), is used based on previous work (de la Cruz and Martin, 2016), to avoid discontinuities within the problem formulation.

$$D_{design}^{clarifier} = \frac{D_{max}^{clarifier}}{1+e^{\left(-F_{digestate}^{in}+0.342\right)\cdot 2.718}} \tag{42}$$

The cost estimation correlation has been developed from the data in WEF (2005), Eq. (43). It includes all the items involved in the operation of such an unit. The correlation must be updated to current prices using the Chemical Engineering Index.

$$Cost_{clarifier\ 1979} = \left(13060 \cdot D_{design}^{clarifier} - 58763\right) \cdot n_{clarifiers} \tag{43}$$

Centrifuge sizing and costing is based on the data by Perry and Green (2008). We assume pusher type with a maximum diameter of 1250 mm. The modelling equation for sizing is given in Eq. (44)

$$D_{Centrifuge}\left(in\right)=0.3308\cdot\frac{F_{digestate}^{in}}{1000}\cdot 3600+9.5092 \tag{44}$$

The number of centrifuges is calculated taking into account the maximum centrifuge diameter, Eq. (45), and the diameter used in the final design will be the minimum value between $D_{Centrifuge}$ and $D_{Centrifuge\ max}$, Eq. (46).

$$n_{centrifuges}\geq\frac{D_{total}^{centrifuge}}{D_{max}^{centrifuge}} \tag{45}$$

$$D_{design}^{centrifuge}=\min\left(D_{max}^{centrifuge},D_{total}^{centrifuge}\right) \tag{46}$$

As in the clarifier, we develop a smooth approximation, Eq. (47), to compute the design diameter avoiding discontinuities as follows:

$$D_{design}^{centrifuge}=\frac{D_{max}^{centrifuge}}{1+e^{\left(-F_{digestate}^{in}+35.369\right)\cdot 0.0395}} \tag{47}$$

Thus, the cost for the centrifuge is estimated based on the data by Perry and Green (2008) as a function of its diameter, Eq. (48). Since the cost correlation is based on 2004 values, the Chemical Engineering Index it used to update the equipment cost.

$$Cost_{centrifuge\ 2004}\left(\$\right)=\left(10272\cdot D_{design}^{centrifuge}-24512\right)\cdot n_{centrifuges} \tag{48}$$

We estimate the operating cost of this system by accounting for the annualized equipment cost (fixed cost), chemicals and labor cost. A similar procedure as before is followed Vian Ortuño (1991) but for the clarifier fixed costs as the correlation to estimate its costs already includes the operating cost, Eq. (49).

$$FC_{coagulation}\left(\frac{€}{year}\right)=\left(Cost_{Coag\ tank}+Cost_{Floc\ tank}+Cost_{centrifuge\ 2016}\right)\cdot f_i\cdot f_j+Cost_{clarifier\ 2016} \tag{49}$$

The chemicals costs are estimated as Eq. (50)

$$ChemC_{coagulation}\left(\frac{€}{year}\right)=\left(F_{Fe_2(SO_4)_3}^{in}\cdot Price_{Fe_2(SO_4)_3}+F_{Al_2(SO_4)_3}^{in}\cdot Price_{Al_2(SO_4)_3}+F_{FeCl_3}^{in}\cdot Price_{FeCl_3}+F_{AlCl_3}^{in}\cdot Price_{AlCl_3}\right)\cdot 3600\cdot h\cdot d \tag{50}$$

To estimate the price for the cake, as in the previous case, we assume the price of each of the nutrients contained (N, P, and K). The price for each nutrient is taken same as before. Thus, the cake price is computed as the weighted sum of each nutrient, as in Eq. (15) (Hernandez et al., 2017).

Finally, the economic benefits or losses of operating this system are calculated as the difference between the credit obtained from the cake and the operating costs of the section of the facility, as in Eq. (16).

3.4.3.-Centrifugation

Centrifugation is a pretreatment that separates solid and liquid phases and that can be used to recover nutrients from the digestate. The advantage of this system is the simple equipment used. Precipitant agents can be added to improve the removal efficiency significantly. Previous studies show that an appropriate mixture of $CaCO_3$ and $FeCl_3$ promotes nutrients recovery. In particular, a ratio of 0.61 kg $CaCO_3$ per kilogram of total solids in the raw material inlet stream, and 0.44 kg of $FeCl_3$ per kilogram of total solids in the raw material inlet stream, achieves a removal efficiency up to 95% and 47 % for P and N respectively (Meixner et al., 2015). Figure 4 presents a scheme of the process.

Figure 4.-Scheme for the centrifugation treatment

Centrifugation process consists of two units, a precipitation tank where $CaCO_3$ and $FeCl_3$ are added, and the centrifuge. These equipment have been modeled using mass balances and removal ratios for the precipitating species. Note that the total solids, carbon, and potassium are assumed to be present in the form of solid compounds, so they will be removed as part of the cake. Moreover, the water content of the centrifuge outlet solids stream is assumed to be 60% ($C_{Wa}^{centrifuge}$) (Wakeman, 2007). Mass balances for the process have been evaluated in Eqs. (51)-(59):

$$F_j^{in} = F_{TS}^{in} \cdot \frac{\varphi_j}{C_j} \quad j \in \{precipitation\ agents\} \tag{51}$$

$$F_j^{in} = F_j^{prec\ tank} = F_j^{centrifuge} = F_j^{cake} \tag{52}$$

$$F_i^{in} = F_i^{prec\ tank} = F_i^{centrifuge} \quad i \in \{P,N\} \tag{53}$$

$$F_i^{cake} = F_i^{centrifuge} \cdot \eta_i \tag{54}$$

$$F_i^{liquid\ effluent} = F_i^{centrifuge} - F_i^{cake} \tag{55}$$

$$F_k^{in} = F_k^{prec\ tank} = F_k^{centrifuge} = F_k^{cake} \quad k \in \{TS, C, K\} \tag{56}$$

$$F_{Wa}^{in} = F_{Wa}^{prec\ tank} = F_{Wa}^{centrifuge} \tag{57}$$

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} + \sum_j F_j^{cake} \right) \cdot \frac{C_{Wa}^{centrifuge}}{1 - C_{Wa}^{centrifuge}} \tag{58}$$

$$F_{Wa}^{liquid\ effluent} = F_{Wa}^{centrifuge} - F_{Wa}^{cake} \tag{59}$$

Where φ_j is the precipitation agent per total solids mass ratio (0.61 kg $CaCO_3$/ kilogram TS and 0.44 kg $FeCl_3$ / kilogram TS).

These units have been designed using correlations as a function of the flow processed. For the design of the precipitation tank (volume, diameter, length, thickness, weight, and cost calculations) the equations provided by Almena and Martin (2016) have been used as before, Eqs. (29)-(37), considering a hydraulic retention time of 2.5 min (Szabó et al., 2008).

$$V_{Prec\ tank}\left(m^3\right) = HRT_{Prec\ tank} \cdot \left(\frac{F_{digestate}^{in}}{\rho_{digestate}} + \frac{F_{FeCl_3}^{in}}{\rho_{FeCl_3}} \right) \tag{60}$$

The volume of $CaCO_3$ added is assumed negligible compared to the volume of the liquid because of it is added as solid. Thus, the diameter of the tanks is computed using Eq. (60) and Eq. (30) as in the previous unit. The cost of the vessel is given by the weight of the metal, using the correlations provided by Almena and Martin (2016), Eqs. (31)-(34). The power required is computed, as in previous cases, using the rules of thumb in Walas (1990), Eq. (35), where the value of $\kappa_{agitator}$ is equal to 10 HP per 1000 gal, in accordance with the data collected in the literature (Walas, 1990). The cost correlation is given by Eq. (36) and updated to 2016 prices. The total cost of the precipitation tank included the vessel and the agitator costs, Eq. (37).

The centrifuge size is characterized by its diameter. We model it as in the previous technology using Eqs. (44)-(48). The operating costs involve fixed, chemicals and labour costs. Fixed costs are estimated using Eq. (61). The labor cost is estimated in Eq. (13), where n_{OP} is equal to 1 (Vian Ortuño, 1991). Total operating cost is given by Eq. (14).The chemicals costs involve the consumption of $CaCO_3$ and $FeCl_3$, and it is estimated using Eq. (62):

$$FC_{centrifugation}\left(\frac{€}{year}\right)=\left(Cost_{centrifuge\ 2016}+Cost_{Prec\ tank}\right)\cdot f_i\cdot f_j \tag{61}$$

$$ChemC_{centrifugation}\left(\frac{€}{year}\right)=\left(F^{in}_{CaCO_3}\cdot Price_{CaCO_3}+F^{in}_{FeCl_3}\cdot Price_{FeCl_3}\right)\cdot 3600\cdot h\cdot d \tag{62}$$

The cake recovered is the main asset of the process. Its price is estimated as the weighted sum of each nutrient, Eq. (15), (Hernandez et al., 2017). Finally, the benefits or losses of operating this system are calculated as the difference between the revenue obtained from the cake and the operating costs of the facility, Eq. (16).

3.4.4.-Struvite production

P and N can be recovered from digestate through the formation of struvite, which is a phosphate mineral with a chemical formula of $MgNH_4PO_4{\cdot}6H_2O$. The advantage of this technology is that struvite is a solid with a high nutrients density, it is easy to transport, and it can be used as slow-release fertilizer without any post-processing (Doyle and Parsons, 2002). The removal of nutrients via struvite production follow the reaction below, requiring the addition of $MgCl_2$, resulting in the production of struvite crystals that can be recovered as solid:

$$Mg^{2+}+NH_4^{+}+H_nPO_4^{3-n}+6H_2O\Leftrightarrow MgNH_4PO_4{\cdot}6H_2O+nH^{+} \tag{63}$$

Due to the presence of potassium in the digestate, together with struvite, another product called potassium struvite or K-Struvite, is also produced. In this case the ammonia cation is substituted by the potassium cation (Wilsenach et al., 2007).

$$K^{+}+Mg^{2+}+H_nPO_4^{3-n}+6H_2O\Leftrightarrow KMgPO_4{\cdot}6H_2O+nH^{+} \tag{64}$$

Since the formation of struvite is favored over the formation of K-Struvite, it is considered that only 15% of the potassium contained in the digestate will react to form K-Struvite (Zeng and Li, 2006). The mass balance for the reactors is given by the stoichiometry of the reactions above.

Two different types of reactors can be used to obtain struvite, either a stirred tank (CSTR) or a fluidized bed reactor (FBR). Figures 5 and 6 provide detailed flowsheets of each case. In case of the FBR, struvite is recovered from the bottoms and the liquid must be processed in a hydrocyclone to avoid discharging fines. In the case of CSTR tanks, we need to use a centrifuge to recover the struvite. We can help the crystal growth by seeding (Doyle and Parsons, 2002; Kumashiro et al., 2001). Due to the substantial increase in the struvite formation yield, we consider the addition of struvite seeds in both cases. The reaction takes place at about 27ºC, with the addition of $MgCl_2$ at a concentration of 57.5 mg/dm^3 (Zhang et al., 2014). A Mg:P molar ratio of 2 (Bhuiyan et al., 2008) is used.

Figure 5.- Scheme for the FBR system

The FBR system is composed of three elements: a mixer tank, a FBR rector, and a hydrocyclone. The system operation consists of a digestate flow which is mixed with a stream of $MgCl_2$ in the mixing tank. The addition of $MgCl_2$ helps precipitate the struvite by increasing the concentration of the species inside the reactor. As the concentration of NH_4^+ is high due to the pH, and the inorganic N and P are the elements we want to recover, the only element which is necessary add is Mg in form of $MgCl_2$.

In the tank there is a suspension of struvite seeds with a size of 0.8 mm which promote the precipitation of struvite. The solid struvite is evacuated from the reactor at the bottom and its moisture is low enough to avoid the use of a dryer. The other stream which leaves the reactor contains liquid water in a high proportion with the excess of Mg, the total solids from the digestate, and low amounts of nutrients and other components. This stream is introduced in a hydrocyclone to recover fines of struvite which can be removed by this stream. 100% of fines removal is assumed but no fines production is considered in the model.

To estimate the cost of this system we evaluate the effect of the following variables, whose operating values are shown between parenthesis:

- Digestate input mass and volume flow (between 1 and 100 kg/s)
- Recovered struvite humidity (5% in mass)
- Amount of phosphorus recovered (90%)
- Mg:P molar ratio with a value of 2

In an FBR there are some variables which influence in the design and hence the cost. The variables considered in this work are showed below with the typical values used in the present study between parenthesis:

- d_p: bed particle diameter, assumed to be 0.8 mm (Jordaan, 2011)
- Sphericity: 0.6 is a standard sphericity for particles used in fluidized bed reactors (Fogler, 2005)

Furthermore, the reaction kinetics and equilibrium are considered to estimate the residence time in the reactor. A first order kinetics, developed by Nelson et al. (2003), has been used, Eqs. (64)-(65). The kinetic constant is 3.42 x10^{-3} s^{-1} for a pH of 9.

$$\frac{-dC}{dt} = k\left(C - C_{eq}\right) \tag{64}$$

$$\ln\left(C - C_{eq}\right) = -kt + \ln\left(C_0 - C_{eq}\right) \tag{65}$$

Struvite formation is an equilibrium reaction. We use the equilibrium ion activity product (IAP_{eq}) value of $7.08\cdot10^{-14}$ (Nelson et al., 2003) to calculate the equilibrium concentrations in the kinetic model, Eq. (66). We assumed that the values of ions concentration are equal to ions activity.

$$IAP_{eq} = \left(Mg^{2+}\right)\left(NH_4^{+}\right)\left(PO_4^{3-}\right) = 7.08\cdot10^{-14} \tag{66}$$

Minimum fluidization velocity is calculated in the first step by considering that the fluid stream is a liquid (Mangin and Klein, 2004). This consideration is motivated because the liquid digestate works as fluidization agent (Le Corre, 2006). The digestate density is 950 kg/m^3 (Rigby and Smith, 2011). The expression used to calculate u_{mf} through Reynolds and Archimedes numbers is given by Eq. (67), (Tisa et al., 2014).

$$u_{mf} = \frac{\mathrm{Re}_{l\,mf}\cdot\mu_{digestate}}{\rho_{digestate} - d_p} \tag{67}$$

Eq. (67) parameters are determined by Eq. (68) and Eq. (69).

$$\mathrm{Re}_{l\,mf} = \sqrt{33.72 + 0.0404 Ar_l\left(1-\alpha_{mf}\right)^3} - 33.7 \tag{68}$$

$$Ar_l = \rho_{digestate}\left(\rho_{struvite} - \rho_{digestate}\right) g \frac{d_p^3}{\mu_{digestate}^2} \tag{69}$$

If the flow has no gas phase, α_{mf} is equal to cero. The terminal velocity is computed using Eq. (70) (Tisa et al., 2014).

$$u_t = \left(\frac{1.78\cdot10^{-2}\cdot\eta^2}{\rho_{digestate}\cdot\mu_{digestate}}\right)^{1/3} d_p \tag{70}$$

Where the parameter η is given by Eq. (71)

$$\eta = g\left(\rho_{struvite} - \rho_{digestate}\right) \tag{71}$$

Finally, the fluid velocity u_0 must be between u_{mf} and u_t. A superficial velocity equal to five times the minimum fluidization velocity is selected (Tejero-Ezpeleta et al., 2004), Eqs. (72)-(73)

$$u_{mf} < u_0 < u_t \tag{72}$$

$$u_0 = 5 \cdot u_{mf} \tag{73}$$

Once the superficial velocity is computed, the area and diameter can be calculated from the mass flow Eqs. (74)-(75).

$$A_{FBR} = \frac{F_{digestate}^{in}}{u_0} \tag{74}$$

$$D_{FBR} = \sqrt{\frac{4A_{FBR}}{\pi}} \tag{75}$$

The length of the bed is determined by the residence time through the kinetics and the equilibrium ion activity product presented above. Consequently, the magnesium and ammonium concentrations can be calculated from the digestate mass balance and the external magnesium added. Using the IAP_{eq} value, the phosphate concentration in equilibrium at the operational conditions can be determined. This equilibrium value will be used in kinetics, Eq. (76).

$$t = \frac{\ln\left(C_0 - C_{eq}\right) - \ln\left(C - C_{eq}\right)}{k} \tag{76}$$

Thus, the bed length is computed as per Eq. (77). Typically, the length of the reactor must be 15% larger than the bed, Eq. (78).

$$L_{bed} = \frac{u_0}{t} \tag{77}$$

$$L_{FBR} = 1.15 \cdot L_{bed} \tag{78}$$

The estimation of the rector cost is carried out assuming that it is a vessel as presented in the processes above, Eqs.(30)-(34), (Almena and Martin, 2016). The cost of the mixer tank is also estimated as that of a vessel , using Eqs. (29)-(34), with a volume given by that to provide a hydraulic retention time of 150 s (Szabó et al., 2008). The impeller is also designed using the same procedure as before, Eqs. (35)-(36), (Wallas, 1990).

Finally, to estimate the cost of the hydrocyclone, a surrogate model using data from Matche's website has been developed (www.matche.com). There is a maximum diameter, therefore, if a unit larger than the standard is required, we actually need to duplicate the equipment, Eq. (80). To estimate the diameter, we considered that there is a linear relationship between the diameter and the flow based on rules of thumb in design literature. A typical unit size of a 20 inch diameter hydrocyclone can process 1000 US gallons per minute, Eq. (79) (Walas, 1990).

$$D_{hydrocyclone}\left(in\right)=F_{digestate}^{in}\left(\frac{US\ gallon}{min}\right)\cdot\frac{20}{1000} \tag{79}$$

$$n_{hydrocyclone} \geq \frac{D_{total}^{hydrocyclone}}{D_{max}^{hydrocyclone}} \tag{80}$$

Where $n_{hydrocyclone}$ is an integer. The maximum diameter for a hydrocyclone, $D_{max}^{hydrocyclone}$, is 30 inch based on standard sizes (www.matche.com). Thus, the design diameter is the lower diameter between $D_{total}^{hydrocyclone}$ and $D_{max}^{hydrocyclone}$, Eq. (81).

$$D_{design}^{hydrocyclone} = \min(D_{total}^{hydrocyclone}, D_{max}^{hydrocyclone}) \tag{81}$$

The estimation of the cost for the fines recovery equipment is computed using Eq. (82) and updated as explained above.

$$Cost_{hydrocyclone\ 2014} = n_{hydrocyclone}\cdot\left(2953.2\cdot D_{design}^{hydrocyclone} - 34131\right) \tag{82}$$

The CSTR process consists of four elements: the CSTR reactor, a centrifuge, and a dryer with its corresponding heat exchanger. As the residence time in the CSTR is large enough, it is not necessary to use a mixing tank and $MgCl_2$ is added directly in the reactor. Thus, struvite is formed in one step in the CSTR. Since the digestate already contains NH_4^+ and P, we need to add $MgCl_2$. As a result, struvite precipitates, and it is recovered from the bottoms of the reactor and dried in a two step process. The first step consists of a centrifuge that recovers struvite with 5% (on weight basis) water (Baasel, 1977). Next, a drum dryer is implemented to remove the residual moisture to reach commercial standards and reduce transportation costs. Figure 6 shows the details of the flowsheet.

Figure 6.-Scheme for the CSTR based struvite production system

The design of the units involved in this process and their cost estimation is based on the following variables:

- Digestate input mass and volume flow (between 1 and 100 kg/s)
- Recovered struvite water content (5% in mass)
- Amount of phosphorus recovered (90%)
- Mg:P molar ratio with a value of 2

The CSTR is assumed to be a stirred vessel; consequently, it is designed as in the previous cases, Eqs. (29)-(37), with a residence of 471.05 s. The residence time is calculated from mass balances and the kinetics described in the FBR process, Eq.(64) and Eq. (65).

The centrifuge size is characterized by its diameter. Both, the size and cost are computed using the data in Perry and Green (2008). We assume a pusher type for the centrifuge with a maximum diameter of 1250 mm as before, Eqs. (44)-(48).

The cost estimation for the drier relies on the amount of water to evaporate, and the evaporation capacity. The evaporation capacity ($e_{capacity}$) is reported in the literature to be equal to 0.01897 (kg/(s·m^2)) (Walas, 1990). Consequently, the dryer cost is computed using a correlation provided by Martin and Grossmann (2011), Eq. (83), updating the cost to current prices using the Chemical Engineering Index.

$$Cost_{dryer\ 2007}(\$) = 1.15 \cdot \left(6477.1 \cdot \frac{F_{water}^{in}}{e_{capacity}} + 102394 \right) \quad (83)$$

The operating cost of the CSTR and the FBR based processes is computed considering three items, fixed, chemicals and labor, and assuming that utilities account for 20% of the operating costs. The correlations for computing each of them are taken from Vian Ortuño (1991) and Sinnot (1999), Eqs. (13) for labour and (14) for total operating cost. Fixed cost for struvite processes is calculated using Eq. (84). We assume that the seeds required for the FBR process are internally produced in the startup of the facility.

$$FC_{struvite}\left(\frac{€}{year} \right) = \left(\sum Cost_{equipment} \right) \cdot f_i \cdot f_j \quad (84)$$

The revenue obtained from the struvite is determined assuming a selling price of 0.763 €/kg, Eq. (85), (Molinos-Senante et al., 2015) .

$$Cost_{struvite}\left(\frac{€}{year}\right)=\left(F_{struvite}^{recovered}\cdot Price_{struvite}\right)\cdot 3600\cdot h\cdot d \tag{85}$$

Finally, the benefits or losses for CSTR and FBR are calculated as the difference between the credit obtained from the struvite and the operating costs of the facility, Eq. (16).

3.5.- Solution procedure

The detailed models for each of the alternatives such as the five filter media or the number of different coagulants result in a large and complex MINLP when cost estimation is involved. We use a two-stage procedure to select the best technology. In the first stage we develop MINLP subproblems to select the appropriate filter media or coagulant. Next, using the detailed models for the best option, surrogate cost models are developed for the five alternative technologies used to process the digestate. However, there are still binary decisions to account for the cost of the active alternative in the superstructure. Thus, the surrogate models are in the form of linear equations. For instance, the surrogate model for the filter to be implemented in the superstructure is given by a linear function as given by Eq. (86).

$$Operating\ cost\left(\frac{€}{year}\right)=20521\cdot F_{design}\left(\frac{ft^3}{\min}\right)-33488\cdot a_{Filter} \tag{86}$$

We avoid the use of binary variables within the formulation (due to highly non linear model of the entire superstructure) by using smooth approximations. We define a_{Filter} as a parameter that takes a value of 0 when F_{design}^{Filter} is 0 and 1 if F_{design}^{Filter} is not equal to 0. The smooth approximation for a_{Filter} is defined as follows, Eq. (87):

$$a_{Filter}=\frac{1}{1+e^{\left(-F_{design}+0.049\right)\cdot 361}} \tag{87}$$

Metal slag is selected as the best filter for the filtration process. For the case of the coagulants, the solution of the subproblem, Eqs. (17)-(50) selects the use of $AlCl_3$. As in the previous case, a surrogate model is developed to be included in the superstructure so that we avoid including binary variables and allow for zero operating costs in case this technology is not selected, Eq. (88).

$$Operating\ cost\left(\frac{€}{year}\right)=1019589.91\cdot F_{digestate}^{in}\left(\frac{kg}{s}\right)-368838.56\cdot a_{Coag} \tag{88}$$

Where the smooth approximation for the term a_{Coag} is given by Eq. (89)

$$a_{Coag} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.068\right)\cdot 863}} \qquad (89)$$

Similar to previous cases we develop a surrogate model to estimate the operating cost for the centrifugation as a function of the flowrate of digestate, Eq. (90):

$$Operating\ cost\left(\frac{€}{year}\right) = 458498.29\cdot F_{digestate}^{in} + 24924.67\cdot a_{Centrifugation} \qquad (90)$$

As before, $a_{Centrifugation}$ is approximated as follows, Eq. (91):

$$a_{Centrifugation} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.068\right)\cdot 863}} \qquad (91)$$

Finally, to include the operating costs for the production of struvite, we again develop surrogate models for the FBR, Eq. (92) and for the CSRT Eq. (94), where a smooth approximation is proposed for the fixed term, a_{FBR} and a_{CSTR} respectively, Eqs. (93) and (95)

$$Operating\ Cost_{FBR}\left(\frac{€}{year}\right) = 245008\cdot F_{digestate}^{in} + 1\cdot 10^{6}\cdot a_{FBR} \qquad (92)$$

$$a_{FBR} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.06785\right)\cdot 862.9679}} \qquad (93)$$

$$Operating\ cost_{CSTR}\left(\frac{€}{year}\right) = 277051\cdot F_{digestate}^{in} + 1\cdot 10^{6}\cdot a_{CSTR} \qquad (94)$$

$$a_{CSTR} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.06785\right)\cdot 862.9679}} \qquad (95)$$

The benefits/losses in the superstructure for any of the technologies to process the digestate is computed as the difference between the revenue obtained from the nutrients and generated power, and the operating costs of the facility.

Finally, the whole superstructure is built (see Figure 1). This superstructure contains models of the fermenter, biogas purification, gas cycle, steam cycle, and digestate treatment processes. The aim of this superstructure is to determine the optimal operating conditions and to select the best digestate treatment technology. Thus, digestate treatment processes have been implemented in the superstructure through detailed mass balances including the solution to the kinetics of the fluidized bed reactors as well as the surrogate models developed in the previous stage to estimate the operating costs. It should be noted that in filtration, centrifugation, and coagulation processes we have included a benefits penalty, $F_{total}^{\text{recovered}}$, due to the fact that the product recovered is a mixture of nutrients and organic matter with a nutrients concentration lower than struvite. This penalty represents the concentration of nutrients in the recovered product given by the ratio between the nutrients recovered and the total recovered mass flow, Eq. (96).

$$\text{Price}\left(\frac{€}{year}\right) = \left(F_P^{\text{recovered}} \cdot \text{Price}_P + F_N^{\text{recovered}} \cdot \text{Price}_N + F_K^{\text{recovered}} \cdot \text{Price}_K\right) \cdot \frac{1}{F_{total}^{\text{recovered}}} \cdot 3600 \cdot h \cdot d \tag{96}$$

The total energy obtained in the system to be optimized is the sum of the one generated at the three sections of the turbine, high, medium and low pressure and that of the gas turbine. We use part of the energy produced to power the compressors used across the facility. The economic benefits or losses of each digestate treatment process are added to the energy benefits.

$$\begin{aligned} Z = & \left[\left(\sum_{i \in turbinebody} \text{W}_{(\text{Turbine})} + W_{(GasTurb)} - \sum_{j \in compressors} W_{(compressors)}\right) \cdot 3600 \cdot h \cdot d \cdot C_{\text{Electricity}}\right] \\ & + \text{Benefits}_{\text{Filtration}} + \text{Benefits}_{\text{Centrif}} + \text{Benefits}_{\text{Coagulation}} + \text{Benefits}_{\text{FBR}} + \text{Benefits}_{\text{CSTR}} \end{aligned} \tag{97}$$

Eq. (97) is the objective function that we maximize to determine the optimal operational conditions and to select the best digestate treatment process subject to the following constraints:

a) Bioreactor and biogas composition model
b) Digestate processing
c) Biogas purification. Described in section 3.2
d) Brayton cycle. Described in section 3.3.1
e) Rankine cycle. Described in section 3.3.2

The main decision variables are related to the selection of the digestate processing technology, among filtration, centrifugation, coagulation and struvite production using CSTR or FBR. The decision variables are also associated with the selection of the type of filter and the coagulation agent. Furthermore, the biogas usage to produce steam requires the operating pressures and temperatures at the gas turbine, and the steam turbine as well as the extraction form the steam turbine to reheat the condensate before regenerating steam using the flue gas from the gas turbine. The superstructure consists of an NLP of approximately 4000 equations and 5000 variables solved using a multistart procedure with CONOPT 3.0 as the preferred solver. The computational time is around 60 min, although it varies for each problem as a consequence of the different data used in each case.

4.-Results

Following the optimization procedure presented in section 3.4 we first decide on the filter media and the coagulant chemical. We solve MINLP subproblems leading to the selection of the filter media and the coagulant agent. We use the metal slag as the filter media and the $AlCl_3$ as the coagulant for all raw materials. Next, we developed surrogate models for the five technologies included in the superstructure and solve a reformulated NLP including smooth approximations for the cost functions of the digestate treatment so as to maximize the power produced and the treatment section. The plant size is assumed to be that which processes 10 kg/s of manure based on the typical amount of manure produced in cattle farms (León, 2015). Four manures have been evaluated on the plant: cattle, pig, poultry and sheep, with the aim of determining, for each one, the power generated the composition of the biogas produced, the optimal digestate treatment technology to recover its nutrients and the biogas-manure and digestate-manure ratios. Section 4.1 summarizes the main operating conditions of the major units in the process and the selection of digestate processing technology. Section 4.2 presents the detail economic evaluation of the four optimal processes, one per manure type. Finally, in section 4.3 an analysis of the effect of the manure composition on the power, operating conditions and digestate treatment is performed.

4.1.-Mass and energy balances

Table 4 shows the main operating conditions of major units for the four different manure types. Cattle, pig, and poultry show similar values among them and to previous work (León and Martín, 2016). The gas in the gas turbine reaches a temperature of 2400 ºC and a pressure of 8.2 bar before expansion for cattle, pig and poultry manure. However, sheep manure shows different values. While the temperature is similar, the pressure is 15.6 bar, almost twice the value found for the rest of the raw materials. Furthermore the flue gas exits the turbine 300 ºC below that when the rest of the manure types are used. Furthermore while the high pressure of the steam turbine is 125 bar for cattle, pig, and poultry manure, in case of sheep manure the steam turbine operates at 95 bar at the high pressure section of the turbine. This is related to the lower gas temperature from the gas turbine since the overheated steam needs to be produced using that stream. Intermediate and low pressures are the same in the steam turbine using any of the manure types, but the exhaust pressure of the steam is higher in case of sheep manure. Table 5 shows the products obtained from the various manure types, power, biogas, and digestate. Poultry is the waste that is more efficient towards power production due to its higher concentration. In all cases an FBR reactor for the production of struvite is the selected technology to recover N and P. In the table we also see the effect of the fact that cattle and pig manure are mostly liquids, since most of the product is digestate, almost 98%, while the use of poultry or sheep manure reduces the production of digestate to 75% and 88% respectively, increasing the production of biogas and power. Finally in Table 6 the biogas composition for each manure considered are presented. The main purpose of the facility is the production of power. However, the biogas composition is typically within a range of values per component that have been imposed as bounds. As a result of maximizing the electricity production for all studied cases, the same biogas composition is obtained, 67.5% molar in CH_4 and the rest is mostly CO_2.

Table 4.- Operating data of the optimal configuration for each raw material.

Table 5.- Process optimization results for considered manures

Table 6.- Biogas composition for considered manures

4.2.-Economic evaluation

This section is divided into the estimation of the investment cost, using a factorial method based on the cost of the units, and the estimation of the electricity production cost.

4.2.1.- Investment cost

We use the factorial method to estimate the investment cost for this facility. This is based on the estimation of the equipment cost and several coefficients to account for pipes, installation, etc. (Sinnot and Towler, 2009). The cost for the different units has been estimated based on Matche's website (www.matche.com), Sinnot and Towler (2009) and Peters and Timmerhaus (2003), updating the cost of the units when required. We assume a plant that processes fluids and solids. Due to the different composition of each manure the specific production of biogas for each one is different, being larger for poultry and sheep than for cattle and pig. The reason for that could be that sheep and poultry manures have less water content while the water content in cattle and pig reaches 98% (http://adlib.everysite.co.uk). For cost estimation proposes the digester maximum size considered is 6000 m^3 per unit, since the larger units could face mixing and homogenization problems (FNR, 2010). This result for the facility investment cost will be different for each raw material. Figure 7 shows the equipment cost distribution where digester and gas turbine are the most important contributions:

- Cattle manure: A plant that processes 10 kg/s of this type of manure requires an investment of 69.1 M€, of which 14.9 M€ represents the equipment cost. The larger cost is assumed by the digester units, with a 75% of the total units cost, followed by the heat exchanger network with a contribution of 12% while both turbines add up to 12%.
- Pig manure: A facility to process 10kg/s of this manure requires in an investment of 69.5 M€, with a cost of 14.9 M€ in equipment. Since the digestate-manure and biogas-manure ratios between cattle and pig manure are very similar, the investment costs are analogous among them. The unit cost distribution is similar to the cattle manure case.
- Poultry manure: The investment for a plant which processes 10kg/s of this manure is 208.0 M€. The units investment adds up to 44.7 M€. In this the units cost distribution is more homogeneous among different items: 60% to digester units, 20% to gas turbine, 10% to heat exchanger network and 9% to steam turbine. It should be noted that, as poultry manure has a high content of dry matter (around 60% on a weight basis), it is necessary to add additional water to decrease the dry matter content to reach 25% with the aim of avoid mixing problems in the digester due to an excessive solids concentration inside.
- Sheep manure: The facility to treat 10kg/s of this manure requires an investment of 105.0 M€, where 22.5 M€ represents the equipment cost. For this plant the main units cost distribution is as follows: 50% for the digester, 25% for gas turbine, 17% for heat exchanger network and 7% for steam turbine.

It is clear that the digester shows the highest share in the investment cost and therefore the concentration of the manure highly determines the cost of the facility. Lantz (2012) presented the investment cost of a facility for heat and power production as a function of its scale. Actually, our plant does not produce steam as a final product but only power. Thus, it is interesting to see that the raw material determines the investment per kW from the 4000€/kW in case of poultry manure or the 7500€/kW in case of sheep manure, to the more than 25000 €/kW in case of pig and cattle.

Figure 7: Units cost distributions for cattle, pig, poultry and sheep manure treatment (ST: Steam turbine, GT: Gas turbine, HX: Heat exchangers, FBR: Fluidized bed reactor).

4.2.2.- Production cost

To calculate the production cost, 20 years of plant life is considered, with a capacity factor of 98%. Apart from the equipment amortization, other items are also taken into account such as salaries, administrative fees, chemicals cost, maintenance cost, utilities and contingency costs. Thus, apart from the annualized equipment cost, 1.5 M€ are spent in Salaries, 0.25 M€ in Administration, 2M€ in Maintenance, 0.25 M€ in other expenses (Martín and León, 2016) while chemicals are computed as described in section 2. The cost of utilities adds up to 0.08 M€, accounting for the cooling water and the steam needed to maintain the operation of the digester and to condition the digestate for its use as a fertilizer. Finally, we assume that the livestock manure is for free. Figure 8 shows the distribution of the production costs for each of the manure types. We see that the figures are very similar. The equipment amortization represents at least 43% of the production costs. This share increases up to 60% for the case of the use of poultry. As the investment is lower, the annual cost for other items is almost constant and their contribution to the electricity cost plays a more important role. Chemicals is the second most important contribution to the cost of electricity with a share of up to 23% for the use of cattle or pig manure and down to 16% in the case of sheep manure. We assume in all cases that waste is for free. Under these considerations the electricity production costs obtained are presented in Table 7.

Table 7: Electricity production cost and NPV for the facility considering different raw materials

The Net Profit Value has also been calculated as a measure of the project profitability, considering an electricity price of sale of 0.06 €/kWh. To compare the profitability of this project a secure investment as the inversion in Spanish national debt has been chosen, considering a discount rate of 3% (Ministerio de Economía, Industria y Competitividad, 2017). The results obtained are presented in Table 7, and it should be noted that facilities for poultry and sheep manures obtain positive NPV while those which use cattle and pig manure as raw material show negative NPV, so from the point of view of NPV as an indicator to decide the project viability, those ones would be disregarded.

Figure 8: Operation cost distribution for cattle, pig, poultry and sheep manure treatment

4.3.- Effect on the power, operating conditions and digestate treatment

The results obtained from the treatment of different manure streams show the influence of the manure composition in the amount produced and the composition of biogas and digestate obtained. Struvite production using FBR is the best choice for digestate treatment. This can be explained by the advantages in recovering nutrients in solid form since they can be easily transported and stored. Furthermore the material is highly concentrated in nutrients with a relatively high selling price.

Biogas production is similar for cattle and pig manures, but is significantly higher in the poultry and sheep cases. The investment cost when processing cattle and pig manure is dominated by the digester, resulting in similar investment and production costs for facilities using either of the two types of manure. However, the higher concentration in organic matter in sheep and poultry manure does not only results in higher power production capacities, but the fact that the contribution to the cost of the turbines is also larger and so is the investment cost of these facilities. On the other hand, the electricity production cost is lower in the last two cases as result of the economies of scale between the investment cost and the biogas produced and the higher amount of struvite produced, with the extreme case of poultry manure where the struvite selling benefits are capable to cover the electricity production costs. Note that the availability of poultry and or sheep manure should be less that than for cattle and pig manure.

5.-Conclusions

In this work, we have designed optimal integrated facilities for the production of biogas-based electrical power and fertilizers from manure. Detailed equation based models for the anaerobic digestion, the Brayton and regenerative Rankine cycles and different technologies for digestate treatment have been developed. To solve the model a two-step procedure has been performed. First, the individual detailed models for each digestate treatment technology are used to formulate a MINLP model aiming at selecting the best configuration for that technology: the best precipitation agent, filter media, etc. In the second step, the best configuration of each technology has been implemented in the entire superstructure. Due to the fact that only one digestate processing technology is allowed and the highly non-linear nature of the model, surrogate models for the cost of each alternatives with a smooth approximations have been developed. For the optimal selection a detailed economic evaluation is performed.

The results show that FBR technologies are preferred to recovery nutrients. Furthermore, in some cases this process can produce electricity at a competitive price (in case of poultry and sheep manure). The investment cost is highly dependent on the water and organic content of the manure type, ranging from 70 M€ to 208 M€ when a large energy production is possible and large gas and steam turbines are to be installed. However, for these cases of high investment cost, the production cost of power is the most competitive due to the large production capacity. Biogas power plants show a wide range of values of power per kW installed depending on the manure concentration. Competitive values of 4000€/kW for poultry manure are obtained, due to the highly concentrated manure, while large values of 25000€/kW installed are reported in case of the diluted cattle or pig manure.

Nomenclature

Sets

$i \in \{P, N\}$

$j \in \{\textit{filter media}\}$

$k \in \{TS, C, K\}$

$a' \in \{CH_4, CO_2, NH_3, H_2S, O_2 \text{ and/or } N_2\}$

$a \in \{H_2O, CH_4, CO_2, NH_3, H_2S, O_2 \text{ and/or } N_2\}$

$d \in \{C, Norg, Nam, P, K, H2O \text{ and/or } Rest\}$

$e \in \{CH_4, NH_3 \text{ and/or } H_2S\}$

$h \in \{CH_4, CO_2, O_2, N_2\}, \{O_2, N_2\} \text{ or } \{CO_2, O_2, N_2\}$

Parameters

$\mathrm{A_{specific}}$: specific clarifier area (m^2 / (ton·day))

$\mathrm{A(i)}$: Antoine A coefficient for vapor pressure of component i

$\mathrm{B(i)}$: Antoine B coefficient for vapor pressure of component i

$\mathrm{C(i)}$: Antoine C coefficient for vapor pressure of component i

$\mathrm{Cp_{sat}}$: specific heat capacity of flue gas.

d : work days per year

d_p : particle diameter (m)

k : kinetic constant (s-1)

IAP_{eq} : equilibrium ion activity product

h : work hours per day

HRT_{unit} : hydraulic retention time of *unit* (s)

$MW_{component}$: molecular weight of component (kg/kmol)

MeP_{ratio} : metal/phosphorus molar ratio in coagulation process

$Price_{component}$: price of the *component* (€/kg)

g : gravity acceleration (m2/s)

k : polytropic coefficient (1.4)

$\kappa_{agitator}$: agitators specific power consumed (HP / 1000 USgallon)

φ_j : precipitation agent *j* per total solids mass ratio

η_c : Compressor's efficiency (0.85)

η_s : Isentropic efficiency (0.9)

η_i^j : *i* component separation yield using in the process the element *j*

P_{atm} : atmospheric pressure (1 bar)

T_{atm} : atmospheric temperature (25 ºC)

R : ideal gas constant (8.314 J/mol·K)

Cp_{H_2O} : specific heat capacity of water (4.18 kJ/kg·ºC)

Variables

$a_{technology}$: parameter which takes the value 0 when $F_{design}^{technology}$ is 0 and 1 if $F_{design}^{technology}$ is not equal to 0

α_{mf} : parameter dependent of the phases number in the FBR

Ar_l : Arquimedes number for liquid

A_{unit} : area of *unit* (m^2)

$Benefits_{technology}$: benefits or losses obtained with *technology*

$C\text{-}N$: carbon to nitrogen molar ratio

C_{eq} : equilibrium concentration (kmol/m^3)

C_0 : initial concentration (kmol/m^3)

$Cost_{unit}$: cost of *unit*

$C_{component}^{unit}$: concentration of *component* in the *unit* inlet stream ($kg_{component}/kg_{total}$)

$ChemC_{technology}$: cost of chemicals for *technology*

D_{unit} : diameter of *unit*

e_{unit} : thickness of *unit*

$Ec_j(T)$: equilibrium constant of component *j* at temperature T.

$F_{component}^{unit}$:mass flow of *component* in the *unit* inlet stream (kg/s)

F_{max}^{unit} : maximum mass inlet flow admitted by a single *unit* (kg/s)

F_{design}^{unit} : mass inlet flow used in the design of *unit* (kg/s)

$FC_{technology}$: fixed cost of *technology*

$F_{total}^{\text{recovered}}$: recovered matter total mass flow (kg/s)

$F_{(unit,unit1)}$: mass flow from stream from unit to unit1 (kg/s)

$fc_{(J,unit,unit1)}$: mass flow of component J from unit to unit1 (kg/s)

$H_{b,(unit,unit1)}$: enthalpy of the stream at the state b from the stream from unit to unit1 (kJ/kg).

$H_{steam\ (isoentropy)}$: enthalpy of the stream at the if the expansion is isentropic (kJ/kg).

l_{j-i} : molar fraction of component *j* in the liquid phase of equilibrium system *i*.

K_{index} : Potassium index of fertilizer.

L_{unit} : length of *unit*

N_{am} : nitrogen contained in ammonia.

N_{org} : nitrogen contained in organic matter.

n_{unit} : number of *units* used in the process

$n_{(unit,unit1)}$: total mol flow from stream from unit to unit1 (kmol/s).

N_{index} : nitrogen index of fertilizer.

$P_{in/compressor}$: inlet pressure to compressor (bar).

$P_{out/compressor}$: outlet pressure of compressor (bar).

$P_j*(T)$: saturation pressure of pure component *j* at temperature T (bar).

P_v : vapor pressure (bar)

P_{index} : phosphorous index of fertilizer.

p_{turbi} : inlet pressure to body *i* in the turbine (bar)

P_{unit} : power of *unit*

$Q_{(unit)}$: heat exchanged in unit (kW).

$R_{C-N/k}$: carbon to nitrogen ratio in *k*.

$R_{C-N/\text{fertilizer}}$: carbon to nitrogen ratio in fertilizer.

$R_{V/F-i}$: rate of evaporation in equilibrium system *i*.

$Rest$: rest of the elements contained in the biomass.

$Re_{l\ mf}$: Reynolds number for liquid in minimum fluidization conditions

$s_{b(unit,unit1)}$: entropy the stream at the state b for the stream from unit to uni1 kJ/kg.K

$T_{turbimin}$: saturating temperature at exit of body *i* (ºC)

$T_{(unit,unit1)}$: temperature of the stream from unit to unit 1 (ºC)

$T_{bubble/i}$: bubble point temperature of equilibrium system *i* (ºC).

$T_{m/i}$: average temperature in equilibrium system *i* (ºC).

$T_{in/compressor}$: inlet temperature to compressor (ºC).

$T_{out/compressor}$: outlet temperature of compressor (ºC).

t : time (s)

u_t : terminal velocity (m/s)

u_0 : fluid velocity (m/s)

u_{mf} : minimum fluidization velocity (m/s)

v_{j-i} : molar fraction of component *j* in the vapor phase of equilibrium system *i*.

$V_{biogas,\,k}$: biogas volume produced per unit of volatile solids (VS) ($m^3_{biogás}/kg_{VS/k}$) associated to *k*.

V_{unit} : volume of *unit*

W_{unit} : weight of *unit*

$w_{DM/k}$: dry mass fraction of *k* ($kg_{DM/k}/kg$).

$w'_{VS/k}$: dry mass fraction of volatile solids out of the dry mass of *k* ($kg_{VS/k}/kg_{DM/k}$).

$w'_{C/k}$: dry mass fraction of C in k ($kg_{C/k}/kg_{DM/k}$).

$w'_{Nam/k}$: dry mass fraction of Nam in k ($kg_{Nam/k}/kg_{DM/k}$).

$w'_{Norg/k}$: dry mass fraction of Norg in k ($kg_{Norg/k}/kg_{DM/k}$).

$w'_{P/k}$: dry mass fraction of P in k ($kg_{P/k}/kg_{DM/k}$).

$w'_{K/k}$: dry mass fraction of K in k ($kg_{K/k}/kg_{DM/k}$).

$w'_{Rest/k}$: dry mass fraction of the rest of the elements contained in k ($kg_{K/k}/kg_{MS/k}$).

$W_{(unit)}$: power produced or consumed in unit (kW).

$x_{a/biogas}$: mass fraction of component *a* in the biogas

y^{j} : binary variable to evaluate the element *j*

y_{biogas} : specific saturated moisture of biogas

$Y_{a'/biogas-dry}$: molar fraction of component a in the dry biogas.

$\Delta H_{reaction|(Bioreactor)}$: Heat of the anaerobic digestion's reaction (kW).

$\Delta H_{comb}(k)$: heat of combustion of component *k* (kW).

$\Delta H_{comb}(e)$: heat of combustion of component *e* (kW).

$\Delta H_{comb}(Digestate-dry)$: heat of combustion of dry digestate (kW)

$\Delta H_f(h)_{T(unit,unit1)}$: heat of formation of component *h* at temperature $T_{(unit,unit1)}$ (kW)

Z : objective function

$\rho_{component}$: *component* density (kg/m^3)

$\mu_{component}$: viscosity of component ($kg/(m\cdot s)$)

Acknowledgments

We acknowledge funding from the National Science Foundation (under grant CBET-1604374) and MINECO (under grant DPI2015-67341-C2-1-R) and EM also acknowledges an undergraduate research grant.

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Optimal integrated facility for waste processing

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Abstract.

A mixed-integer nonlinear programming strategy is proposed to design integrated facilities to simultaneously recover power and nutrients from organic waste. The facilities consider anaerobic digestion of different types of manure (cattle, pig, poultry, and sheep). The products from this step are biogas and a nutrient-rich effluent. The biogas produced is cleaned and used in a gas turbine to produce power while the hot flue gas obtained from combustion produces steam that is fed to a steam turbine to produce additional power. The nutrient-rich effluent is processed to recover the nutrients using different technologies that include filtration, coagulation, centrifugation, and struvite precipitation in ~~mixed~~stirred and fluidized bed reactors. This processing step provides a mechanism to prevent phosphorus or nitrogen release to the environment and to avoid the development of eutrophication processes. It is found that struvite ~~r~~production~~ecovery il~~ in fluidized beds is the technology of choice to recover nutrients from all manure sources. Furthermore, power production depends strongly on manure composition and exhibits high cost variability (from 4000 €/kW in the case of poultry manure to 25000 €/kW in the case of cattle and pig manure).

~~In this work, an integrated facility for the processing of waste to power and high added value products such as fertilizers and chemicals has been conceptually designed. The process is based on the anaerobic digestion of different types of manure (cattle, pig, poultry and sheep). Specifically, the products obtained in the process are biogas composed of methane and carbon dioxide,~~ and ~~a nutrient-rich effluent. The biogas produced is cleaned and used in a gas turbine to produce power while the hot flue gas obtained from combustion is fed to a steam turbine to produce extra power. The nutrient-rich effluent can be~~is ~~processed to recover the nutrients~~s ~~using different technologies that include filtration, coagulation, centrifugation and struvite precipitation in mixed and~~

[1] Corresponding Author. M Martín: mariano.m3@usal.es

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~~fluidized bed reactors to avoid P or N accumulation on the soil. The model is formulated as a Mixed-Integer Nonlinear Programming (MINLP) for the optimal power production and nutrient recovery technology, which incorporates detailed physical and economic evaluation models. We have found that sStruvite production in fludizedfluidized beds is the technology of choice to recover nutrients from all manure sources. We have also found Furthermore, that the power production reduced strongly depends on the manure composition and exhibits high cost variability from . In particular we have found that the break-even price of the power produced ranges from 4000€/kW in case of poultry manure to 25000€/kW in the case of cattle or pig manure.~~

Keywords: Biogas, Digestate, Anaerobic digestion, Manure, Power production, Mathematical optimization

1.- Introduction

Countries across the globe generate large amounts of organic waste that include urban residues and sludge and manure from livestock activities. While many of these waste streams can be used as a source for power and chemical products, identifying suitable cost-effective technologies is challenging. Anaerobic digestion (AD) is a promising technology to treat these residues to produce biogas, which can be used as a source for thermal energy and electrical power (León and Martín, 2016) or chemicals (Hernández and Martín, 2016). However, AD technologies also generate a nutrient-rich residual stream called digestate, that must be further processed to prevent waste and soil contamination. In particular, nutrient management is needed to prevent losses of phosphorous and nitrogen to surface and underground water bodies which leads to eutrophication processes (Sampat et al., 2017, Garcia Serrano et al., 2009). There are a number of technologies that can be used to process the digestate that range from simple mechanical separations such as filters (Gustafsson et al., 2008) and centrifugation units (Meixner et al., 2015) to chemical processing such as struvite precipitation (Bhuiyan et al., 2008). Recent studies have analyzed the production of highly concentrated nutrient products such

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as struvite (Lin et al., 2015). The variability in the recovered product quality, selling price, and production cost present complex trade-offs for the optimal use of the digestate. Existing studies have only addressed the performance of various treatment mechanisms and lack a systematic design perspective that evaluates the performance of coupled biogas and nutrient recovery technologies (Drosg et al., 2015). This is necessary, for instance, to assess economic performance of nutrient recovery in the face of strong variations in the digestate content obtained from AD (Al Seadi et al., 2008).

~~Countries across the globe generate large amounts of waste that include urban residues and sludge and manure from livestock activities. While many of these waste streams can be used as source for power and chemical products, identifying suitable cost-effective technologies is challenging. Anaerobic digestion (AD) is deemed as a promising technology to treat those residues to produce biogas, which can be used as a source for energy, thermal or power (León and Martín, 2016) or chemicals (Hernández and Martín, 2016). However, AD technologies also generate a nutrient-rich residual stream called digestate, that must to be processed further to prevent waste and soil contamination. In particular, nutrient management is needed to prevent high concentration of phosphorous and nitrogen in the soils of the US (Sampat et al., 2017) and Europe (Garcia Serrano et al., 2009; Sampat et al., 2017). There are a number of technologies that can be used to process the digestate to recover phosphorous and nitrogen and with this discharge digestate more safely. Such technologies range from simple mechanical separations that remove solids and nutrients such as filters (Gustafsson et al., 2008), centrifugation (Meixner et al., 2015), the production of struvite (Bhuiyan et al., 2008). Recent reviews are available. (Lin et al., 2015; Gustafsson et al., 2008; Szabó et al., 2008; Meixner et al., 2015; Bhuiyan et al., 2008), to the production of highly concentrated products such as struvite. The variability in the recovered product quality, selling price, and production cost presents complex trade-offs for the optimal use of the digestate. Existing studies have only addressed the performance of various treatment mechanisms but lacks a systems-wide process design perspective that evaluates the performance of coupled biogas and nutrient recovery technologies (Drosg et al., 2015). This is necessary, for instance, to assess economic performance of nutrient recovery in the face of strong variations in the digestate content obtained from AD (Al Seadi et al., 2008).~~

In this work we propose a systematic design framework to optimize the simultaneous production of energy from the biogas obtained by anaerobic digestion of cattle, sheep, poultry and pig manure, along with the recovery of nitrogen and phosphorous from the digestate. The proposed framework determines the optimal

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technology configuration, equipment sizing, and operational conditions for various compositions of manure and digestate and revenues for biogas, electricity, and fertilizer.

The paper is organized as follows. In section 2 we present a brief description of the process and the flowsheet. In section 3 we focus on the modelling of the digestate processing technologies and costing. Section 4 presents the results for various feedstocks and section 5 draws conclusions.

~~In this work we proposed an process design framework to optimize the simultaneous production of energy from the biogas obtained by anaerobic digestion of manure from cattle, sheep, poultry and pig, and the recovery of nitrogen and phosphorous from digestate processing. The proposed framework determines the optimal technology configuration, equipment sizing, and operational conditions for various compositions of manure and digestate and credits for biogas, electricity and fertilizer. The paper is organized as follows. In section 2 we present a brief description of the process and the flowsheet. In section 3 we focus on the modeling~~modelling ~~of the digestate processing technologies and costing. Section 4 presents the results for various scales and feedstocks and section 5 draws conclusions.~~

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2.- Process description

The proposed process consists of four sections: biogas production, biogas purification (~~biomethane~~ biogas generation), electrical power generation, and nutrient recovery from digestate. This is illustrated in Figure 1.

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The biomass together with water and nutrients (manure slurry) are fed to a bioreactor through stream 1, where the mixture is anaerobically digested to produce biogas and a decomposed substrate (digestate). The biogas, composed of methane, carbon dioxide, nitrogen, hydrogen sulfide, ammonia and moisture leaves the bioreactor through stream 2, and it is then sent to the purification section to remove H_2S in a fixed-bed reactor and to eliminate CO_2 and traces of NH_3 in a second step by using a Pressure Swing Adsorption (PSA) system. The purified biogas (stream 3) is used in a Brayton cycle, modelled as a furnace and an expansion, producing power. Air is fed via stream 4 and the exhaust gases (stream 7) are fed to a regenerative Rankine cycle, where it produces high pressure overheated steam extracted in stream 8. This overheated steam is fed to a steam turbine, where it is expanded to produce power. The exhaust steam from the turbine is recovered in stream 9 and

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reused in the Rankine cycle through stream 10. Between streams 9 and 10 hot flue gases from the gas turbine reheat and produced overheated steam from the recycled water (León and Martín, 2016).

The digestate is released from the digester through stream 12, and it can be processed through a number of technologies to remove nitrogen and phosphorous. We consider filtration, centrifugation, coagulation, and struvite production using either a fluidized bed reactor (FBR) or a continuous stirred tank reactor (CSTR). These technologies are described in detail in Section 3.4.

Four manure types have been considered as raw material for the process: cattle, pig, poultry and sheep manure. Table 1 shows the composition and properties of each type of manure.

Table 1: Manure composition and properties (Kowalski et., 2013; Lorimor et al., 2004, Al Seadi et al., 2008; Martins das Neves et al., 2009).

~~The biomass together with water and nutrients (manure slurry) are fed to a bioreactor through stream 1, where the mixture is anaerobically digested to produce biogas and a decomposed substrate (digestate). The biogas, composed of methane, carbon dioxide, nitrogen, hydrogen sulphide, ammonia, and moisture, leaves the bioreactor through stream 2, and it is sent to the purification section to remove H_2S in a fixed-bed reactor and to eliminate CO_2 and traces of NH_3 in a second step by using a Pressure Swing Adsorption (PSA) system. The purified biogas, stream 3, is used in a Brayton cycle, modelled as a furnace and an expansion, producing power. Air is fed via stream 4. The exhaust gases, stream 7 are fed to a regenerative Rankine cycle, where it produces high pressure overheated steam extracted in stream 8. This overheated steam is fed to a turbine to be expanded producing power. The exhaust steam from the turbine is recovered in stream 9 and reused in the Rankine cycle through stream 10. Between streams 9 and 10 there is a heat exchange integration to recover the residual heat contained in the turbine outlet streams in the heating of the recycled water (León and Martín, 2016).~~

~~The digestate is evacuated from the digestor through stream 12, and it can be processed through a number of technologies to remove nitrogen and phosphorous. We consider filtration, centrifugation or coagulation, and struvite production using a fluidized bed reactor (FBR) or a continuous stirred tank reactor (CSTR). These technologies are detailed described in Section 3.4.~~

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~~Four manure types have been considered as raw material for the process: cattle, pig, poultry and sheep manures. Table 1 shows the composition and properties of each type of manure where the properties of the pig slurry come from (Kowalski et., 2013;) and the rest from (Lorimor et al., 2004, Al Seadi et al., 2008; Martins das Neves et al., 2009); Kowalski et., 2013).~~

~~Table 1: Manure composition and properties~~

Figure 1.-Flowsheet for the production of power and fertilizers

3.- Modelling issues

We evaluate the performance of the different unit operations in the process by using detailed models that comprise mass and energy balances, thermodynamics, chemical and vapor-liquid equilibria, and product yield calculations. The global process model comprises total mass flows, component mass flows, component mass fractions, temperatures and pressures of the streams in the process network. The components that are ~~tracked~~considered in our calculations belong to the set shown in Table 2.

Table 2: Set of components

~~{Wa, CO_2, CO, O_2, N_2, H_2S, NH_3, CH_4, SO_2, C, H, O, N, Norg, P, K, S, Rest, Cattle_slurry, Pig_slurry, Poultry_slurry, P_2O_5, K_2O, CaCO3, FeCl3, Antifoam, Fe2SO4_3, Al2SO4_3, AlCl3, MgCl2, NaOH, Struvite_seeds, Mg, Cl, Struvite, KStruvite, MgCl2_CSTR, NaOH_CSTR, Mg_CSTR, Cl_CSTR, Struvite_CSTR, KStruvite_CSTR, FeCl3_Coag}.~~ In the following subsections, we briefly present the main equations used to characterize the operation of the different units. For the sake of brevity, simpler balances based on removal efficiency or stoichiometry and equations connecting units are omitted. The power production system is described in detail in previous work (León and Martín, 2016) and we thus only provide a brief description.~~We evaluate the performance of the different unit operations in the process by using models that comprise mass and energy balances, thermodynamic relationships, chemical and vapor-liquid equilibria, and product yield calculations. The global process model comprises total mass flows, component mass flows, component mass fractions, temperatures and pressures of the streams in the process network. The components that are tracked in our~~

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~~calculations belong to the set {Wa, CO_2, CO, O_2, N_2, H_2S, NH_3, CH_4, SO_2, C, H, O, N, Norg, P, K, S, Rest, Cattle_slurry, Pig_slurry, Poultry_slurry, P_2O_5, K_2O, CaCO3, FeCl3, Antifoam, Fe2SO4_3, Al2SO4_3, AlCl3, MgCl2, NaOH, Struvite_seeds, Mg, Cl, Struvite, KStruvite, MgCl2_CSTR, NaOH_CSTR, Mg_CSTR, Cl_CSTR, Struvite_CSTR, KStruvite_CSTR, FeCl3_Coag}. In the following subsections we briefly present the main equations used to characterize the operation of the different units. For the sake of the length of the paper, simpler balances based on removal efficiency or stoichiometry, or the equations that connect two units are omitted and only the conversion, the chemical reactions and the removal efficiency are presented. Furthermore, the power production section was already described in a previous work (León and Martín, 2016), therefore only a brief description is presented below.~~

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The cost estimation for the alternatives and for the entire process is based on the estimation of the unit costs from different sources using the factorial method. From the units cost, the facility cost is estimated using the coefficients in Sinnot (1999), so that the total physical plant cost involving equipment erection, piping instrumentation, electrical, buildings, utilities, storages, site development, and ancillary buildings is 3.15 times the total equipment cost for processes which use fluids and solids. On the other hand, the fixed cost, which includes design and engineering, contractor's fees, and contingency items is determined as 1.4 times the total physical plant cost for the fluid and solid processes. In the subsequent cost estimation procedures these parameters are denoted as f_i for the total physical plant parameter and f_j for the fixed cost parameter.

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~~The estimation of the cost of the alternatives and that of the entire project is based on the estimation of the unit costs from different sources using the factorial method. From the units cost, the facility cost is estimated using the coefficients in Sinnot (1999) so that the total physical plant cost involving equipment erection, piping instrumentation, electrical, buildings, utilities, storages, site development and ancillary buildings, is 3.15 times the total equipment cost for processes which uses fluids and solids. On the other hand, the fixed cost, which includes design and engineering, contractor's fee and contingency items, its determined as 1.4 times the total physical plant cost for fluid and solid processes. In the subsequent cost estimation these parameters are designed as f_i for the total physical plant parameter and f_j for the fixed cost parameter.~~

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3.1.- Biogas production

AD is a complex microbiological process that decomposes organic matter in the absence of oxygen. It produces a gas mixture following hydrolysis, acidogenesis, acetogenesis, and methanogenesis steps, consisting mainly of methane and carbon dioxide (biogas), and decomposed substrate (digestate). The anaerobic reactor is modeled using mass balances of the species involved in the production of biogas and digestate. Inorganic nitrogen, organic nitrogen, sulfur, carbon, and phosphorus balances are formulated by using the composition of volatile solids in manure, see Table 1 (Al Seadi et al., 2008; Martins das Neves et al., 2009). Typical bounds for the biogas composition are provided. The reactor operates at 55 ºC. We refer the reader to the supplementary material and León and Martín (2016) for details on the modelling of the digester.

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3.2.- Biogas purification

This system consists of a number of stages to remove H_2S, CO_2 and NH_3. Here we highlight some basics about the operation of these stages. For further details we refer the reader to previous work (León and Martín, 2016).

The removal of H_2S is carried out in a bed of Fe_2O_3, that operates at 25-50 ºC producing Fe_2S_3. The regeneration of the bed uses oxygen to produce elemental sulfur and Fe_2O_3

CO_2 is adsorbed using a packed bed of zeolite 5A. The typical operating conditions for PSA systems are low temperature (25 ºC) and moderate pressure (4.5 bar). The recovery of the PSA system is assumed to be

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100% for NH_3 and H_2O (because of their low total quantities in the biogas, in general), 95 % for CO_2, and 0% for all other gas of the mixture.

In both cases the system is modelled as two beds in parallel so that one bed is in adsorption mode while the second one is in regeneration mode, to allow for continuous operation of the plant. Further details can be found in the supplementary material.

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~~In both cases the system is modelled as two beds in parallel so that one bed is in adsorption mode while the second one is in regeneration mode, to allow for continuous operation of the plant. Further details can be found in the supplementary material.~~

3.3.- Electricity generation

We consider two stages for the generation of power. The initial one consists of the use of a gas turbine, a common alternative for using any gas fuel. However, the flue gas that exits the gas turbine is at high temperature. We can either produce steam as a utility or use that steam within a regenerative Rankine cycle to enhance the production of power. The details for the process appear in León and Martín (2016) or in the supplementary material.~~We consider two stages for the generation of power. The initial one consists of the use of a gas turbine, a common alternative to use any gas fuel. However, the flue gas that exits the gas turbine is at high temperature. We can either produce steam as a utility or use that steam within a Regenerative Ranking cycle to enchanceenhance the production of power. The details for the process appear in (León and Martín, 2016) or in the supplementary material.~~

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3.3.1.- Brayton cycle

We model the Brayton cycle as a double-stage compression system (one for the air and one for the fuel) with intercooling with variable operating pressure for the gas turbine. The compression is assumed to be polytropic with a coefficient equal to 1.4 and an efficiency of 85% (Moran and Shapiro, 2003).

The combustion of methane from the biogas is assumed to be adiabatic, heating up the mixture. We consider the combustion chamber as an adiabatic furnace. We use an excess of 20% of air with respect to the stoichiometry and 100 % conversion of the reaction:

~~We model the Brayton cycle as a double multistage compression system (one for the air and one for the fuel) with intercooling with variable operating pressure for the gas turbine. The compression is assumed to be polytropic with a coefficient k equal to 1.4 with an efficiency of 85% (Moran and Shapiro, 2003)~~

~~The combustion of the methane from the biogas is assumed to be adiabatic, heating up the mixture. We consider the combustion chamber as an adiabatic furnace. We use an excess of 20 % of air with respect to the stoichiometry and 100 % conversion of the reaction:~~

$$\cancel{CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O}\; CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \qquad (1)$$

The hot flue gas is expanded in the gas turbine to generate power and the expansion is assumed polytropic. In this case, a value of 1.3 is used based on an offline simulation using CHEMCAD®, with an efficiency of 85 % (Moran and Shapiro, 2003). Finally, the exhaust gas is cooled down and used to generate high-pressure steam to be fed to the Rankine cycle.

~~The hot flue gas is expanded in the gas turbine to generate power, assumed polytrophic. In this case, a value of 1.3 is used based on an offline simulation using CHEMCAD ®, with an efficiency of 85 % (Moran and Shapiro, 2003). Finally, the exhaust gas is cooled down and used to generate high pressure steam to be fed to the Rankine cycle.~~

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3.3.2.- Rankine cycle

We use the hot flue gas from the turbine to generate steam following a scheme that consists of using the hot gas in the order that follows. First, the hot flue gas is used for the superheating stage of the steam that is to be fed to the turbine. Next, the hot gas is used in the regenerative stage of the Rankine cycle, reheating the steam from the expansion of the high pressure turbine. Subsequently, the flue gas is used in the evaporation and preheating of the condensed water, see Figure 1. The details of the modelling of the Rankine cycle can be seen in Martín and Martín (2013). We assume an isentropic efficiency of 0.9 for each expansion.

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~~We use the hot flue gas from the turbine to generate steam following a scheme that consists of using the hot gas in the order that follows. First, the hot flue gas is used for the superheating stage of the steam that is to be fed to the turbine. Next, the hot gas is used in the regenerative stage of the ranking cycle, reheating up the steam from the expansion of the high pressure turbine. Subsequently, the flue gas is used in the evaporation and preheating of the condensed water., See Figure 1. The details of the modelling of the Rankine cycle can be seen in Martín and Martín, 2013. We assume an isentropic efficiency of 0.9 of each expansion.~~

3.4.- Digestate conditioning

Four different alternatives are considered to process the digestate including filtration, centrifugation, coagulation, and struvite production. For struvite production, the performance of fluidized bed reactors (FBR) and stirred tanks reactors (CSTR) systems is compared. For filtration, centrifugation, and coagulation technologies, nutrients output is a cake composed of different solids and nutrients, with a complex composition. The credit that we can get from the cake has been estimated based on the amount of nutrients contained. The prices for the nutrients (N, P and K) are assumed as follows: 0.45€/kg for N, 0.24€/kg for K and 0.32€/kg for P (Hernández et al., 2017).

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3.4.1.-Filtration

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Filtration is a low-cost technology that is appropriate for small installations where the amount of P to be removed is moderate. This technology consists of a filter that contains a reactive medium to help remove phosphorus. P removal using reactive filtration takes place through various mechanisms depending on the characteristics of the filter media. For instance, filter media made of compounds rich in cations under basic environments (usually containing calcium silicates at pH values above 9) form orthophosphate precipitates in the form of calcium phosphates, principally as hydroxyapatite (Pratt et al., 2012). Metallurgical slag captures P by adsorption over metal at pH close to 7 (Pratt et al., 2012). In this work we consider the use of five different types of filter media. Among them, we have studied wollastonite as a filter media rich in alkaline calcium silicates, dolomite Polonite® as calcium carbonate based components, and Filtra P® as calcium hydroxide based product (Österberg, 2012; Vohla et al., 2011). For the metallurgical slag, we have considered the blast furnace slag described by Cucarella et al. (2008). These filters are used in wastewater treatment facilities (Gustafson et al., 2008) and further analysis can be found in Shilton et al. (2006). Details on Ca-rich filters can be found in Koiv et al. (2010). The removal yield of P and N for the different filter media is shown in Table 3~~2~~. It is possible to combine this filter medium with nitrogen-philic filters to simultaneously remove nitrogen and phosphorous. An advantage of this technology is that the cake produced can be used as soil fertilizer (Hylander et al., 2006). The removal yield of nitrogen for Filtra P has been considered similar to the limestone nitrogen removal yield, as Filtra P is a limestone derived product.

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~~Filtration is a low-cost technology appropriate for small installations where the amount of P to remove is moderate. This technology consists of a filter that contains a reactive medium to help remove phosphorus. P removal using reactive filtration takes place through various mechanisms depending on the medium characteristics. For instance, filter media made of compounds rich in cations under basic environments (usually containing calcium silicates at pH values above 9) form orthophosphate precipitates in the form of calcium phosphates, principally as hydroxyapatite (Pratt et al., 2012; Gustafson et al., 2008). Metallurgical slag captures P by adsorption over metal at pH close to 7 (Pratt et al., 2012). In this work we consider the use of five different types of filter media. Among them, as filter media rich in alkaline calcium silicates we have studied wollastonite, as calcium carbonate based components have been considered dolomite Polonite®, and as calcium hydroxide based product Filtra P® has been evaluated (Österberg, 2012; Vohla et al., 2011). As metallurgical slag we have~~

~~considered the blast furnace slag described by Cucarella et al. (2008) . These filters are used in waste water treatment (Gustafson et al., 2008) and further analysis is found in ; (Shilton et al., 2006) or Ca-rishrich filters (Koiv et al., 2010); Koiv et al., 2010). The removal yield of P and N for the different filter media is shown en Table 2. It is possible to combine this filter medium with nitrogen filic filters to simultaneously remove nitrogen and phosphorous. An advantage of this technology is that the cake produced can be used as soil fertilizer (Hylander et al., 2006). The removal yield of nitrogen for Filtra P has been considered similar to the limestone N removal yield due to Filtra P is a limestone derived product.~~

Table ~~2~~3. Recovered P and N yield for different filter media

The model for the filtration is based on the removal efficiency per filter media, see Figure 2. It has been considered that materials such as total solids, carbon and potassium are forming solid compounds, so they will be retained by the filter media, Eqs. (4~~3~~)-(7~~6~~)~~:~~.

$$F_i^{cake} \geq F_i^{in} \cdot \eta_i^j - M \cdot (1 - y^j)$$
$$i \in \{P, N\}$$
$$j \in \{filter\ media\} \qquad (2~~1~~)$$

$$\sum y^j = 1 \qquad (3~~2~~)$$

$$F_i^{liquid\ effluent} = F_i^{in} - F_i^{cake} \qquad (4~~3~~)$$

$$F_k^{cake} = F_k^{in} \quad k \in \{TS, C, K\} \qquad (5~~4~~)$$

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} \right) \cdot \frac{C_{Wa}^{cake}}{1 - C_{Wa}^{cake}} \qquad (6~~5~~)$$

$$F_{Wa}^{liquid\ effluent} = F_{Wa}^{in} - F_{Wa}^{cake} \qquad (7~~6~~)$$

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As we have considered five filter media, we have used a Big M formulation to select one of them assigning a binary variable $y^{filter\ media}$ for each filter media, Eqs. (2~~1~~)-(3~~2~~). This variable takes a value of 1 for the selected filter media and 0 for the rest, so that we are able to evaluate one filter media per time.

It is assumed that the cake obtained contains moisture with a value of ~~15~~55% in weight basis (C_{Wa}^{cake}).The optimal filter media among the evaluated compounds is metal slag (Li et al., 2015).

~~Here, $y^{filter\ media}$ is a binary variable to select among the different filter media. The constant M is a large number that is used to relax constraints when a unit is not installed. It is assumed that the cake obtained contains moisture with a value of 15% in weight basis (C_{Wa}^{cake}). The optimal filter media among the evaluated compounds is metal slag.~~

Figure 2.- Scheme of the filter

The cost of each alternative has been estimated according to the number of filters, which depends on the maximum flow they can process. The maximum flow per filter unit, $F_{\max}^{filter}$ ~~$F_{\max}$~~, is 1300 ft³/min (Loh et al., 2002). To design the filter units we have taken the lower value between the flow provided by mass balances and the maximum flow allowed per filter, Eqs. (8~~7~~)-(10~~9~~).~~The cost of each alternative is estimated as follows. The number of filters needed depends on the maximum flow they can process. Consequently,~~

$$F\left(\frac{ft^3}{min}\right) = \frac{F_{in}}{\rho_{digestate}} \qquad (8\cancel{7})$$

$$n_{filters} \geq \frac{F_{total}^{filter}}{F_{max}^{filter}} \qquad (9\cancel{8})$$

~~Where $n_{filters}$ is an integer number that represents the number of filters. The maximum flow per filter unit, $F_{\max}$, is 1300 ft³/min (Loh, Lyons and White, 2002). We define a design variable for the filter as:~~

$$F_{design}^{filter}\left(\frac{ft^3}{\min}\right) = \min\left(F_{\max}^{filter}, F_{total}^{filter}\right) \qquad (10\cancel{9})$$

In fact, since the maximum flow for a cartridge filter is 1300 ft³/min, for this facility the number of filters considered in this work will always be one and the design flow is equal to the flow provided by mass

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balances.~~Since the maximum flow for a cartridge filter is 1300 ft³/min (Loh et al., 2002), the number of filters considered in this work will always be one.~~

The correlation used to calculate the filter cost, Eq. (1~~0~~1), is obtained from data reported in Loh et al. (2002). This correlation provides the price in 1998 dollars, so we use the Chemical Engineering Index to update it.

~~The correlation used to calculate the filter cost is obtained from data reported in Loh et al. (2002). This correlation provides the price in 1998 dollars and we use the Chemical Engineering Index to update it to current time.~~

$$FC_{filtration}(\$) = 4.7436 \cdot F_{design}^{filter} + 807.6923 \qquad (11\cancel{0})$$

The operating cost is estimated using a simple correlation, Eq. (14~~3~~), ~~in (Vian Ortuño, 1991)~~where we ~~We~~ assume that the utilities contribute 20% of the total (Vian Ortuño, 1991). The other economical contributions considered are ~~C~~chemicals, estimated as in Eq. (12~~10~~), labour, as per Eq. (12~~3~~) and the contribution of the investment cost of the units ~~give~~given by ~~e~~Eqs. (10~~9~~)-(11~~0~~). The filter media are considered as chemicals that will be ~~repalces~~replaces annually.

$$ChemC_{filtration}\left(\frac{€}{year}\right) = \frac{\dfrac{F_P^{in} \cdot 3600 \cdot h \cdot d}{kg_P}}{kg_{\,filter\ media}} \cdot Price_{filter\ media} \qquad (12\cancel{1})$$

~~Where~~ In Eq. (12~~1~~) $kg_{filter\ media}$ are calculated as the P content in the inlet stream divided by the filter media P adsorption capacity.

$$Labour\ cost\left(\frac{€}{year}\right) = \left(61.33 \cdot F_P^{recovered} \cdot 3.6 \cdot h^{(-0.82)}\right) \cdot \left(F_P^{recovered} \cdot 3.6 \cdot h \cdot d\right) \cdot \left(\frac{Salary}{h \cdot d}\right) n_{OP} \qquad (13\cancel{2})$$

The number of operations considered, n_{OP} ~~nOP~~, is equal to 1.

$$Operating\ cost\left(\frac{€}{year}\right) = \frac{ChemC + 1.5 \cdot Labour\ cost + 0.3 \cdot FixedCost \cdot f_i \cdot f_j}{(1 - Utilities)} \qquad (13\cancel{4})$$

Finally the credit obtained from the cake is computed as the weighted sum of each nutrient value, Eq.(15~~4~~), (Hernandez et al., 2017), and the benefits (or losses) are computed as the difference between the credit obtained from the cake and the operating costs of the facility, Eq. (16~~5~~).

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~~The credit obtained from this cake is computed as the weighted sum of each nutrient (Hernandez et al., 2017):~~

$$Cost_{cake}\left(\frac{€}{year}\right)=\left(F_P^{recovered}\cdot Price_P+F_N^{recovered}\cdot Price_N+F_K^{recovered}\cdot Price_K\right)\cdot 3600\cdot h\cdot d \qquad (15~~4~~)$$

~~Finally, the benefits (or losses) are computed by the following equation:.~~

$$Benefits_{Filtration}\left(\frac{€}{year}\right)=Cost_{cake}-Operating\ cost \qquad (16~~5~~)$$

3.4.2.-Coagulation

Coagulation is a chemical treatment to process the digestate. The goal of this process is to destabilize colloidal suspensions by reducing the attractive forces, followed by a flocculation process to form flocs from the previously destabilized colloids and to subsequently precipitate them. The nutrients are then recovered with other sedimented solids by clarification. Both N and P can be removed from the influent through coagulation-flocculation, where phosphorus is removed primarily in the form of metal hydroxides, which is the dominant process at typical plant pH values (Szabó et al., 2008). Nitrogen elimination is related to the removal of the colloidal matter (Aguilar et al., 2002). Different coagulation agents are considered aiming at selecting the optimal one: ~~$FeCl_3$,~~$FeCl_3$, $Fe_2(SO_4)_3$~~ 3~~, $Al_2(SO_4)_3$, and $AlCl_3$. -The flowsheet for the process of coagulation is presented in Figure 3.

~~Coagulation is a chemical treatment to process the digestate. The goal of this process is to destabilize colloidal suspensions by reducing the attractive forces, followed by a flocculation process to form flocs from the previously destabilized colloids and to subsequently precipitate them. The nutrients are then recovered with other sedimented solids by clarification. Both N and P can be removed from the influent through coagulation-flocculation, where phosphorus is removed primarily in the form of metal hydroxides, which is the dominant process at typical plant pH values (Szabó et al., 2008). Nitrogen elimination is related to the removal of the colloidal matter (Aguilar et al., 2002). Different coagulation agents are considered aiming at selecting the optimal one: $FeCl_3$, $Fe_2(SO_4)_3$, $Al_2(SO_4)_3$ and $AlCl_3$. The flowsheet for the process of coagulation is presented in Figure 3.~~

Figure 3.- Scheme of the coagulation process.

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The removal efficiency achieved is similar for the different coagulant agents, with values up to 99% for phosphorus and 57% for nitrogen (Aguilar et al., 2002). The main variables which influence the coagulation-flocculation process are the initial ratio of metal to phosphorus, pH, and Chemical Oxygen Demand (COD). The initial metal-phosphorus molar ratio must be between 1.5 and 2.0, and the recommended pH range is from 5.5 to 7. COD has a negative impact on the removal efficiency when its value is increased (Szabó et al., 2008).

To determine the amount of coagulant agent to be added to the system, it has been considered that a metal/phosphorus molar ratio of 1.75 must be achieved (Szabó et al., 2008). Given the relationship between the P in the raw material stream, the metal added, and the metal concentration in the commercial presentation of the coagulant agent, we are able to compute the coagulant agent amount that should be added. In the coagulation and flocculation tanks the flocs are formed and nutrients are recovered in the sediment together with coagulation agents and organic solids contained in the raw material. In the decanter, it has been considered that the stream with solids has a water content of 50% ($C_{Wa}^{sedimentator}$) (Williams and Esteves, 2011) and the water content of the centrifuge outlet solids stream is 60% ($C_{Wa}^{centrifuge}$) (Wakeman, 2007).

Other elements present in the digestate, such as total solids, carbon, and potassium are assumed to be present in the solid forming compounds that sediment. Thus, they are among species that constitute the cake. Taking into account the elements mentioned above mass balances have been formulated with the corresponding removal ratios. To select and evaluate the different coagulant agents, the problem has been modelled using a mixed-integer nonlinear programming (MINLP) formulation with Big-M constraints, Eqs. (17~~6~~)-(18~~7~~)

~~The removal efficiency achieved is similar for the different coagulant agents, with values of up to 99% for phosphorus and 57% for nitrogen (Aguilar et al., 2002). The main variables which influence the coagulation flocculation process are initial ratio metal to phosphorus, pH and COD. The initial metal-phosphorus molar ratio must be between 1.5 to 2.0 and the recommending pH value range is from 5.5 to 7. COD has a negative impact on the removal efficiency when its value is increased (Szabó et al., 2008).~~

~~To determine the amount of coagulant agent that is necessary to add in the system it has been considered that a metal/phosphorus molar ratio of 1.75 must be achieved (Szabó et al., 2008). Given the relation between the P in the raw material stream and the metal added and the metal concentration in the commercial~~

~~presentation of the coagulant agent we are able to calculate the coagulant agent amount that it should be added. In the coagulation and flocculation tanks the flocs are formed and nutrients are recovered in the sediment together with coagulation agents and organic solids contained in the raw material. In the sedimentator separation it has been considered that the stream with solids has a content in water of 50% (~~$C_{Wa}^{sedimentator}$~~) (Williams and Esteves, 2011) and the water content of the centrifuge outlet solids stream is 5% (~~$C_{Wa}^{centrifuge}$~~) (Williams and Esteves, 2011; Baasel, 1977). Other elements presented in the digestate, such as total solids, carbon and potassium are considered that they are present forming solid compounds, so they will sediment and are part of the cake formed. Taking into account the elements mentioned above mass balances have been calculated as follows to evaluate the different coagulant agents:~~

$$F_j^{coag\ tank} \geq \frac{F_P^{in}}{MW_P} \cdot MeP_{ratio} \cdot \frac{MW_j}{C_{Me}} - M \cdot (1 - y^j) \quad j \in \{coagulant\ agents\}$$ (17~~6~~)

$$\sum y^j = 1$$ (18~~7~~)

Where MeP is the metal/phosphorus ratio and M is a number larger to formulate the Big-M disjunction ~~as a Big-M~~ to select and evaluate the different coagulant agents. Mass balances are computed using Eqs. (189)-(287).

$$F_j^{coag\ tank} = F_j^{floc\ tank} = F_j^{sedimentator} = F_j^{centrifuge} = F_j^{cake} \; j \in \{coagulants\}$$ (19~~8~~)

$$F_i^{in} = F_i^{coag\ tank} = F_i^{floc\ tank} = F_i^{sedimentator} \quad i \in \{P, N\}$$ (20~~19~~)

$$F_i^{cake} = F_i^{centrifuge} = F_i^{sedimentator} \cdot \eta_i^j$$ (21~~0~~)

$$F_i^{sink1} = F_i^{sedimentator} - F_i^{cntrifuge}$$ (22~~1~~)

$$F_k^{in} = F_k^{coag\ tank} = F_k^{floc\ tank} = F_k^{sedimentator} = F_k^{centrifuge} = F_k^{cake} \quad k \in \{TS, C, K\}$$ (23~~2~~)

$$F_{Wa}^{in} = F_{Wa}^{coag\ tank} = F_{Wa}^{floc\ tank} = F_{Wa}^{sedimentator}$$ (24~~3~~)

$$F_{Wa}^{centrifuge} = \left(F_{TS}^{centrifuge} + \sum_i F_i^{centrifuge} + \sum_j F_j^{centrifuge} \right) \cdot \frac{C_{Wa}^{sedimentator}}{1 - C_{Wa}^{centrifuge}}$$ (25~~4~~)

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$$F_{Wa}^{sink1} = F_{Wa}^{sedimentator} - F_{Wa}^{centrifuge}$$ (26~~5~~)

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} + \sum_j F_j^{cake} \right) \cdot \frac{C_{Wa}^{centrifuge}}{1 - C_{Wa}^{centrifuge}}$$ (27~~6~~)

$$F_{Wa}^{sink2} = F_{Wa}^{centrifuge} - F_{Wa}^{cake}$$ (28~~7~~)

The estimation of the size and cost of both the coagulation and flocculation tanks has been carried out using a correlation developed by Almena and Martin (2016) as a function of the weight of the vessels. To simplify the mass balances it is considered that the volume provided by the coagulant agents is negligible with respect to the processed stream of the digestate. The vessel size is computed from the residence time. The hydraulic retention time considered in the coagulation tank is 4 min (Zhou et al., 2008). ~~The estimation of the size and cost of both the coagulation and flocculation tanks has been carried out using a correlation developed by Almena and Martin (2016) as a function of the weight of the vessels, Eq. (33). To simplify the mass balances it is considered that the volume provided by the coagulant agents is negligible with respect to the processed stream of digestate. The hydraulic retention time considered in the coagulation tank is 4 min (Zhou et al., 2008).~~ The vessel size is computed from the residence time, Eq. (29~~8~~). Using this data~~With it~~, the diameter and length are computed using rules of thumb, Eqs. (30~~29~~)-(31~~0~~). Finally, a correlation for the thickness as a function of the diameters allows determining the mass of metal required for the vessel and its ~~weight ,~~weight, Eqs. (32~~1~~)-(33~~2~~). Vessel cost estimation is provided by Eq. (34~~3~~).

$$V_{Coag\ tank}\left(m^3\right) = HRT_{Coag\ tank} \cdot \frac{F_{digestate}^{in}}{\rho_{digestate}}$$ (29~~8~~)

$$D_{Coag\ tank}\left(m\right) = \left(\frac{6 \cdot V_{Coag\ tank}}{7 \cdot \pi}\right)^{1/3}$$ (30~~29~~)

$$L_{Coag\ tank}\left(m\right) = 4 \cdot D_{Coag\ tank}$$ (31~~0~~)

$$e_{Coag\ tank}\left(m\right) = 0.0023 + 0.003 \cdot D_{Coag\ tank}$$ (32~~1~~)

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$$W_{Coag\ tank}(kg) = \rho_{SS316}\cdot\left[\pi\cdot\left(\left(\frac{D_{Coag\ tank}}{2}+e_{Coag\ tank}\right)^2-\left(\frac{D_{Prec\ tank}}{2}\right)^2\right)\cdot L_{Coag\ tank}+\frac{4}{3}\cdot\pi\cdot\left(\left(\frac{D_{Coag\ tank}}{2}+e_{Coag\ tank}\right)^3-\left(\frac{D_{Coag\ tank}}{2}\right)^3\right)\right] \quad (33~~2~~)$$

$$Cost_{Vessel}=6839.8\cdot V_{Coag\ tank}\left(m^3\right)^{0.65} \quad (34~~3~~)$$

To estimate the power consumed by the agitator, Eq. (35~~4~~), the rules of thumb have been used where the specific power consumed, $\kappa_{agitator}$, is tabulated in Walas (1990). For our slurries a value of $\kappa_{agitator}$ equal to 10 HP per 1000 US gallons is the most appropriate.

~~To estimate power consumed by the agitator, the rules of thumb in Walas (1990) have been used.~~

$$P_{agitator}(HP)=V_{Coag\ tank}(US\ gallon)\cdot\frac{\kappa_{agitator}}{1000} \quad (35~~4~~)$$

The agitator cost is also estimated using a correlation from Walas (1990), Eq. (36~~5~~). For cost estimation purposes we have considered stainless steel 316 as construction material and a dual impeller operating at speed between 56 and 100 rpm depending on the tanks size. With this considerations the values for a, b and c are 8.8200, 0.1235 and 0.0818 respectively (Walas, 1990). This correlation provides the cost in 1985 dollars, so it is necessary to update the result using the Chemical Engineering Index as before.

~~where $\kappa_{agitator}$ is the agitator specific power consumed in HP per 1000 US gallons, whose value is tabulated in Walas (1990). For our slurries a value of $\kappa_{agitator}$ equal to 10 HP per 1000 US gallons is the most appropriate. The cost correlation in Eq (35) is provided 1985 dollars, where parameters a, b and c are a function of the agitator material, the impeller speed and the impeller configuration (single impeller or dual impeller). We have considered stainless steel 316, dual impeller and speed between 56 and 100 rpm based on the tanks size. With this considerations the values for a, b and c are 8.8200, 0.1235 and 0.0818 respectivelly~~respectively ~~(Walas, 1990). We use the Chemical Engineering Index to update it as before.~~

$$Cost_{agitator\ 1985}(\$)=e^{a+b\cdot ln\left(P_{agitator}(HP)\right)+c\cdot\left[ln\left(P_{agitator}(HP)\right)\right]^2} \quad (36~~5~~)$$

The total cost of the coagulation tank is equal to the sum of the vessel cost and the agitator cost, Eq. (37~~6~~).

$$Cost_{Coag\ tank} = Cost_{Vessel} + Cost_{agitator\ 2016}$$ (37~~6~~)

The flocculation tank is designed similarly to that of the coagulation, using Eqs. (29~~8~~)-(37~~6~~4). For this step the hydraulic retention time is 25 min (Zhou et al., 2008).

The decanter is assumed to be circular because of its lower operating and maintenance costs. The area, Eq. (38~~7~~), is computed using the parameter $A_{specific}$, which is the specific clarifier area in m² per ton of inlet flow per day (WEF, 2005). The typical value, 10 m²/(t/day), is taken from Perry and Green (2008). The diameter of the clarifier, $D_{clarifier}$, is computed from the area value, Eq. (39~~8~~).~~The sedimentator is assumed to be circular because of its lower operating and maintenance costs. The area is computed using Eq. (37) (WEF, 2005)~~

$$A_{clarifier} = \frac{A_{specific} \cdot F^{in}_{digestate} \left(\frac{m^3}{day}\right)}{1000}$$ (38~~7~~)

~~Where $A_{specific}$ is the specific clarifier area in m² per ton of inlet flow per day. The typical value, 10 m²/(t/day), is taken from Perry and Green (2008). The diameter of the clarifier is computed from the area as follows:~~

$$D_{clarifier} = \left(\frac{4 \cdot A_{clarifier}}{\pi}\right)^{1/2}$$ (39~~8~~)

The number of clarifiers is an integer value that has been computed rounding up the ratio between the clarifier diameter calculated before and the maximum clarifier diameter, $D^{clarifier}_{max}$, Eq. (40~~39~~). The maximum clarifier diameter value considered is 40 m (Perry and Green, 2008). ~~The number of clarifiers is computed as an integer value as follows:~~

$$n_{clarifiers} \geq \frac{D^{clarifier}_{total}}{D^{clarifier}_{max}}$$ (~~39~~40)

~~where $D^{clarifier}_{max}$, the maximum clarifier diameter, is 40 m (Perry and Green, 2008).~~

The diameter used in the final design will be the smallest between $D_{clarifier}$ and $D^{clarifier}_{max}$, Eq. (~~40~~41).

$$D^{clarifier}_{design} = min(D^{clarifier}_{max}, D^{clarifier}_{total})$$ (~~40~~41)

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To model the minimization function and compute $D_{clarifier\ design}$, the following smooth function approximation, given by Eq. (~~41~~42), is used based on previous work (de la Cruz and Martin, 2016), to avoid discontinuities within the problem formulation.

~~To model the min function and compute $D_{clarifier\ design}$, the following smooth function approximation, given by eql Eq. (41), based on previous work is proposed to avoid discontinuities within the problem formulation (de la Cruz and Martin, 2016).~~

$$D_{design}^{clarifier}=\frac{D_{max}^{clarifier}}{1+e^{\left(-F_{digestate}^{in}+0.342\right)\cdot 2.718}} \qquad (\cancel{41}42)$$

The cost estimation correlation has been developed from the data in WEF (2005), Eq. (4~~2~~3). It includes all the items involved in the operation of such an unit. The correlation must be updated to current prices using the Chemical Engineering Index.

~~It includes all the items involved in the operation of such an unit. The correlation must be updated to current prices using Chemical Engineering Index:~~

$$Cost_{clarifier\ 1979}=\left(13060\cdot D_{design}^{clarifier}-58763\right)\cdot n_{clarifiers} \qquad (4\cancel{2}3)$$

Centrifuge sizing and costing is based on the data by Perry and Green (2008). We assume pusher type with a maximum diameter of 1250 mm. ~~Centrifuge sizing and costing is based on the data by Perry and Green (2008). We assumed pusher type with a maximum diameter of 1250 mm.~~ The modelling equation~~s for sizing~~ for sizing is ~~and costing are~~ given ~~below:~~ in Eq. (4~~3~~4)

$$D_{Centrifuge}\left(in\right)=0.3308\cdot\frac{F_{digestate}^{in}}{1000}\cdot 3600+9.5092 \qquad (4\cancel{3}4)$$

The number of centrifuges is calculated taking into account the maximum centrifuge diameter, Eq. (4~~4~~5), and the diameter used in the final design will be the minimum value between $D_{Centrifuge}$ and $D_{Centrifuge\ max}$ Eq. (4~~5~~6).

$$n_{centrifuges}\geq\frac{D_{total}^{centrifuge}}{D_{max}^{centrifuge}} \qquad (4\cancel{4}5)$$

~~Where $n_{centrifuges}$ is an integer number and the maximum centrifuge diameter, $D_{Centrifuge\ max}$, is 49.21 in (Perry and Green, 2008)1250mm as indicated before.~~

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~~The diameter used in the final design will be the smallest between $D_{Centrifuge}$ and $D_{Centrifuge\ max}$.~~

$$D_{design}^{centrifuge} = \min\left(D_{max}^{centrifuge}, D_{total}^{centrifuge}\right) \qquad (4\cancel{5}6)$$

As in the clarifier, we develop a smooth approximation, Eq. (47~~6~~), to compute the design diameter avoiding discontinuities as follows:

~~As in the clarifier, we use a smooth approximation to compute the design diameter avoiding discontinuities as follows:~~

$$D_{design}^{centrifuge} = \frac{D_{max}^{centrifuge}}{1+e^{\left(-F_{digestate}^{in}+35.369\right)\cdot 0.0395}} \qquad (47\cancel{6})$$

Thus, the cost for the centrifuge is estimated based on the data by Perry and Green (2008) as a function of its diameter, Eq. (48~~7~~). Since the cost correlation is based on 2004 values, the Chemical Engineering Index it used to update the equipment cost. ~~Thus, the cost for the centrifugecentrifuge is estimated as in Eq. (47). Since the cost correlation is based on 2004 data, the Chemical Engineering Index it used to update the equipment cost.~~

$$Cost_{centrifuge\ 2004}(\$) = \left(10272\cdot D_{design}^{centrifuge} - 24512\right)\cdot n_{centrifuges} \qquad (48\cancel{7})$$

We estimate the operating cost of this system by accounting for the annualized equipment cost (fixed cost), chemicals and labor cost. A similar procedure as before is followed Vian Ortuño (1991~~) but~~) but for the clarifier fixed costs as the correlation to estimate its costs already includes the operating cost, Eq. (4~~8~~9).

~~We estimate the operating costs of this system accounting for the fixed, the chemicals and labour cost. We estimate these items using Vian Ortuño (1991) as presetedpresented above, but for the clarifier fixed costs due to the fact that the correlation to estimate the clarifier costs already includes their operational cost.~~

$$FC_{coagulation}\left(\frac{€}{year}\right) = \left(Cost_{Coag\ tank} + Cost_{Floc\ tank} + Cost_{centrifuge\ 2016}\right)\cdot f_i \cdot f_j + Cost_{clarifier\ 2016} \qquad (4\cancel{9}8)$$

The chemicals costs are estimated as Eq. (50~~49~~)

$$ChemC_{coagulation}\left(\frac{€}{year}\right) = \left(F_{Fe_2(SO_4)_3}^{in}\cdot Price_{Fe_2(SO_4)_3} + F_{Al_2(SO_4)_3}^{in}\cdot Price_{Al_2(SO_4)_3} + F_{FeCl_3}^{in}\cdot Price_{FeCl_3} + F_{AlCl_3}^{in}\cdot Price_{AlCl_3}\right)3600\cdot h\cdot d \qquad (\cancel{49}50)$$

To estimate the price for the cake, as in the previous case, we assume the price of each of the nutrients contained (N, P, and K). The price for each nutrient is taken same as before. Thus, the cake price is computed as the weighted sum of each nutrient, as in Eq. (~~14~~15) (Hernandez et al., 2017).

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Finally, the economic benefits or losses of operating this system are calculated as the difference between the credit obtained from the cake and the operating costs of the section of the facility, as in Eq. (~~15~~16).

~~While the labor costs are estimated as Eq. (12) and the operating cost as in Eq (13) where n_{OP} is equal to one. To estimate the price for the cake, as in the previous case, we assume the price of each of the nutrients contained in it, N, P and K . The price for each is taken as before. Thus, the cake price is computed as the weighted sum of each nutrient as in Eq. (14) (Hernandez et al., 2017).~~

~~Finally, the economic benefits or losses the benefits of operating this system are calculated using Eq. (15), applied to this case.~~

3.4.3.-Centrifugation

Centrifugation is a pretreatment that separates solid and liquid phases and that can be used to recover nutrients from the digestate. The advantage of this system is the simple equipment used. Precipitant agents can be added to improve the removal efficiency significantly. Previous studies show that an appropriate mixture of $CaCO_3$ and $FeCl_3$ promotes nutrients recovery. In particular, a ratio of 0.61 kg $CaCO_3$ per kilogram of total solids in the raw material inlet stream, and 0.44 kg of $FeCl_3$ per kilogram of total solids in the raw material inlet stream, achieves a removal efficiency up to 95% and 47 % for P and N respectively (Meixner et al., 2015). Figure 4 presents a scheme of the process.

~~Centrifugation is a pretreatment that separates solid and liquid phases and that can be used to recover nutrients from the digestate. The advantage of this system is the simple equipment used. Precipitant agents can be added to improve significantly the removal efficiency. Previous studies show that an appropriate mix of $CaCO_3$ and $FeCl_3$ promotes nutrients recovery. In particular, proportions of in proportion of 0.61 kg $CaCO_3$ per kilogram of total solids in the raw material inlet stream and 0.44 kg of $FeCl_3$ per kilogram of total solids in the raw material inlet stream, achieving a removal efficiency up to 95% and 47 % for P and N respectively (Meixner et al., 2015). Figure 4 presents a scheme of the process.~~

Figure 4.-Scheme for the centrifugation treatment

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Centrifugation process consists of two units, a precipitation tank where $CaCO_3$ and $FeCl_3$ are added, and the centrifuge. These equipment have been modeled using mass balances and removal ratios for the precipitating species. Note that the total solids, carbon, and potassium are assumed to be present in the form of solid compounds, so they will be removed as part of the cake. Moreover, the water content of the centrifuge outlet solids stream is assumed to be 60% ($C_{Wa}^{centrifuge}$) (Wakeman, 2007). Mass balances for the process have been evaluated ~~as follows~~in Eqs. (51~~0~~)-(59~~8~~):

$$F_j^{in} = F_{TS}^{in} \cdot \frac{\varphi_j}{C_j} \quad j \in \{precipitation\ agents\} \qquad (51\cancel{0})$$

$$F_j^{in} = F_j^{prec\ tank} = F_j^{centrifuge} = F_j^{cake} \qquad (52\cancel{1})$$

$$F_i^{in} = F_i^{prec\ tank} = F_i^{centrifuge} \quad i \in \{P,N\} \qquad (53\cancel{2})$$

$$F_i^{cake} = F_i^{centrifuge} \cdot \eta_i \qquad (54\cancel{3})$$

$$F_i^{liquid\ effluent} = F_i^{centrifuge} - F_i^{cake} \qquad (55\cancel{4})$$

$$F_k^{in} = F_k^{prec\ tank} = F_k^{centrifuge} = F_k^{cake} \quad k \in \{TS,C,K\} \qquad (56\cancel{5})$$

$$F_{Wa}^{in} = F_{Wa}^{prec\ tank} = F_{Wa}^{centrifuge} \qquad (57\cancel{6})$$

$$F_{Wa}^{cake} = \left(F_{TS}^{cake} + \sum_i F_i^{cake} + \sum_j F_j^{cake} \right) \cdot \frac{C_{Wa}^{centrifuge}}{1 - C_{Wa}^{centrifuge}} \qquad (58\cancel{7})$$

$$F_{Wa}^{liquid\ effluent} = F_{Wa}^{centrifuge} - F_{Wa}^{cake} \qquad (59\cancel{8})$$

Where φ_j is the precipitation agent per total solids mass ratio (0.61 kg $CaCO_3$/ kilogram TS and 0.44 kg $FeCl_3$ / kilogram TS).~~ We should point out a couple considerations with regard to the materials such as total solids, carbon, and potassium. First, these materials are assumed to be present in form of solid compounds, so they will be removed as part of the cake. Moreover, the water content of the centrifuge outlet solids stream is assumed to be 560% ($C_{Wa}^{centrifuge}$) (BaaselWakeman, 19772007).~~

~~To model design this process we have used some correlations to design the units as a function of the flow through the unit. For the design of the precipitation tank (volume, diameter, length, thickness, weight and cost calculations) the equations provided by Almena and Martin (2016) have been used as before, Eqs. (28)-(36).~~

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~~considering an hydraulic retention time of 2.5 min (Szabó et al., 2008):~~These~~is~~ units ha~~ves~~ been designed using correlations as a function of the flow processed. For the design of the precipitation tank (volume, diameter, length, thickness, weight, and cost calculations) the equations provided by Almena and Martin (2016) have been used as before, Eqs. (2~~9~~8)-(3~~7~~6), considering a hydraulic retention time of 2.5 min (Szabó et al., 2008).

$$V_{Prec\ tank}\left(m^3\right) = HRT_{Prec\ tank} \cdot \left(\frac{F^{in}_{digestate}}{\rho_{digestate}} + \frac{F^{in}_{FeCl_3}}{\rho_{FeCl_3}} \right) \qquad (~~59~~60)$$

~~The volume of $CaCO_3$ added is considered negligible compared to the volume of the liquid. Thus, the diameter of the tanks is computed using Eqs. (28)-(29) as in the previous unit. The cost of the vessel is given by the weight of the metal, this we compute it using Eqs. (31)-(36) as before. The power requires is computed using Eq (34) where $\kappa_{agitator}$ is taken to be 10 (Walas, 1990). The cost correlation is given by Eq. (35) and updated to 2016 prices as before. The total cost of the precipitation tank involved the vessel and the agitator, Eq. (36):~~The volume of $CaCO_3$ added is assumed negligible compared to the volume of the liquid because of it is added as solid. Thus, the diameter of the tanks is computed using Eqs. (60~~59~~28) and Eq. ~~(~~(29~~30~~) as in the previous unit. The cost of the vessel is given by the weight of the metal, using the correlations provided by Almena and Martin (2016), Eqs. (3~~1~~3031)-(36~~343~~). The power required is computed, as in previous cases, using the rules of thumb in Walas (1990), Eq. (3~~5~~4), where the value of $\kappa_{agitator}$ is equal to 10 HP per 1000 gal, in accordance with the data collected in the literature (Walas, 1990). The cost correlation is given by Eq. (3~~6~~5) and updated to 2016 prices. The total cost of the precipitation tank included the vessel and the agitator costs, Eq. (3~~7~~6).

The centrifuge size is characterized by its diameter. We model it as in the previous technology using Eqs. (4~~3~~4)-(4~~8~~76).

The operating costs involve fixed, chemicals and labour costs. Fixed costs are estimated using Eq. (61~~60~~47). The labor cost is estimated in Eq. (1~~3~~2), where n_{OP} ~~n_{OP}~~ is equal to ~~1~~ (1 (Vian Ortuño, ~~(~~1991). Total operating cost is given by Eq. (1~~4~~3).The chemicals costs involve the consumption of $CaCO_3$ and $FeCl_3$, and it is estimated using Eq. (6~~2~~1):

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$$FC_{centrifugation}\left(\frac{€}{year}\right)=\left(Cost_{centrifuge\ 2016}+Cost_{Prec\ tank}\right)\cdot f_i\cdot f_j$$ (6~~0~~1)

$$ChemC_{centrifugation}\left(\frac{€}{year}\right)=\left(F^{in}_{CaCO_3}\cdot Price_{CaCO_3}+F^{in}_{FeCl_3}\cdot Price_{FeCl_3}\right)\cdot 3600\cdot h\cdot d$$ (62~~1)~~

~~The cake recovered is the main asset of the process. The credit obtained is computed using Eq. (14) as before. Finally the benefits (or losses) are computed using Eq. (15) applied to this case~~.

The cake recovered is the main asset of the process. Its price is estimated as the weighted sum of each nutrient, Eq. (1~~4~~5), (Hernandez et al., 2017). Finally, the benefits or losses of operating this system are calculated as the difference between the revenue obtained from the cake and the operating costs of the facility, Eq. (1~~5~~6).

3.4.4.-Struvite production

P and N can be recovered from digestate through the formation of struvite, which is a phosphate mineral with a chemical formula of $MgNH_4PO_4{\cdot}6H_2O$. The advantage of this technology is that struvite is a solid with a high nutrients density, it is easy to transport, and it can be used as slow-release fertilizer without any post-processing (Doyle and Parsons, 2002). The removal of nutrients via struvite production follow the reaction below, requiring the addition of $MgCl_2$, resulting in the production of struvite crystals that can be recovered as solid:

~~Phosphorus and nitrogen can be recovered from digestate through the formation of struvite, which is a phosphate mineral with a chemical formula of~~ $\cancel{MgNH_4PO_4{\cdot}6H_2O}$~~. The great advantage of this technology is that struvite is a solid with a high nutrients density, easy to transport, and that can be used as slow-release fertilizer without any post-processing (Doyle and Parsons, 2002). The removal of nutrients via struvite production follows the reaction below, requiring the addition of MgCl2, resulting in the production of struvite crystals that can be recovered as a solid:~~

$$Mg^{2+}+NH_4^{+}+H_nPO_4^{3-n}+6H_2O\Leftrightarrow MgNH_4PO_4{\cdot}6H_2O+nH^{+}$$ (63)

Due to the presence of potassium in the digestate, together with struvite, another product called potassium struvite or K-Struvite, is also produced. In this case the ammonia cation is substituted by the potassium cation (Wilsenach et al., 2007).

~~Due to the presence of potassium in the digestate, together with struvite, a variety called potassium struvite, or K-Struvite, is also produced. In this case the ammonia cation is substituted by the potassium cation (Wilsenac et al., 2007).~~

$$K^{+} + Mg^{2+} + H_nPO_4^{3-n} + 6H_2O \Leftrightarrow KMgPO_4{\cdot}6H_2O + nH^{+} \qquad (64)$$

Since the formation of struvite is favored over the formation of K-Struvite, it is considered that only 15% of the potassium contained in the digestate will react to form K-Struvite (Zeng and Li, 2006). The mass balance for the reactors is given by the stoichiometry of the reactions above.

Two different types of reactors can be used to obtain struvite, either a stirred tank (CSTR) or a fluidized bed reactor (FBR). Figures 5 and 6 provide detailed flowsheets of each case. In case of the FBR, struvite is recovered from the bottoms and the liquid must be processed in a hydrocyclone to avoid discharging fines. In the case of CSTR tanks, we need to use a centrifuge to recover the struvite. We can help the crystal growth by seeding (Doyle and Parsons, 2002; Kumashiro et al., 2001). Due to the substantial increase in the struvite formation yield, we consider the addition of struvite seeds in both cases. The reaction takes place at about 27ºC, with the addition of $MgCl_2$ at a concentration of 57.5 mg/dm^3 (Zhang et al., 2014). A Mg:P molar ratio of 2 (Bhuiyan et al., 2008) is used.

~~Since the formation of struvite is favored over the formation of K-Struvite, it is considered that only a 15% of the potassium contained in the digestate will react to form K-Struvite (Zeng and Li, 2006). The mass balance to the reactors is given by the stoichiometry of the reactions above.~~

~~Two different types of reactors can be used to obtain struvite, either a CSTR or a fluidized bed reactor (FBR). Figures 5 and 6 provide detailed flowsheets of each case. In the case of FBR, struvite is recovered from the bottoms and the liquid must be processed in a hydrocyclon to avoid dirchangingdischarging fines. In the case of CSTR tanks, we need to use a centrifuge to recover the struvite. We can help the crystal growing procedure by seeding or not (Doyle and Parsons, 2002; Kumashiro et al., 2001). Due to the substantial increase in the struvite formation yield, we consider the addition of struvite seeds in both cases. The reaction takes place at about 27ºC, adding as a source of magnesium $MgCl_2$ with a concentration of 57.5 mg/dm^3 (Zhang et al., 2014). It is used a Mg:P molar ratio of 2 (Bhuiyan et al., 2008)~~

Figure 5.- Scheme for the FBR system

The FBR system is composed of three elements: a mixer tank, a FBR rector, and a hydrocyclone. The system operation consists of a digestate flow which is mixed with a stream of $MgCl_2$ in the mixing tank. The addition of $MgCl_2$ helps precipitate the struvite by increasing the concentration of the species inside the reactor.

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As the concentration of NH_4^+ is high due to the pH, and the inorganic N and P are the elements we want to recover, the only element which is necessary add is Mg in form of $MgCl_2$.~~The FBR system is composed of three elements: a mixer tank, a FBR rector, and a hydrocyclone. The system operation consists of a digestate flow which is mixed with a stream of $MgCl_2$ in the mixer tank. The addition of $MgCl_2$ helps precipitate the struvite by rising the concentration of the species inside the reactor. As the concentration of NH^{4+} is high and P is the element which we want to recover, the only element which is necessary add is Mg in form of $MgCl_2$.~~

In the tank there is a suspension of struvite seeds with a size of 0.8 mm which promote the precipitation of struvite. The solid struvite is evacuated from the reactor at the bottom and its moisture is low enough to avoid the use of a dryer. The other stream which leaves the reactor contains liquid water in a high proportion with the excess of Mg, the total solids from the digestate, and low amounts of nutrients and other components. This stream is introduced in a hydrocyclone to recover fines of struvite which can be removed by this stream. 100% of fines removal is assumed but no fines production is ~~considered in~~considered in the ~~model .~~model.

~~In the tank there is a suspension of struvite seeds with a size of 0.8 mm which promoted the precipitation of struvite on them. The solid struvite is evacuated from the reactor by the bottom and its moisture is low enough to avoid the use of a dryer. The other stream which leaves the reactor contains liquid water in a high proportion with the excess of Mg, the total solids from the digestate and low amounts of nutrients and other components. This stream is introduced in a hydrocyclone to recover fines of struvite which can be dragged by this stream. 100% of fines removal is assumed but no fines production is considered in the model .~~

To estimate the cost of this system we evaluate the effect of the following variables, whose operating values are shown between parenthesis: ~~To estimate the cost of this system we evaluate the effect of the following variables, whose operating values are shown between parenthesis:~~

- Digestate input mass and volume flow (between 1 and 100 kg/s)
- Recovered struvite humidity (5% in mass)
- Amount of phosphorus recovered (90%)
- Mg:P molar ratio with a value of 2

- ~~Digestate input mass and volume flow (between 1 and 100 kg/s)~~
- ~~Recovered struvite humidity (5% in mass)~~

- ~~Amount of phosphorus recovered (90%)~~
- ~~Mg:P molar ratio with a value of 2~~

In an FBR there are some variables which influence in the design and hence the cost. The variables considered in this work are showed below with the typical values used in the present study between parenthesis:

- d_p: bed particle diameter, assumed to be 0.8 mm (Jordaan, 2011)
- Sphericity: 0.6 is a standard sphericity for particles used in fluidized bed reactors (Fogler, 2005)

- ~~d_p: bed particle diameter, assumed to be 0.8 mm (Jordaan, 2011)~~
- ~~Sphericity: 0.6 is a standard sphericity for particles used in fluidized bed reactors (Fogler, 2005)~~

Furthermore, the reaction kinetics and equilibrium are considered to estimate the residence time in the reactor. A first order kinetics, developed by Nelson et al. (2003), has been used, Eqs. (64~~2~~)-(65~~3~~). The kinetic constant is 3.42 $x10^{-3}$ s^{-1} for a pH of 9.~~Furthermore, the reaction kinetics and equilibrium are considered to estimate the residence time on the reactor. A first order kinetics, developed by Nelson et al. (2003), has been used. The kinetic constant is $3.42 \cdot 10^{-3}$ s^{-1} for a pH of 9.~~

$$\frac{-dC}{dt} = k\left(C - C_{eq}\right) \qquad (64\cancel{2})$$

$$\ln\left(C - C_{eq}\right) = -kt + \ln\left(C_0 - C_{eq}\right) \qquad (65\cancel{3})$$

Struvite formation is an equilibrium reaction. We use the equilibrium ion activity product (IAP_{eq}) value of $7.08 \cdot 10^{-14}$ (Nelson et al., 2003) to calculate the equilibrium concentrations in the kinetic model, Eq. (664). We ~~have~~ assumed that the values of ions concentration are equal to ions activity.

~~Struvite formation is an equilibrium, which equilibrium ion activity product (IAP_{eq}) value used to calculate the equilibrium concentrations in the kinetic model is $7.08 \cdot 10^{-14}$ (Nelson et al., 2003). As approximation ions concentrations has been taken equal to ion concentrations.~~

$$IAP_{eq} = \left(Mg^{2+}\right)\left(NH_4^{+}\right)\left(PO_4^{3-}\right) = 7.08 \cdot 10^{-14} \qquad (664\cancel{3})$$

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Minimum fluidization velocity is calculated in the first step by considering that the fluid stream is a liquid ~~Minimum fluidization velocity is calculated in a first step considering that the fluid is a liquid~~ (Mangin and Klein, 2004). This consideration is motivated because the liquid digestate works as fluidization agent ~~This consideration is motivated because the own liquid digestate works as fluidizant agent~~ (~~Mangin and Klein, 2004;~~ Le Corre, 2006). The digestate density is 950 kg/m³ (Rigby and Smith, 2011). The expression used to calculate u_{mf} ~~umf~~ through Reynolds and Archimedes numbers is given by Eq. (6475), (Tisa et al., 2014).

$$u_{mf} = \frac{\mathrm{Re}_{l\,mf} \cdot \mu_{digestate}}{\rho_{digestate} - d_p} \qquad (6457)$$

Eq. (675) parameters are determined by Eq. (686) and Eq. (697) ~~.~~). ~~Where~~

$$\mathrm{Re}_{l\,mf} = \sqrt{33.72 + 0.0404 Ar_l \left(1-\alpha_{mf}\right)^3} - 33.7 \qquad (6865)$$

$$Ar_l = \rho_{digestate}\left(\rho_{struvite} - \rho_{digestate}\right) g \frac{d_p^3}{\mu_{digestate}^2} \qquad (6976)$$

If the flow has no gas phase, α_{mf} is equal to cero. The terminal velocity is computed using Eq. (70687) (Tisa et al., 2014).

~~Tisa et al (2014).~~

$$u_t = \left(\frac{1.78 \cdot 10^{-2} \cdot \eta^2}{\rho_{digestate} \cdot \mu_{digestate}}\right)^{1/3} d_p$$

(70687)

Where the parameter η ~~ETA~~ is given by Eq. (6971)

$$\eta = g\left(\rho_{struvite} - \rho_{digestate}\right)$$

(71698)

Finally, the fluid velocity u_0 must be between u_{mf} ~~umf~~ and u_t ~~ut~~. A superficial velocity equal to five times the minimum fluidization velocity is selected ~~chosen~~ (Tejero-Ezpeleta et al., 2004), Eqs. (697072)-(73710)

$$u_{mf} < u_0 < u_t$$

(697072)

$$u_0 = 5 \cdot u_{mf}$$

(73710)

Once the superficial velocity is computed~~, the area and diameter can be calculated for the processing flow as follows:~~, the area and diameter can be calculated from the mass flow Eqs. (7274)-(7375).

$$A_{FBR} = \frac{F_{digestate}^{in}}{u_0}$$

(74721)

$$D_{FBR} = \sqrt{\frac{4A_{FBR}}{\pi}}$$

(75732)

The length of the bed is determined by the residence time through the kinetics and the equilibrium ion activity product presented above. Consequently, the magnesium and ammonium -concentrations can be calculated from the digestate mass balance and the external magnesium added. Using ~~, and knowing~~ the IAP_{eq} value, the phosphate concentration in equilibrium ~~in~~ at the operational conditions can be determined. This equilibrium value will be used in kinetics, Eq. (76743).

$$t = \frac{\ln\left(C_0 - C_{eq}\right) - \ln\left(C - C_{eq}\right)}{k} \qquad (7643)$$

Thus, the bed length is computed as per Eq. (74577). Typically, the length of the reactor must be 15% larger than the bed, Eq. (78765) ~~.~~).

$$L_{bed} = \frac{u_0}{t}$$

(74577))

~~Typically, the length of the reactor must be 15% larger than the bed.~~

$$L_{FBR} = 1.15 \cdot L_{bed}$$

(78765)

The estimation of the rector ~~cost of the reactor~~cost is carried out assuming that it is a vessel as presented in the processes above, ~~presented above in~~ Eeqs.(2930)-(3334)~~.~~, (Almena and Martin, ~~(~~2016). The cost of the mixer tank is also estimated as that of a vessel , using Eeqs. (2829)-(3334), with a volume given by that to provide a hydraulic retention time of 150 s (Szabó et al., 2008). The impeller is also designed using the same procedure as before, Eqs. (~~34~~35)-(~~35~~36), (Wallas, ~~(~~1990).

Finally, to estimate the cost of the hydrocyclone, a surrogate model using data from Matche's website has been developed (www.matche.com). There is a maximum diameter, therefore, if a unit larger than the standard is required, we actually need to duplicate the equipment, Eq. (~~78~~80). To estimate the diameter, we considered that there is a linear relationship between the diameter and the flow based on rules of thumb in design literature. A typical unit size of a 20 inch diameter hydrocyclone can process 1000 US gallons per minute, Eq. (~~77~~79) (Walas, 1990).

~~A typical unit size is that a hydrocyclone with a diameter of 20 inch can process 1000 US gallons per minute (Walas, 1990).~~

$$D_{hydrocyclone}(in)=F_{digestate}^{in}\left(\frac{US\ gallon}{min}\right)\cdot\frac{20}{1000} \quad (7~~6~~97)$$

$$n_{hydrocyclone}\geq\frac{D_{total}^{hydrocyclone}}{D_{max}^{hydrocyclone}}$$

(80~~77~~8)

Where ~~n_{HC}~~ $n_{hydrocyclone}$ is an integer. The maximum diameter for a hydrocyclone, $D_{max}^{hydrocyclone}$, is 30 inch based on standard sizes (www.matche.com). Thus, the design diameter is the lower diameter between $D_{total}^{hydrocyclone}$ and $D_{max}^{hydrocyclone}$, Eq. (81~~79~~). ~~If a diameter larger than the maximum is required, the unit must be duplicated.~~

$$D_{design}^{hydrocyclone}=\min(D_{total}^{hydrocyclone},D_{max}^{hydrocyclone})$$

(~~78~~981)

The estimation of the cost for the fines recovery equipment is computed using Eq. (82~~079~~) and updated as explained above.

$$Cost_{hydrocyclone\ 2014}=n_{hydrocyclone}\cdot\left(2953.2\cdot D_{design}^{hydrocyclone}-34131\right)$$

(~~798~~082)

The CSTR process consists of four elements: the CSTR reactor, a centrifuge, and a dryer with its corresponding heat exchanger. As the residence time in the CSTR is large enough, it is not necessary to use a

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mixing tank and $MgCl_2$ is added directly in the reactor. Thus, struvite is formed in one step in the CSTR. Since the digestate already contains NH_4^+ and P, we need to add $MgCl_2$. As a result, struvite precipitates, and it is recovered from the bottoms of the reactor and dried in a two step process. The first step consists of a centrifuge that recovers struvite with 5% (on weight basis) water (Baasel, 1977). Next, a drum dryer is implemented to remove the residual moisture to reach commercial standards and reduce transportation costs. Figure 6 shows the details of the flowsheet.

~~The CSTR process consists of four elements: the CSTR reactor, a centrifuge and a dryer with its corresponding heat exchanger. As the residence time in the CSTR is large enough, it is not necessary to use a previous mixer tank and $MgCl_2$ is added directly in the reactor. Thus, struvite is formed in one step in the CSTR. Since the digestate already contains NH^{4+} and P, we need to add $MgCl_2$. As a result, struvite precipitates, and it is recovered from the bottoms of the reactor and dried in a two step process. The first step consists of a centrifuge that recovers struvite with 5%wt water~~ (Baasel, 1977)~~. Next, a drum dryer is implemented to remove the residual moisture to reach commercial standards and reduce transportation costs. Figure 6 shows the detail of the flowsheet.~~

Figure 6.-Scheme for the CSTR based struvite production system

The design of the units involved in this process and their cost estimation is based on the following variables:

- Digestate input mass and volume flow (between 1 and 100 kg/s)
- Recovered struvite water content (5% in mass)
- Amount of phosphorus recovered (90%)
- Mg:P molar ratio with a value of 2

~~The design of the units involved in this process and its cost estimation is based on the following variables:~~

- ~~Digestate input mass and volume flow (between 1 and 100 kg/s)~~
- ~~Recovered struvite humidity (5% in mass)~~
- ~~Amount of phosphorus recovered (90%)~~
- ~~Mg:P molar ratio with a value of 2~~

The CSTR is assumed to be a stirred vessel; consequently, it is designed as in the previous cases, Eqs. (289)-(3764), with a residence of 471.05 s~~, as before, Eqs. (28-34) with a residence of 471.05s~~. The residence time is calculated from mass balances and the kinetics described in the FBR process, Eq.(642) and Eq. (653).

The centrifuge size is characterized by its diameter. Both, the size and cost are computed using the data in Perry and Green (2008). We assume a pusher type for the centrifuge with a maximum diameter of 1250 mm ~~that is modelled~~ as before, Eqs. (443)-(4876).

The cost estimation for the drier relies on the amount of water to evaporate, and the evaporation capacity. The evaporation capacity ($e_{capacity}$) is reported in the literature to be equal to 0.01897 (kg/(s·m²)) (Walas, 1990)~~.)~~. Consequently, the dryer cost is computed using a correlation provided by Martin and Grossmann (2011), Eq. (8183), updating the cost to current prices using the Chemical Engineering Index.

~~, a factor provided in the literature, $e_{capacity}$, equal to 0.01897 (Walas, 1990). Consequently, the dryer cost is computed using Eq. (80), updating the cost to current prices using the chemical index:~~

$$Cost_{dryer\ 2007}(\$) = 1.15 \cdot \left(6477.1 \cdot \frac{F^{in}_{water}}{e_{capacity}} + 102394 \right)$$

(83810)

The operating cost of the CSTR and the FBR based processes is computed considering three items, fixed, chemicals and labor, and assuming that utilities account for 20% of the operating costs. The correlations for computing each of them are taken from Vian Ortuño (1991) and Sinnot (1999)~~ as before~~, Eqs. (132) for labour and (143) for ~~the~~ total operating cost. Fixed cost for struvite processes is calculated using Eq. (8284)~~ where~~. We assume that the seeds required for the FBR process are internally produced in the startup of the facility. ~~:~~

$$FC_{struvite}\left(\frac{€}{year}\right) = \left(\sum Cost_{equipment}\right) \cdot f_i \cdot f_j \qquad (81284)$$

The revenue obtained from the struvite is determined assuming a selling price of 0.763 €/kg, Eq. (8385), (Molinos-Senante et al., 2015) .

~~We assume that the seeds required for the FBR process are internally produced in the startup of the facility.~~

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~~The credit obtained from the struvite is determined assuming a selling price of 0.763 €/kg (Molinos-Senante et al., 2015) .~~

$$Cost_{struvite}\left(\frac{€}{year}\right)=\left(F_{struvite}^{recovered}\cdot Price_{struvite}\right)\cdot 3600\cdot h\cdot d \qquad (85~~83~~2)$$

Finally, the benefits or losses for CSTR and FBR are calculated as the difference between the credit obtained from the struvite and the operating costs of the facility, Eq. (16~~5~~).

~~Finally the benefits (or losses) of CSTR and FBR processes are calculated using a similar expresion~~expression ~~than Eq. 15~~

3.5.- Solution procedure

The detailed models for each of the alternatives such as the five filter media or the number of different coagulants result in a large and complex MINLP when cost estimation is involved. We use a two-stage procedure to select the best technology. In the first stage we develop MINLP subproblems to select the appropriate filter media or coagulant. Next, using the detailed models for the best option, surrogate cost models are developed for the five alternative technologies used to process the digestate. However, there are still binary decisions to account for the cost of the active alternative in the superstructure. Thus, the surrogate models are in the form of linear equations. For instance, the surrogate model for the filter to be implemented in the superstructure is given by a linear function as given ~~by Eq~~by Eq. (~~84~~86).

~~The detailed models for a each of the alternatives such as the five filter media or the number of different coagulants result in a large and complex MINLP when cost estimation is involved. We use a two-stage procedure to select the best technology. In a first stage we develop MINLP subproblems to select the appropriate filter media or coagulant. Next, using the detailed models for the best option, surrogate cost models are developed for the five alternative technologies used to process the digestate. However, they are still integer variables to to include the cost of the active alternative in the superstructure. Thus, the surrogate models are in the form given by Eq. (82). For instance, the surrogate model for the filter to be implement in the superstructure is given by a linear function as given by Eq. (82).~~

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$$Operating\ cost\left(\frac{€}{year}\right) = 20521 \cdot F_{design}\left(\frac{ft^3}{\min}\right) - 33488 \cdot a_{Filter}$$

(82486)

We avoid the use of binary variables within the formulation (due to highly non linear model of the entire superstructure) by using smooth approximations. We define a_{Filter} as a parameter that takes a value of 0 when F_{design}^{Filter} is 0 and 1 if F_{design}^{Filter} is not equal to 0. The smooth approximation for a_{Filter} is defined as follows, Eq. (8587):

~~We avoid the use of binary variables within the formulation (due to highly non inearlinear model of the entire superstucturesuperstructure) by using smooth approximations. We define a_{Filter} as a parameter that takes the value 0 when F_{design}^{Filter} is 0 and 1 if F_{design}^{Filter} is not equal to 0. The smooth approximation for a_{Filter} is defined as follows:~~

$$a_{Filter} = \frac{1}{1+e^{(-F_{design}+0.049) \cdot 361}}$$

(87853)

Metal slag is selected as the best filter for the filtration process. For the case of the ~~coagulants~~ ,coagulants, the solution of the subproblem, Eqs. (176)-(5049) -selects the use of $AlCl_3$. As in the previous case, a surrogate model is developed to be included in the superstructure so that we avoid including binary variables and allow for zero operating costs in case this technology is not selected, Eq. (88).

$$Operating\ cost\left(\frac{€}{year}\right) = 1019589.91 \cdot F_{digestate}^{in}\left(\frac{kg}{s}\right) - 368838.56 \cdot a_{Coag}$$

(88864)

Where the smooth ~~approaximation~~approximation for the term a_{Coag} is given by Eq. (85789)

$$a_{Coag} = \frac{1}{1+e^{(-F_{digestate}^{in}+0.068) \cdot 863}}$$

(89875)

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Similar to previous cases we develop a surrogate model to estimate the operating cost for the centrifugation as a function of the flowrate of digestate,~~Similar to previous cases we develop a surrogate model to estimate the operating cost for the centrifugation as a function of the flowrate of digestate,~~ Eq. (90886):

$$Operating\ cost\left(\frac{€}{year}\right) = 458498.29 \cdot F_{digestate}^{in} + 24924.67 \cdot a_{Centrifugation}$$

(86890)

As before, $a_{Centrifugation}$ is ~~approxiamted~~approximated as follows, Eq. (8991):

$$a_{Centrifugation} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.068\right)\cdot 863}}$$

(87991)

Finally, to include the operating costs for the production of struvite, we again develop ~~again we develop~~ surrogate ~~models for~~models for the FBR, Eq. (929088) and for the CSRT Eq. (90294), where a smooth approximation is proposed for the fixed term, a_{FBR} and a_{CSTR} respectively, Eqs. (899193) and (95934)

$$Operating\ Cost_{FBR}\left(\frac{€}{year}\right) = 245008 \cdot F_{digestate}^{in} + 1\cdot 10^{6} \cdot a_{FBR}$$

(88920)

$$a_{FBR} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.06785\right)\cdot 862.9679}}$$

(89931)

$$Operating\ cost_{CSTR}\left(\frac{€}{year}\right) = 277051 \cdot F_{digestate}^{in} + 1\cdot 10^{6} \cdot a_{CSTR} \quad (9042)$$

$$a_{CSTR} = \frac{1}{1+e^{\left(-F_{digestate}^{in}+0.06785\right)\cdot 862.9679}} \quad (9531)$$

The benefits/losses in the superstructure for any of the technologies to process the digestate is computed as the difference between the revenue obtained from the nutrients and generated power, and the operating costs of the facility.

Finally, the whole superstructure is built (see Figure 1). This superstructure contains models of the fermenter, biogas purification, gas cycle, ~~water~~steam cycle, and digestate treatment processes. The aim of this

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superstructure is to determine the optimal operating conditions and to select the best digestate treatment technology. Thus, digestate treatment processes have been implemented in the superstructure through detailed mass balances including the solution to the kinetics of the fluidized bed reactors as well as the surrogate models developed in the previous stage to estimate the operating costs. It should be noted that in filtration, centrifugation, and coagulation processes we have included a benefits penalty, $F_{total}^{recovered}$, due to the fact that the product recovered is a mixture of nutrients and organic matter with a nutrients concentration lower than struvite. This penalty represents the concentration of nutrients in the recovered product given by the ratio between the nutrients recovered and the total recovered mass flow, Eq. (~~94~~96).

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~~The benefits/losses in the superstructure for any of the technologies to process the digestate is compuredcomputed as per Eq. (15).~~

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~~Finally, the whole superstructure is built (see Figure 1). This superstructure contains models of the fermenter, biogas purification, gas cycle, water cycle and digestate treatment processes. The aim of this superstructure is to determine the optimal operating conditions and to select the best digestate treatment technology. To do this, digestate treatment processes have been implemented in the superstructure through the detailed mass balance including the solution to the kinetics of the fluidized bed reactors as well as the surrogate models developed in the previous stage to estimate the operating costs. It should be noted that in filtration, centrifugation and coagulation processes it has been added a benefits penalty,~~ $\cancel{F_{total}^{recovered}}$~~, due to the fact that the product recovered is a mixture of nutrients and organic matter with a nutrients concentration lower than struvite. This penalty represents the concentration of nutrients in the recovered product given by the ration between the nutrients recovered and the total recovered mass flow, Eq. 8192.~~

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$$\cancel{\text{Price}\left(\frac{€}{year}\right)=\left(F_P^{\text{recovered}}\cdot\text{Price}_P+F_N^{\text{recovered}}\cdot\text{Price}_N+F_K^{\text{recovered}}\cdot\text{Price}_K\right)\cdot\frac{1}{F_{total}^{\text{recovered}}}\cdot 3600\cdot h\cdot d}$$

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$$\text{Price}\left(\frac{€}{year}\right)=\left(F_P^{\text{recovered}}\cdot\text{Price}_P+F_N^{\text{recovered}}\cdot\text{Price}_N+F_K^{\text{recovered}}\cdot\text{Price}_K\right)\cdot\frac{1}{F_{total}^{\text{recovered}}}\cdot 3600\cdot h\cdot d$$

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(~~924~~96)

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The total energy obtained in the system to be optimized is the sum of the one generated at the three sections of the turbine, high, medium and low pressure and that of the gas turbine. We use part of the energy produced to power the compressors used across the facility. The economic benefits or losses of each digestate treatment process are added to the energy benefits.

~~The total energy obtained in the system to be optimized is the sum of the one generated at the three bodies of the turbine, high, medium and low pressure and that of the gas turbine. We use part of the energy produce to power the compressors used across the faciliity~~facility~~. To the energy benefits are added the economic benefits or losses of each digestate treatment process.~~

$$Z = \left[\left(\sum_{i \in turbinebody} W_{(\mathrm{Turbine})} + W_{(GasTurb)} - \sum_{j \in compressors} W_{(compressors)} \right) \cdot 3600 \cdot h \cdot d \cdot C_{\mathrm{Electricity}} \right] + \mathrm{Benefits}_{\mathrm{Filtration}} + \mathrm{Benefits}_{\mathrm{Centrif}} + \mathrm{Benefits}_{\mathrm{Coagulation}} + \mathrm{Benefits}_{\mathrm{FBR}} + \mathrm{Benefits}_{\mathrm{CSTR}}$$

(~~8393~~597)

Eq. (~~95~~97) is the objective function that we maximize to determine the optimal operational conditions and to select the best digestate treatment process subject to the following constraints:

~~This is the objective function which we search miximize~~maximize ~~to determine the optimal operational conditions and select the best digestate treatment process subject to the following constraints:~~

a) Bioreactor and biogas composition model
b) Digestate processing
c) Biogas purification. Described in section 3.2
d) ~~Bryaton~~Brayton cycle. Described in section 3.3.1
e) Rankine cycle. Described in section 3.3.2

The main decision variables are related to the selection of the digestate processing technology, among filtration, centrifugation, coagulation and~~,~~ struvite production using CSTR or FBR. The decision variables are also associated with the selection of the type of filter and the coagulation agent. Furthermore, the biogas usage to produce steam requires the operating pressures and temperatures at the gas turbine, and the steam turbine as well as the extraction form the steam turbine to reheat the condensate before regenerating steam using the flue gas from the gas turbine. The superstructure consists of an NLP of approximately 4000 equations and 5000 variables solved using a ~~multistrat~~multistart procedure with CONOPT 3.0 as the preferred solver. The computational time is around 6~~2~~0 min, although it varies for each problem as a consequence of the different data used in each case.

~~The main decision variables are related to the selctionselection of the digestate processing, either filtration, the type of filter, centribugationcentrifugation, coagulation, the coagulation agent, or struvite production, using a~~

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~~CSTR or an FBR. Furthermore, the biogas usage to produce steam requires the operating pressures and temperatures at the gas turbine, and the steam turbine as well as the extrationextraction form the steam turbimeturbine to reheat up the condensate before regenerating steam using the hour flue gas from the gas turbine. The superstructure consists of an NLP of approximatellyapproximately 4000 eqs equations and 5000 var variables solved using a multistrat procedure with CONOPT 3.0 as the preferedpreferred solver.~~

4.-Results

Following the optimization procedure presented in section 3.4 we first decide on the filter media and the coagulant chemical. We solve MINLP subproblems leading to the selection of the filter media and the coagulant agent. We use the metal slag as the filter media and the $AlCl_3$ as the coagulant for all raw materials. Next, we developed surrogate models for the five technologies included in the superstructure and solve a reformulated NLP including smooth approximations for the cost functions of the digestate treatment so as to maximize the power produced and the treatment section. The plant size is assumed to be that which processes 10 kg/s of manure based on the typical amount of manure produced in cattle farms (León, 2015). Four manures have been evaluated on the plant: cattle, pig, poultry and sheep, with the aim of determining, for each one, the power generated the composition of the biogas produced, the optimal digestate treatment technology to recover its nutrients and the biogas-manure and digestate-manure ratios. Section 4.1 summarizes the main operating conditions of the major units in the process and the selection of digestate processing technology. Section 4.2 presents the detail economic evaluation of the four optimal processes, one per manure type. Finally, in section 4.3 an analysis of the effect of the manure composition on the power, operating conditions and digestate treatment is performed.

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4.1.-Mass and energy balances

Table 4~~3~~ shows the main operating conditions of major units for the four different manure types. Cattle, pig, and poultry show similar values among them and to previous work (León and Martín, 2016). The gas in the gas turbine reaches a temperature of 2400 °C and a pressure of 8.2 bar before expansion for cattle, pig and poultry manure. However, sheep manure shows different values. While the temperature is similar, the pressure is 15.6 bar, almost twice the value found for the rest of the raw materials. Furthermore the flue gas exits the turbine 300 °C below that when the rest of the manure types are used. Furthermore while the high pressure of the steam turbine is 125 bar for cattle, pig, and poultry manure, in case of sheep manure the steam turbine operates at 95 bar at the high pressure section of the turbine. This is related to the lower gas temperature from the gas turbine since the overheated steam needs to be produced using that stream. Intermediate and low pressures are the same in the steam turbine using any of the manure types, but the exhaust pressure of the steam is higher in case of sheep manure. Table 4~~5~~ shows the products obtained from the ~~vapours~~various manure types, power, biogas, and digestate. Poultry is the waste that is more efficient towards power production due to its higher concentration. In all cases an FBR reactor for the production of struvite is the selected technology to recover N and P. In the table we also see the effect of the fact that cattle and pig manure are mostly liquids, since most of the product is digestate, almost 98%, while the use of poultry or sheep manure reduces the production of digestate to 75% and 88% respectively, increasing the production of biogas and power. Finally in Table 5~~6~~ the biogas composition for each manure considered are presented. The main purpose of the facility is the production of power. However, the biogas composition is typically within a range of values per component that have been imposed as bounds. As a result of maximizing the electricity production for all studied cases, the same biogas composition is obtained, 67.5% molar in CH_4 and the rest is mostly CO_2.

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~~Table 3 shows the main operating conditions of major units for the four different manure types. Cattle, pig and poultry show similar values among them and to previous work (León and Martín, 2016). The gas in the gas turbine reaches a value of 2400 ºC and 8.2 bar before expansion for cattle, pig and poultry manure. However, sheep manure shows different values. While the temperature is similar, the pressure is 15.6 bar, almost twice the value found for the rest of the raw materials. Furthermore the flue gas exits the turbine 300 ºC below that when the rest of the manure types are used. Furthermore while the high pressure of the steam turbine is 125 bar for cattle, pig and poultry, sheep manure steam turbine operate at 95 bar at the high pressure section of the turbine. The fact is related to the lower gas temperature from the gas turbine since the overheated steam needs to be produced using that stream. Intermediate and low pressures are the same in the steam turbine using any of the manure types, but the exhaust pressure of the steam is higher in case of using sheep manure. Table 4 shows the products obtained from the vapours manure types, power, biogas and digestate. Poultry is the waste that is more efficient towards power production due to its higher concentration. In all cases an FBR reactor for the production of struvite is the selected technology to recover N and P. In the table we also see the effect of the fact that cattle and pig manure are mostly liquids, since most of the product is digestate, almost 98%, while the use of poultry or Sheep manure reduces the production of digestate to 75 and 88% respectively, increasing the production of biogas and power. Finally in Table 5 the biogas composition for each manure considered are presented. As the considered plant purpose considered is the power generation, the biogas composition has been delimited within a range of possible values for each content which set a high methane content. As a result of maximizing the electricity production a for all studied cases, the same biogas composition is obtained, 67.5% molar in CH_4 and the rest mostly CO_2.~~

Table 4~~3~~.- Operating data of the optimal configuration for each raw material.

Table 5~~4~~.- Process optimization results for considered manures

Table 6~~5~~.- Biogas composition for considered manures

4.2.-Economic evaluation

This section is divided into the estimation of the investment cost, using a factorial method based on the cost of the units, and the estimation of the electricity production cost.~~This section is divided into the estimation of the investment cost, using a factorial method based on the cost of the units, and the estimation of the electricity production cost.~~

4.2.1.- Investment cost

We use the factorial method to estimate the investment cost for this facility. This is based on the estimation of the equipment cost and several coefficients to account for pipes, installation, etc. (Sinnot and Towler, 2009). The cost for the different units has been estimated based on Matche's website (www.matche.com), Sinnot and Towler (2009) and Peters and Timmerhaus (2003), updating the cost of the units when required. We assume a plant that processes fluids and solids. Due to the different composition of each manure the specific production of biogas for each one is different, being larger for poultry and sheep than for cattle and pig. The reason for that could be that sheep and poultry manures have less water content while the water content in cattle and pig reaches 98% (http://adlib.everysite.co.uk). For cost estimation proposes the digester maximum size considered is 6000 m^3 per unit, since the larger units could face mixing and homogenization problems (FNR, 2010). This result for the facility investment cost will be different for each raw material. Figure 7 shows the equipment cost distribution where digest~~o~~er and gas turbine are the most important contributions:

- Cattle manure: A plant that processes 10 kg/s of this type of manure requires an investment of 69.1 M€, of which 14.9 M€ represents the equipment cost. The larger cost is assumed by the digester units, with a 75% of the total units cost, followed by the heat exchanger network with a contribution of 12% while both turbines add up to 12%.
- Pig manure: A facility to process 10kg/s of this manure requires in an investment of 69.5 M€, with a cost of 14.9 M€ ~~in equipment~~in equipment. Since the digestate-manure and biogas-manure ratios between cattle and pig manure are very similar, the investment costs are analogous among them. The unit cost distribution is similar to the cattle manure case.
- Poultry manure: The investment for a plant which processes 10kg/s of this manure ~~is 208.0~~is 208.0 M€. The units investment adds up to 44.7 M€. In this the units cost distribution is more

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homogeneous among different items: 60% to digest~~o~~er units, 20% to gas turbine, 10% to heat exchanger network and 9% to steam turbine. It should be noted that, as poultry manure has a high content of dry matter (around 60% on a weight basis), it is necessary to add additional water to decrease the dry matter content to reach 25% with the aim of avoid mixing problems in the digester due to an excessive solids concentration inside.

- Sheep manure: The facility to treat 10kg/s of this manure requires an investment of 105.0 M€, where 22.5 M€ represents the equipment cost. For this plant the main units cost distribution is as follows: 50% for the ~~digestor~~digester, 25% for gas turbine, 17% for heat exchanger network and 7% for steam turbine.

~~We use the factorial method to estimate the investment cost for this facility. This is based on the estimation of the equipment cost and several coefficients to account for pipes, installation, etc.(Sinnot and Towler, 2009). The cost for the different units has been estimated based on MATCHE, Sinnot and Towler (2009) and Peters and Timmerhaus (2003) updating the cost of the units when required. We assume a plant that processes fluids and solids. Due to the different composition of each manure the specific production of biogas for each one is different, being larger for poultry and sheep than for cattle and pig. A reason for that could be that sheep and poultry manures have a less content in water while moisture content in cattle and pig reach 98% (http://adlib.everysite.co.uk/adlib/defra/content.aspx?id=2RRVTHNXTS.88UETAXXR0DD2). For cost estimation proposes the digester maximum size considered is 6000 m³ per unit, due to larger units could be mixing and homogenization problems (FNR, 2010). This results in that facility investment cost will be different for each raw material. Figure 7 shows the equipment cost distribution where digestor and gas turbine are the most important contributions.~~

- ~~Cattle manure: a plant that processes 10 kg/s of this type of manure requires an investment of 69.1 M€, of which 14.9 M€ represents the equipment cost. The larger cost is assumed by the digestor units, with a 75% of the total units cost, followed by the heat exchanger network with a contribution of 12% while both turbines add up to 12%.~~
- ~~Pig manure: a facility to process 10kg/s of this manure results in an investment of 69.5 M€, with a cost of 14.9 M€ in equipment. Due to the digestate-manure and biogas-manure ratios~~

~~between cattle and pig manure are very similar, the investment cost are analogous among them. The unit cost distribution is similar to the cattle manure case.~~

- ~~Poultry manure: the investment for a plant which process 10kg/s of this manure is 208.0 M€. The units investment adds up to 44.7 M€. In this the units cost distribution is more homogeneous among different items: 60% to digestor units, 20% to gas turbine, 10% to heat exchanger network and 9% to steam turbine. It should be noted that, as poultry manure have a high content in dry matter (around 60 % in weight basis), it is necessary to add additional water to decrease the dry matter content to reach 25% with the aim of avoid mixing problems in the digestor due to an excessive solids concentration inside.~~
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It is clear that the digester shows the highest share in the investment cost and therefore the concentration of the manure highly determines the cost of the facility. Lantz (2012) presented the investment cost of a facility for heat and power production as a function of its scale. Actually, our plant does not produce steam as a final product but only power. Thus, it is interesting to see that the raw material determines the investment per kW from the 4000€/kW in case of poultry manure or the 7500€/kW in case of sheep manure, to the more than 25000 €/kW in case of pig and cattle. ~~It is clear that the digestor shows the highest share in the investment and therefore the concentration of the manure highly determines the cost of the facility. Lantz (2012) presented the investment cost of a facility for heat and power production as a function of its scale. Actually, our plant does not produce steam as a final product but only power. Thus, it is interesting to see that the raw material determines the investment per kW from the 4000€/kW of the poultry manure or the 7500€/kW of the sheep manure, to the more than 25000 €/kW in case of pig and cattle.~~

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Figure 7: Units cost distributions for cattle, pig, poultry and sheep manure treatment (ST: Steam turbine, GT: Gas turbine, HX: Heat exchangers, FBR: Fluidized bed reactor).~~Units cost distributions for cattle, pig, poultry~~

~~and sheep manure treatment (ST: Steam turbine, GT: Gas turbine, HX: Heat exchangers, FBR: Fluidized bed reactor).~~

4.2.2.- Production cost

To calculate the production cost, 20 years of plant life is considered, with a capacity factor of 98%. Apart from the equipment amortization, other items are also taken into account such as salaries, administrative fees, chemicals cost, maintenance cost, utilities and contingency costs. Thus, apart from the annualized equipment cost, 1.5 M€ are spent in Salaries, 0.25 M€ in Administration~~, 2M~~, 2M€ in Maintenance, 0.25 M€ in other expenses (Martín and León, 2016) while chemicals are computed as described in section 2. The cost of utilities adds up to 0.08 M€, accounting for the cooling water and the steam needed to maintain the operation of the digester and to condition the digestate for its use as a fertilizer. Finally, we assume that the livestock manure is for free. Figure 8 shows the distribution of the production costs for each of the manure types. We see that the figures are very similar. The equipment amortization represents at least 43% of the production costs. This share increases up to 60% for the case of the use of poultry. As the investment is lower, the annual cost for other items is almost constant and their contribution to the electricity cost plays a more important role. Chemicals is the second most important contribution to the cost of electricity with a share of up to 23% for the use of cattle or pig manure and down to 16% in the case of sheep manure. We assume in all cases that waste is for free. Under these considerations the electricity production costs obtained are presented in Table ~~6~~7.

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Table 7~~6~~: Electricity production cost and NPV for the facility considering different raw materials

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The Net Profit Value has also been calculated as a measure of the project profitability, considering an electricity price of sale of 0.06 €/kWh. To compare the profitability of this project a secure investment as the inversion in Spanish national debt has been chosen, considering a discount rate of 3% (Ministerio de Economía, Industria y Competitividad, 2017). The results obtained are presented in Table ~~6~~7, and it should be noted that facilities for poultry and sheep manures obtain positive NPV while those which use cattle and pig manure as raw material show negative NPV, so from the point of view of NPV as an indicator to decide the project viability, those ones would be disregarded.

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Figure 8: Operation cost distribution for cattle, pig, poultry and sheep manure treatment

4.3.- Effect on the power, operating conditions and digestate treatment

The results obtained from the treatment of different manure streams show the influence of the manure composition in the amount produced and the composition of biogas and digestate obtained. Struvite production using FBR is the best choice for digestate treatment. This can be explained by the advantages in recovering nutrients in solid form since they can be easily transported and stored. Furthermore the material is highly concentrated in nutrients with a relatively high selling price.

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Biogas production is similar for cattle and pig manures, but is significantly higher in the poultry and sheep cases. The investment cost when processing cattle and pig ~~manure is~~manure is dominated by the ~~digestor~~digester, resulting in similar investment and production costs for facilities using either of the two types of manure. However, the higher concentration in organic matter in sheep and poultry manure does not only results in higher power production capacities, but the fact that the contribution to the cost of the turbines is also larger and so is the investment cost of these facilities. On the other hand, the electricity production cost is lower in the last two cases as result of the economies of scale between the investment cost and the biogas produced and the higher amount of struvite produced, with the extreme case of poultry manure where the struvite selling benefits are capable to cover the electricity production costs. Note that the availability of poultry and or sheep manure should be less that than for cattle and pig manure.

~~Biogas production is similar for cattle and pig manures, but is significantly higher in the poultry and sheep cases. The high share of the cost due to the turbines results in the fact that the investment for the facilities processing cattle and pig manures are similar and their value between all manures evaluated is the lowest. However, the higher production capacity of power from sheep, and specially for poultry, results in the need for larger turbines and higher investment cost. On the other hand, the electricity production cost is lower in the two last cases as result of the scale economies between the investment cost and the biogas produced and the higher amount of struvite produced, with the extreme case of poultry manure where the struvite selling benefits are capable to cover the electricity production costs.~~

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~~4.-Results~~

~~Following the optimization procedure presented in section 3.4 we first decide on the filter media and the coagulant chemical. We solve MINLP subproblems leading to the selection of the metal slag We use the metal slag as filter media and the $AlCl_3$ as coagulant for all raw materials. Next, we developed surrogate models for the five technologies included in the superstructure and solve a reformulated NLP including smooth approximations for the cost functions of the digestate treatment so as to maximize the power produced and the treatment section. The plant size is assumed to be that which processes 10 kg/s of manure based on the typical amount of manure produced in cattle farms [31]. Four manures have been evaluated on the plant: cattle, pig, poultry and sheep, with the aim of determining, for each one, the power generated the composition of the biogas produced, the optimal digestate treatment technology to recover its nutrients and the biogas-manure and digestate-manure ratios. Section 4.1 summarizes the main operating condition of the major units in the process and the selection of digestate processing technology. Section 4.2 presents the detail economic evaluation of the four optimal processes, one per manure type. Finally, in section 4.3 an analysis of the effect of the manure composition on the power, operating conditions and digestate treatment is performed.~~

~~4.1.-Mass and energy balances~~

~~Table 3 shows the main operating conditions of major units for the four different manure types. Cattle , pig and poultry show similar values among them and to previous work (León and Martín, 2016). However, Sheep manure shows different values. An expansion of 8.2 bar is suggested for the gas turbine but for the sheep manure, whose expansion is almost twice. Furthermore while the high pressure of the steam turbine is 125 bar for cattle, pig and poultry, sheep manure steam turbine operate at 95 bar at the high pressure section of the turbine. The fact is related to the lower gas temperature from the gas turbine. Table 4 shows the products obtained from the vapours manure types, power, biogas and digestate. Poultry is the waste that is more efficient due to its higher concentration. Finally in Table 5 the biogas composition for each manure considered are presented. As the considered plant purpose considered is the power generation, the biogas composition has been delimited within a range of possible values for each content which set a high methane content. As result of maximize the electricity production a for all studied cases it is selected the same high biogas composition.~~

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~~Table 3.- Operating data of the optimal configuration for each raw material.~~

~~Table 4.- Process optimization results for considered manures~~

~~Table 5.- Biogas composition for considered manures~~

~~4.2.-Economic evaluation~~

~~This section is divided into the estimation of the investment cost, using a factorial method based on the cost of the units, and the estimation of the electricity production cost.~~

~~4.2.1.- Investment cost~~

~~We use the factorial method to estimate the investment cost for this facility. This is based on the estimation of the equipment cost and several coefficients to account for pipes, installation, etc.(Sinnot and Towler, 2009). The cost for the different units has been estimated based on MATCHE, Sinnot and Towler (2009) and Peters and Timmerhaus (2003) updating the cost of the units when required. We assume a plant that processes fluids and solids. Due to the different composition of each manure the specific production of biogas for each one is different, being larger for poultry and sheep than for cattle and pig. A reason for that could be that sheep and poultry manures have a less content in water while moisture content in cattle and pig reach 98% (http://adlib.everysite.co.uk/adlib/defra/content.aspx?id=2RRVTHNXTS.88UETAXXR0DD2). For cost estimation proposes the digester maximum size considered is 6000 m^3 per unit, due to larger units could be mixing and homogenization problems (FNR, 2010). This results in that facility investment cost will be different for each raw material. Figure 7 shows the equipment cost distribution where digestor and gas turbine are the most important contributions.~~

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~~4.2.2.- Production cost~~

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5.-Conclusions

In this work, we have designed optimal integrated facilities for the production of biogas-based electrical power and fertilizers from manure. Detailed equation based models for the anaerobic digestion, the Brayton and regenerative Rankine cycles and different technologies for digestate treatment have been developed. To solve the model a two-step procedure has been performed. First, the individual detailed models for each digestate treatment technology are used to formulate a MINLP model aiming at selecting the best configuration for that technology: the best precipitation agent, filter media, etc. In the second step, the best configuration of each technology has been implemented in the entire superstructure. Due to the fact that only one digestate processing technology is allowed and the highly non-linear nature of the model, surrogate models for the cost of each alternatives with a smooth approximations have been developed. For the optimal selection a detailed economic evaluation is performed.

The results show that FBR technologies are preferred to recovery nutrients. Furthermore, in some cases this process can produce electricity at a competitive price (in case of poultry and sheep manure). The investment cost is highly dependent on the water and organic content of the manure type, ranging from 70M€ to 208M€ when a large energy production is possible and large gas and steam turbines are to be installed. However, for these cases of high investment cost, the production cost of power is the most competitive due to the large production capacity. Biogas power plants show a wide range of values of power per kW installed depending on the manure concentration. Competitive values of 4000€/kW for poultry manure are obtained, due to the highly concentrated manure, while large values of 25000€/kW installed are reported in case of the diluted cattle or pig manure.

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Nomenclature

Sets

$i \in \{P, N\}$

$j \in \{filter\ media\}$

$k \in \{TS, C, K\}$

$a' \in \{CH_4, CO_2, NH_3, H_2S, O_2\ and/or\ N_2\}$

$a \in \{H_2O, CH_4, CO_2, NH_3, H_2S, O_2 \text{ and/or } N_2\}$

$d \in \{C, Norg, Nam, P, K, H2O \text{ and/or } Rest\}$

$e \in \{CH_4, NH_3 \text{ and/or } H_2S\}$

$h \in \{CH_4, CO_2, O_2, N_2\}, \{O_2, N_2\} \text{ or } \{CO_2, O_2, N_2\}$

Parameters

$A_{specific}$: specific clarifier area (m² / (ton·day))

$A(i)$: Antoine A coefficient for vapor pressure of component i

$B(i)$: Antoine B coefficient for vapor pressure of component i

$C(i)$: Antoine C coefficient for vapor pressure of component i

Cp_{sat} : specific heat capacity of flue gas.

d : work days per year

d_p : particle diameter (m)

k : kinetic constant (s-1)

IAP_{eq} : equilibrium ion activity product

h : work hours per day

HRT_{unit} : hydraulic retention time of *unit* (s)

$MW_{component}$: molecular weight of component (kg/kmol)

MeP_{ratio} : metal/phosphorus molar ratio in coagulation process

$Price_{component}$: price of the *component* (€/kg)

g : gravity acceleration (m2/s)

k := polytropic coefficient (1.4)

$\kappa_{agitator}$: agitators specific power consumed (HP / 1000 USgallon)

φ_j : precipitation agent *j* per total solids mass ratio

η_c : Compressor's efficiency (0.85)

η_s : Isentropic efficiency (0.9)

η_i^j : *i* component separation yield using in the process the element *j*

P_{atm} : atmospheric pressure (1 bar)

T_{atm} : atmospheric temperature (25 °C)

R : ideal gas constant (8.314 J/mol·K)

Cp_{H_2O} : specific heat capacity of water (4.18 kJ/kg·°C)

Variables

$a_{technology}$: parameter which takes the value 0 when $F_{design}^{technology}$ is 0 and 1 if $F_{design}^{technology}$ is not equal to 0

α_{mf} : parameter dependent of the phases number in the FBR

Ar_l : Arquimedes number for liquid

A_{unit} : area of *unit* (m^2)

$Benefits_{technology}$: benefits or losses obtained with *technology*

$C\text{-}N$: carbon to nitrogen molar ratio

C_{eq} : equilibrium concentration (kmol/m^3)

C_0 : initial concentration (kmol/m^3)

$Cost_{unit}$: cost of *unit*

$C^{unit}_{component}$: concentration of *component* in the *unit* inlet stream (kg$_{component}$/kg$_{total}$)

$ChemC_{technology}$: cost of chemicals for *technology*

D_{unit} : diameter of *unit*

e_{unit} : thickness of *unit*

$Ec_j(T)$: equilibrium constant of component *j* at temperature T.

$F^{unit}_{component}$:mass flow of ~~the~~ *component* in the *unit* inlet stream (kg/s)

F^{unit}_{max} : maximum mass inlet flow admitted by a single *unit* (kg/s)

F^{unit}_{design} : mass inlet flow used in the design of *unit* (kg/s)

$FC_{technology}$: fixed cost of *technology*

$F^{recovered}_{total}$: recovered matter total mass flow (kg/s)

$F_{(unit,unit1)}$: mass flow from stream from unit to unit1 (kg/s)

$fc_{(J,unit,unit1)}$: mass flow of component J from unit to unit1 (kg/s)

$H_{b,(unit,unit1)}$: enthalpy of the stream at the state b from the stream from unit to unit1 (kJ/kg).

$H_{steam\ (isoentropy)}$: enthalpy of the stream at the if the expansion is isentropic (kJ/kg).

l_{j-i} : molar fraction of component *j* in the liquid phase of equilibrium system *i*.

K_{index} : Potassium index of fertilizer.

L_{unit} : length of *unit*

N_{am} : nitrogen contained in ammonia.

N_{org} : nitrogen contained in organic matter.

n_{unit} : number of *units* used in the process

$n_{(unit,unit1)}$: total mol flow from stream from unit to unit1 (kmol/s).

N_{index} : nitrogen index of fertilizer.

$P_{in/compressor}$: inlet pressure to compressor (bar).

$P_{out/compressor}$: outlet pressure of compressor (bar).

$P_j{}^*(T)$: saturation pressure of pure component *j* at temperature T (bar).

P_v : vapor pressure (bar)

P_{index} : phosphorous index of fertilizer.

p_{turbi} : inlet pressure to body *i* in the turbine (bar)

P_{unit} : power of *unit*

$Q_{(unit)}$: heat exchanged in unit (kW).

$R_{C-N/k}$: carbon to nitrogen ratio in *k*.

$R_{C-N/\text{fertilizer}}$: carbon to nitrogen ratio in fertilizer.

$R_{V/F-i}$: rate of evaporation in equilibrium system *i*.

$Rest$: rest of the elements contained in the biomass.

$Re_{l\,mf}$: Reynolds number for liquid in minimum fluidization conditions

$s_{b(unit,unit1)}$: entropy the stream at the state b for the stream from unit to uni1 kJ/kg.K

$T_{turbimin}$: saturating temperature at exit of body *i* (ºC)

$T_{(unit,unit1)}$: temperature of the stream from unit to unit 1 (ºC)

$T_{bubble/i}$: bubble point temperature of equilibrium system *i* (ºC).

$T_{m/i}$: average temperature in equilibrium system *i* (ºC).

$T_{in/compressor}$: inlet temperature to compressor (ºC).

$T_{out/compressor}$: outlet temperature of compressor (ºC).

t : time (s)

u_t : terminal velocity (m/s)

u_0 : fluid velocity (m/s)

u_{mf} : minimum fluidization velocity (m/s)

v_{j-i} : ~~$v_{i,j}$=~~molar fraction of component *j* in the vapor phase of equilibrium system *i*.

$V_{biogas,\,k}$: ~~$V_{biogas,k}$= B~~biogas volume produced per unit of volatile solids (VS) ($m^3_{biogás}/kg_{VS/k}$) associated to *k*.

V_{unit} : volume of *unit*

W_{unit} : weight of *unit*

$w_{DM/k}$: dry mass fraction of *k* ($kg_{DM/k}$/kg).

$w'_{VS/k}$: dry mass fraction of volatile solids out of the dry mass of *k* ($kg_{VS/k}/kg_{DM/k}$).

$w'_{C/k}$: dry mass fraction of C in k ($kg_{C/k}/kg_{DM/k}$).

$w'_{Nam/k}$: dry mass fraction of Nam in k ($kg_{Nam/k}/kg_{DM/k}$).

$w'_{Norg/k}$: dry mass fraction of Norg in k ($kg_{Norg/k}/kg_{DM/k}$).

$w'_{P/k}$: dry mass fraction of P in k ($kg_{P/k}/kg_{DM/k}$).

$w'_{K/k}$: dry mass fraction of K in k ($kg_{K/k}/kg_{DM/k}$).

$w'_{Rest/k}$: dry mass fraction of the rest of the elements contained in k ($kg_{K/k}/kg_{MS/k}$).

$W_{(unit)}$: power produced or consumed in unit (kW).

$x_{a/biogas}$: mass fraction of component *a* in the biogas

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y^{j} : binary variable to evaluate the element *j*

y_{biogas} : specific saturated moisture of biogas

$Y_{a'/biogas-dry}$: molar fraction of component a in the dry biogas.

$\Delta H_{reaction|(Bioreactor)}$: Heat of the anaerobic digestion's reaction (kW).

$\Delta H_{comb}(k)$: heat of combustion of component *k* (kW).

$\Delta H_{comb}(e)$: heat of combustion of component *e* (kW).

$\Delta H_{comb}(Digestate-dry)$: heat of combustion of dry digestate (kW)

$\Delta H_{f}(h)_{T(unit,unit1)}$: heat of formation of component *h* at temperature $T_{(unit,unit1)}$ (kW)

Z : objective function

$\rho_{component}$: *component* density (kg/m^3)

$\mu_{component}$: viscosity of component (kg/(m·s))

~~In this work, we have optimized the production of biogas-based electrical power and fertilizers from manure. Detailed equation based models for the anaerobic digestion, the Brayton and regenerative Rankine cycles and different technologies for digestate treatment have been developed. To solve the model a two-step procedure is performed. First, the individual detailed models for each digestate treatment technology are used to formulate a MINLP model aiming at selecting the best configuration for that technology: the best precipitation agent, filter media, etc. In a second step the best configuration of each technology has been implemented in the entire superstructure. Due to the fact that only one digestate processing technology is allowed and the highly non linear nature of the model, surrogate models for the cost of each alternatives with a smooth approximations have been developed. For the optimal selection a detailed economic evaluation is performed.~~

~~The results show that FBR technologies are preferred to recovery nutrients. Furthermore, in some cases this process can produce electricity at a competitive i.e. Poultry and sheep manure. The investment cost is highly dependent on the water and organic content of the manure type, from 70M€ to 135M€ when a large energy production is possible and large gas and steam turbines are to be installed. However, for these cases of high investment cost, the production cost of power is the most competitive resulting in values from 4000€/MW for the largest investment but also largest power production to 25000€/kW produced in case of cattle or pig manure.~~

~~**Nomenclature**~~

~~Parameters~~

~~η_i^j : *i* component separation yield using in the process the element *j*~~

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$Price_{component}$: price of the *component*

$MW_{component}$: molecular weight of *component*

MeP_{ratio} : metal/phosphorus molar ratio in coagulation process

$\kappa_{agitator}$: agitators specific power consumed (HP / 1000 USgallon)

$A_{specific}$: specific clarifier area (m^2 / (ton·day))

h : work hours per day

d : work days per year

φ_j : precipitation agent / per total solids mass ratio

HRT_{unit} : hydraulic retention time of *unit* (s)

k : kinetic constant (s-1)

IAP_{eq} : equilibrium ion activity product

d_p : particle diameter

g : gravity acceleration

Variables

$F^{unit}_{component}$:mass flow of the *component* in the *unit* inlet stream (kg/s)

η^j_i : *i* component separation yield using in the process the element *j*

y^j : binary variable to evaluate the element *j*

$C^{unit}_{component}$: concentration of *component* in the *unit* inlet stream (kg$_{component}$/kg$_{total}$)

$\rho_{component}$: *component* density (kg/m^3)

n_{unit} : number of *units* used in the process

F^{unit}_{max} : maximum mass inlet flow admitted by a single *unit* (kg/s)

F^{unit}_{design} : mass inlet flow used in the design of *unit* (kg/s)

$FC_{technology}$: fixed cost of *technology*

$ChemC_{technology}$: cost of chemicals for *technology*

$Price_{component}$: price of the *component*

$MW_{component}$: molecular weight of *component*

MeP_{ratio} : metal/phosphorus molar ratio in coagulation process

V_{unit} : volume of *unit*

D_{unit} : diameter of *unit*

L_{unit} : lenghtlength of *unit*

e_{unit} : thickness of *unit*

W_{unit} : weight of *unit*

$Cost_{unit}$: cost of *unit*

P_{unit} : power of *unit*

$\kappa_{agitator}$: agitators specific power consumed (HP / 1000 USgallon)

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~~A_{unit} : area of *unit* (m^2)~~

~~$A_{specific}$: specific clarifier area (m^2 / (ton·day))~~

~~h : work hours per day~~

~~d : work days per year~~

~~φ_j : precipitation agent *j* per total solids mass ratio~~

~~HRT_{unit} : hydraulic retention time of *unit* (s)~~

~~t : time (s)~~

~~k : kinetic constant (s-1)~~

~~IAP_{eq} : equilibrium ion activity product~~

~~C_{eq} : equilibrium concentration (kmol/m^3)~~

~~C_0 : initial concentration (kmol/m^3)~~

~~u_{mf} : minimum fluidization velocity (m/s)~~

~~$\mathrm{Re}_{l\ mf}$: Reynolds number for liquid in minimum fluidization conditions~~

~~$\mu_{component}$: viscosity of component (kg/(m·s))~~

~~d_p : particle diameter~~

~~Ar_l : Arquimedes number for liquid~~

~~u_t : terminal velocity (m/s)~~

~~g : gravity acceleration~~

~~u_0 : fluid velocity (m/s)~~

~~$a_{technology}$: parameter which takes the value 0 when $F_{design}^{technology}$ is 0 and 1 if $F_{design}^{technology}$ is not equal to 0~~

~~α_{mf} : parameter dependantdependent of the phases number in the FBR~~

~~$F_{total}^{recovered}$: recovered matter total mass flow (kg/s)~~

~~Z : objective function~~

~~$\mathrm{Benefits}_{technology}$: benefits or losses obtained with *technology*~~

~~W_{unit} : Power produced or consumed in *unit* (kW)~~

Acknowledgments

We acknowledge funding from the National Science Foundation (under grant CBET-1604374) and MINECO (under grant DPI2015-67341-C2-1-R) and EM also acknowledges an undergraduate research grant.

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~~Appendix~~

~~Temperature in ºC~~

~~Compressed liquid~~

~~H(kJ/kg)=4.2921*(TEMPERATURE) + 4.1269~~

~~S (kJ/(kg K))=1.1902*10~~$^{-5}$~~*(TEMPERATURE)~~^{3} ~~– 3.7465*10~~$^{-3}$~~*(TEMPERATURE)~~^{2} ~~+ 4.5352*(TEMPERATURE) + 0.64547~~

~~Saturated liquid~~

~~H(kJ/kg)=3.6082*10~~$^{-12}$~~*(TEMPERATURE)~~^{6} ~~– 3.4120*10~~$^{-9}$~~*(TEMPERATURE)~~^{5} ~~+ 1.2303*10~~$^{-6}$~~*(TEMPERATURE)~~^{4} ~~– 2.0306*10~~$^{-4}$~~*(TEMPERATURE)~~^{3} ~~+~~

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~~1.5552*10^{-2}*(TEMPERATURE)^{2} + 3.7216*(TEMPERATURE) + 3.0035~~
~~S (kJ/(kg K))= 1.0372*10^{-12}*(TEMPERATURE)^{5} – 8.6494*10^{-10}*(TEMPERATURE)^{4} +~~
~~2.8965*10^{-7}*(TEMPERATURE)^{3} – 5.6730*10^{-5}*(TEMPERATURE)^{2} +~~
~~1.6802*10^{-2}*(TEMPERATURE) - 2.1997*10^{-2}~~

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~~Saturated Vapor~~
~~H (kJ/kg)=-6.5690*10^{-12}*(TEMPERATURE)^{6} +~~
~~6.3049*10**(-9)*(TEMPERATURE)^{5} – 2.3080*10^{-6}*(TEMPERATURE)^{4} +~~
~~3.8339*10**(-4)*(TEMPERATURE)^{3} – 3.0632*10^{-2}*(TEMPERATURE)^{2} +~~
~~2.7553*(TEMPERATURE) + 2.4957*10^{3}~~
~~S (kJ/(kg K))=~~
~~-2.0373*10^{-12}*(TEMPERATURE)^{5} + 1.8589*10^{-9}*(TEMPERATURE)^{4} –~~
~~7.1901*10^{-7}*(TEMPERATURE)^{3} + 1.6112*10^{-4}*(TEMPERATURE)^{2} –~~
~~2.8904*10^{-2}*(TEMPERATURE) + 9.1915~~

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~~Superheated steam (Up to 10 bar)~~
~~H (kJ/kg)= (-0.0000063293*(Pressure(bar)) +~~
~~0.00033179)*(TEMPERATURE)^{2}+(0.0124*(Pressure(bar)) + 1.8039)*(TEMPERATURE)+(-~~
~~6.0707*(Pressure(bar))+2504.6)~~
~~S (kJ/(kg K))=~~
~~0.000000000942*(TEMPERATURE)^{3}-~~
~~0.00000309*(TEMPERATURE)^{2}+0.00524*(TEMPERATURE)+(6.8171*((Pressure(bar)))$^{(-0.069455)}$)~~

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~~Superheated steam (10 bar -150bar)~~
~~H (kJ/kg)= (-0.00000000000011619*(Pressure(bar))^{2} – 0.0000000000087596*(Pressure(bar)) -~~
~~0.00000000022611)*(TEMPERATURE) ^{4}+~~
~~(0.0000000004298*(Pressure(bar))^{2} + 0.00000003276*(Pressure(bar)) +~~
~~0.0000007313)*(TEMPERATURE)^{3}+~~
~~(-0.0000005801*(Pressure(bar))^{2} - 0.000046*(Pressure(bar)) - 0.0005009)*(TEMPERATURE)^{2}+~~
~~(0.0003383*(Pressure(bar))^{2} + 0.02947*(Pressure(bar)) + 2.195)*(TEMPERATURE)+~~
~~(-0.072042*(Pressure(bar))^{2} – 7.7877*(Pressure(bar)) + 2440.8)~~
~~S (kJ/(kg K))= (0.000000000015719*(Pressure(bar)) + 0.00000000074013)*(TEMPERATURE)^{3}+~~
~~(-0.00000000010074*(Pressure(bar))^{2} - 0.000000030171*(Pressure(bar)) -~~
~~0.0000028872)*(TEMPERATURE)^{2}+~~
~~(0.000000094914*(Pressure(bar))^{2} + 0.000029097*(Pressure(bar)) + 0.0050938)*(TEMPERATURE)+~~
~~(0.000041223*(Pressure(bar))^{2} – 0.028841*(Pressure(bar)) + 5.9537)~~

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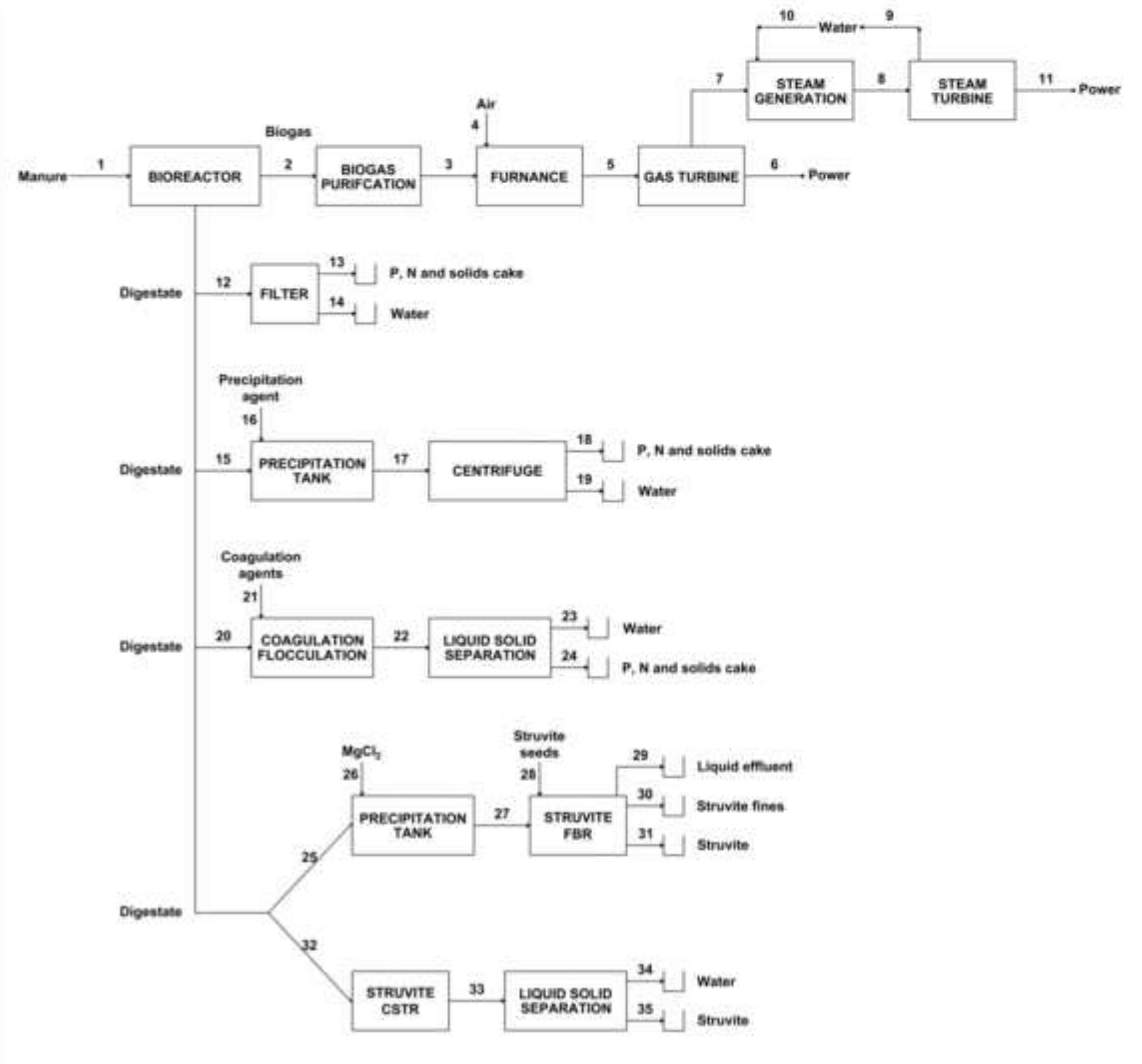
Figure 1



Manure
1
BIOREACTOR
Biogas
2
BIOGAS PURIFCATION
3
Air
4
FURNANCE
5
GAS TURBINE
6
Power
7
STEAM GENERATION
8
STEAM TURBINE
11
Power
10
Water
9
Digestate
12
FILTER
13
P, N and solids cake
14
Water
Precipitation agent
16
Digestate
15
PRECIPITATION TANK
17
CENTRIFUGE
18
P, N and solids cake
19
Water
Coagulation agents
21
Digestate
20
COAGULATION FLOCCULATION
22
LIQUID SOLID SEPARATION
23
Water
24
P, N and solids cake
$MgCl_2$
26
Struvite seeds
28
PRECIPITATION TANK
27
STRUVITE FBR
29
Liquid effluent
30
Struvite fines
31
Struvite
25
Digestate
32
STRUVITE CSTR
33
LIQUID SOLID SEPARATION
34
Water
35
Struvite

Figure 2

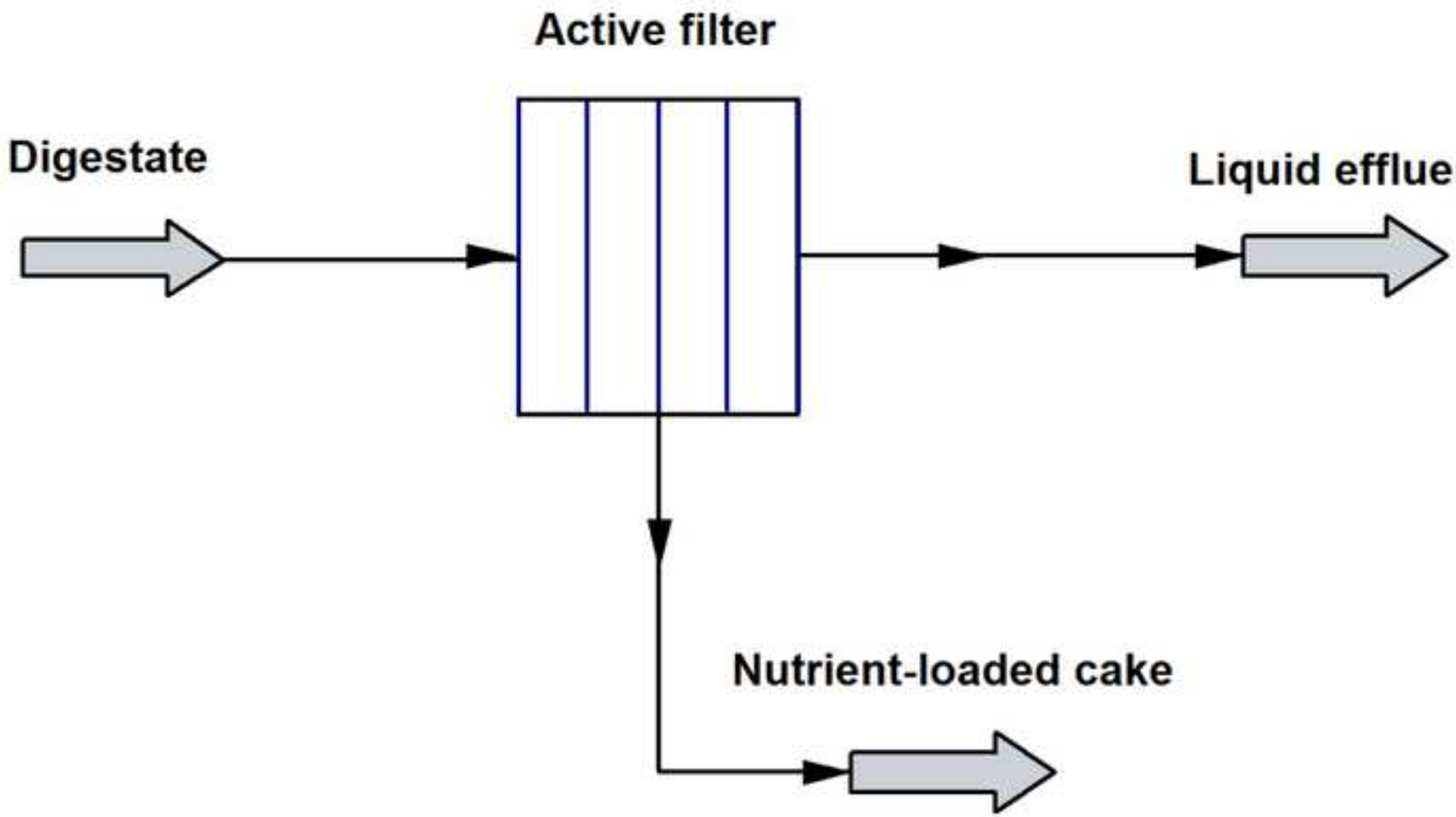
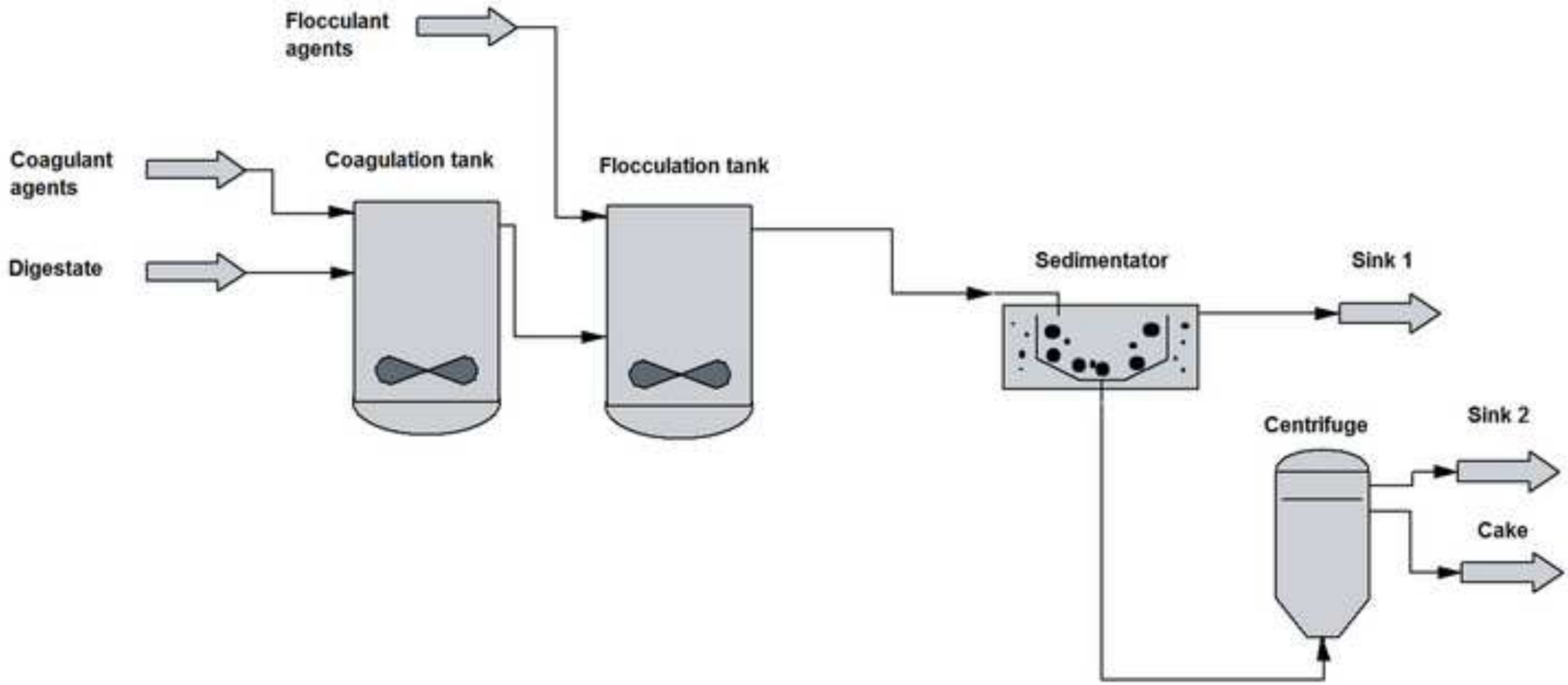

Active filter
Digestate
Liquid effluent
Nutrient-loaded cake

Figure 3



Flocculant agents
Coagulant agents
Coagulation tank
Flocculation tank
Digestate
Sedimentator
Sink 1
Centrifuge
Sink 2
Cake

Figure 4

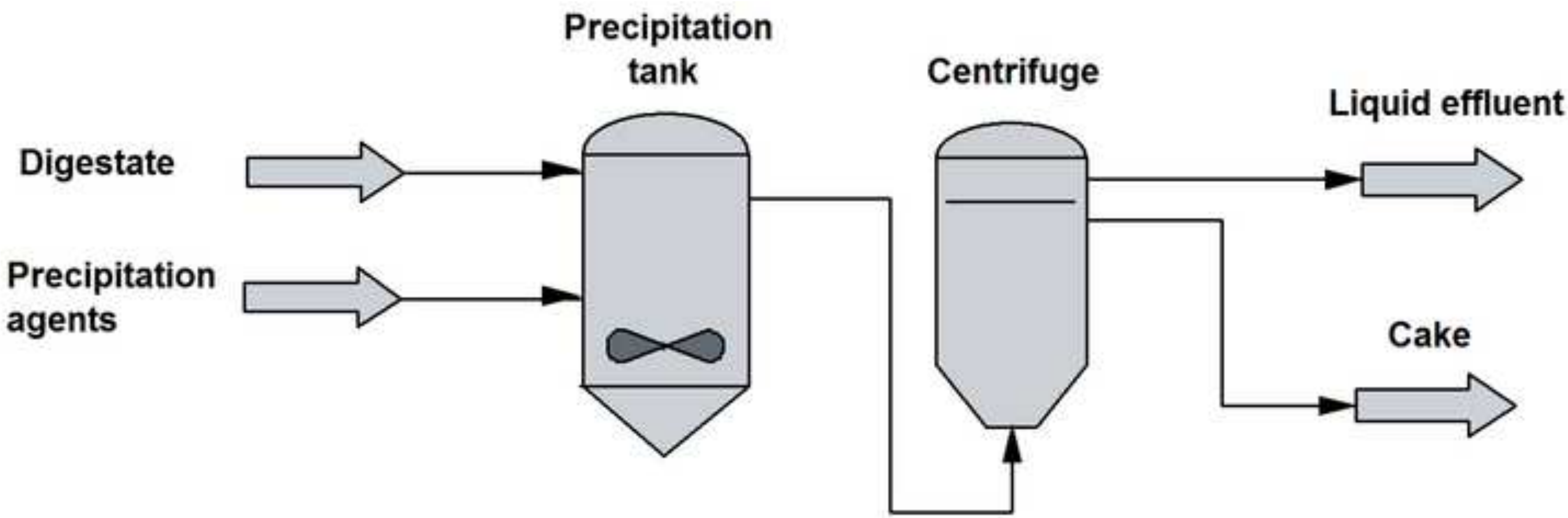
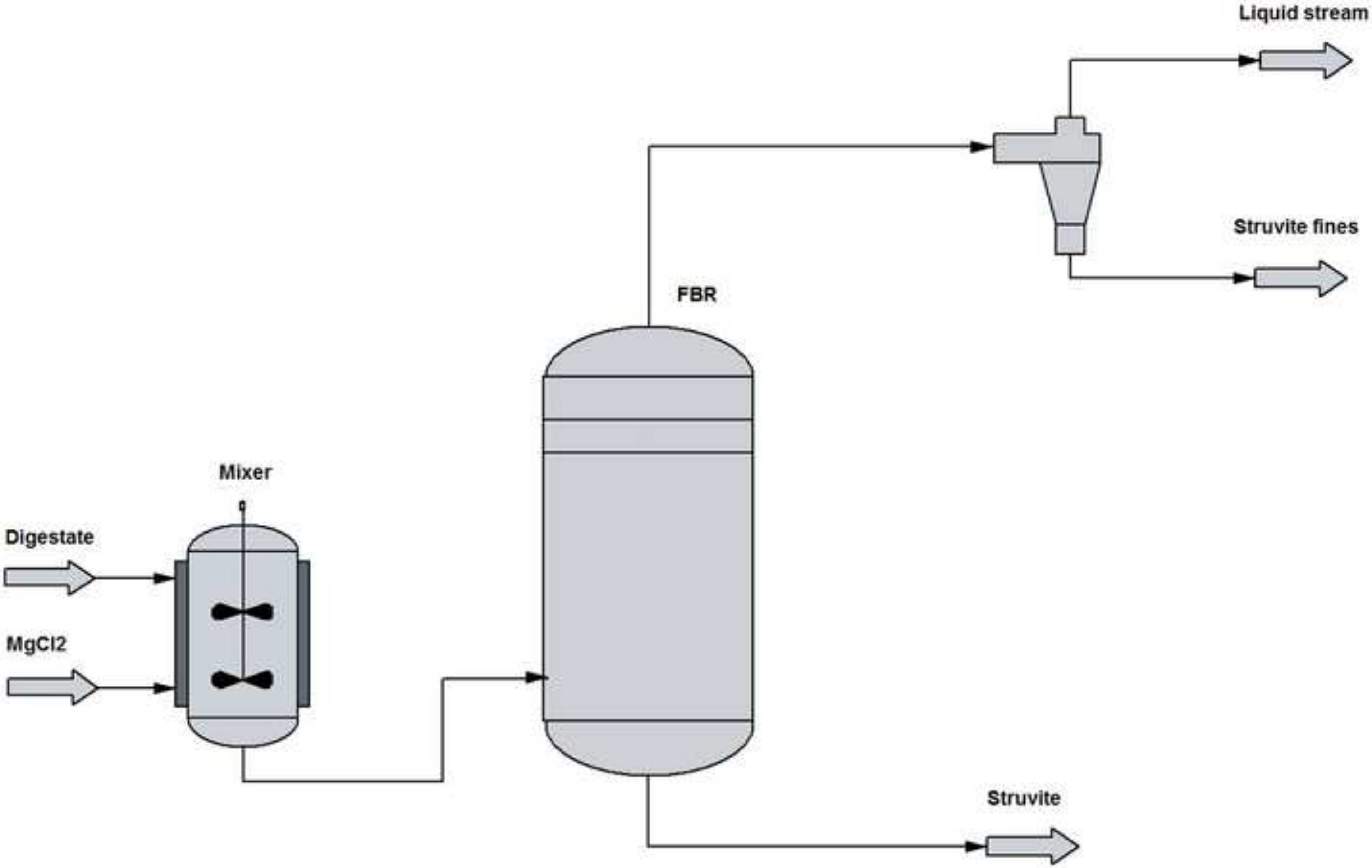
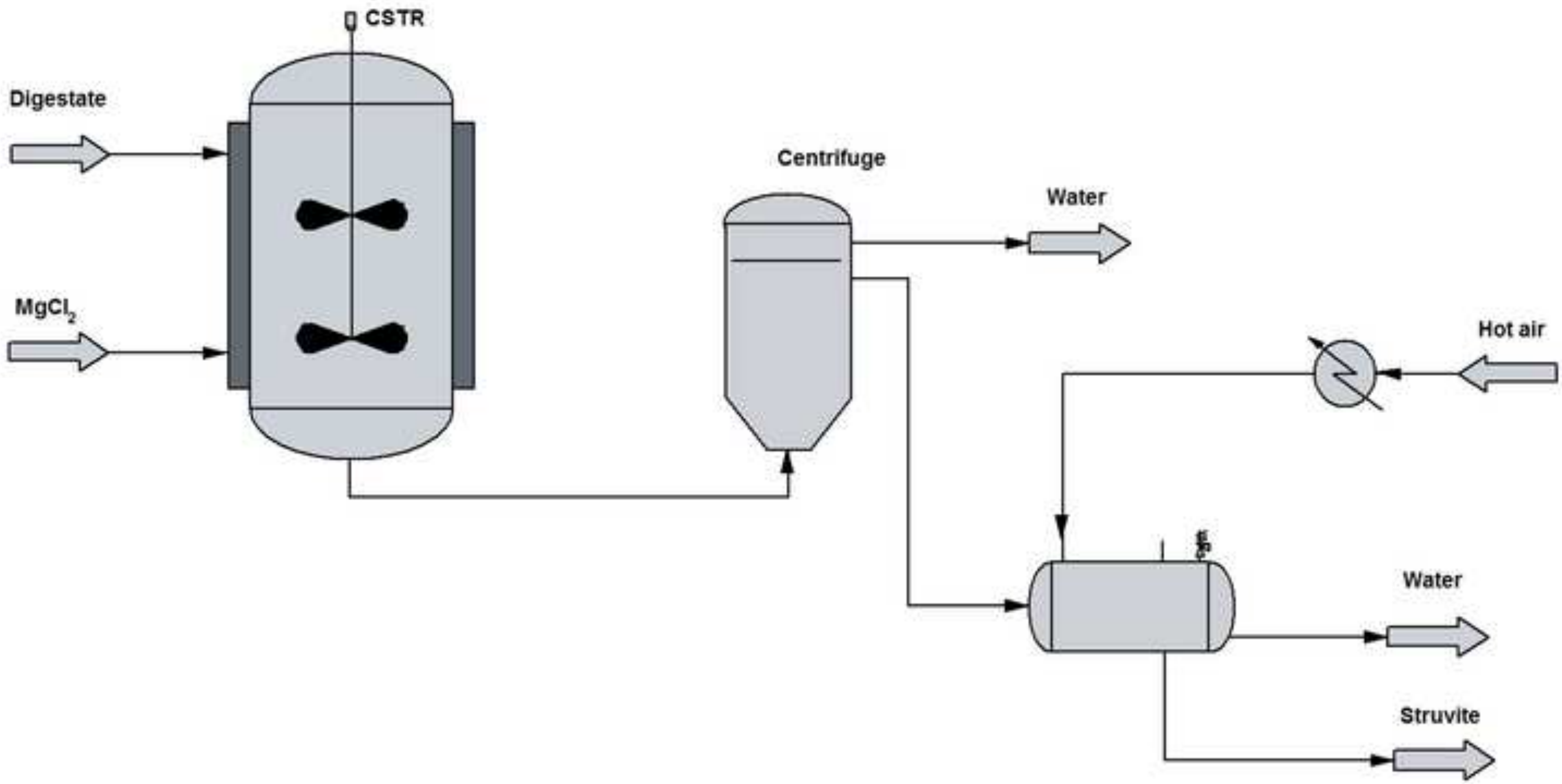

Precipitation
tank
Centrifuge
Liquid effluent
Digestate
Precipitation
agents
Cake

Figure 5



Liquid stream
Struvite fines
FBR
Mixer
Digestate
MgCl2
Struvite

Figure 6



CSTR
Digestate
$MgCl_2$
Centrifuge
Water
Hot air
Water
Struvite

Figure 7

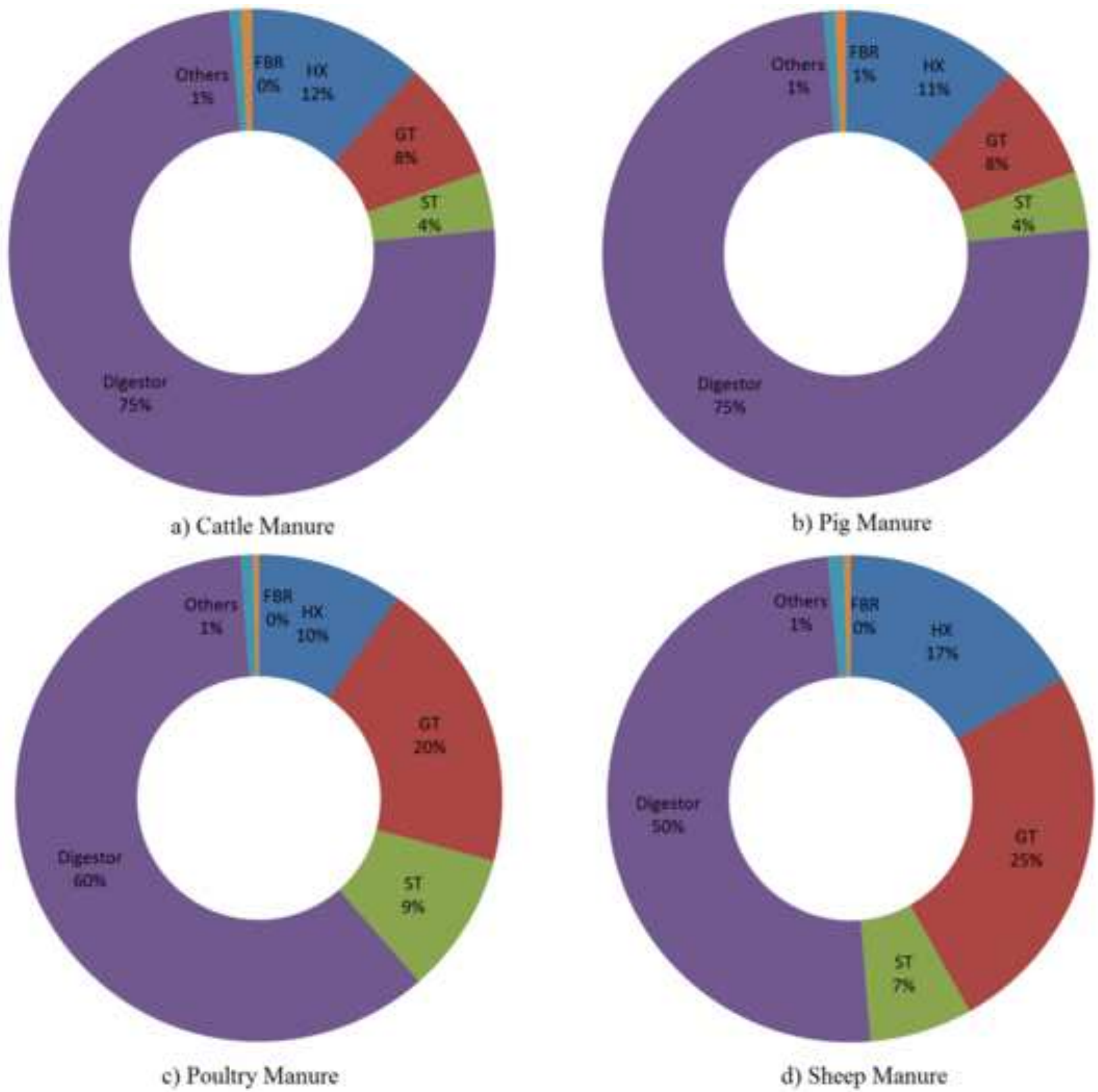
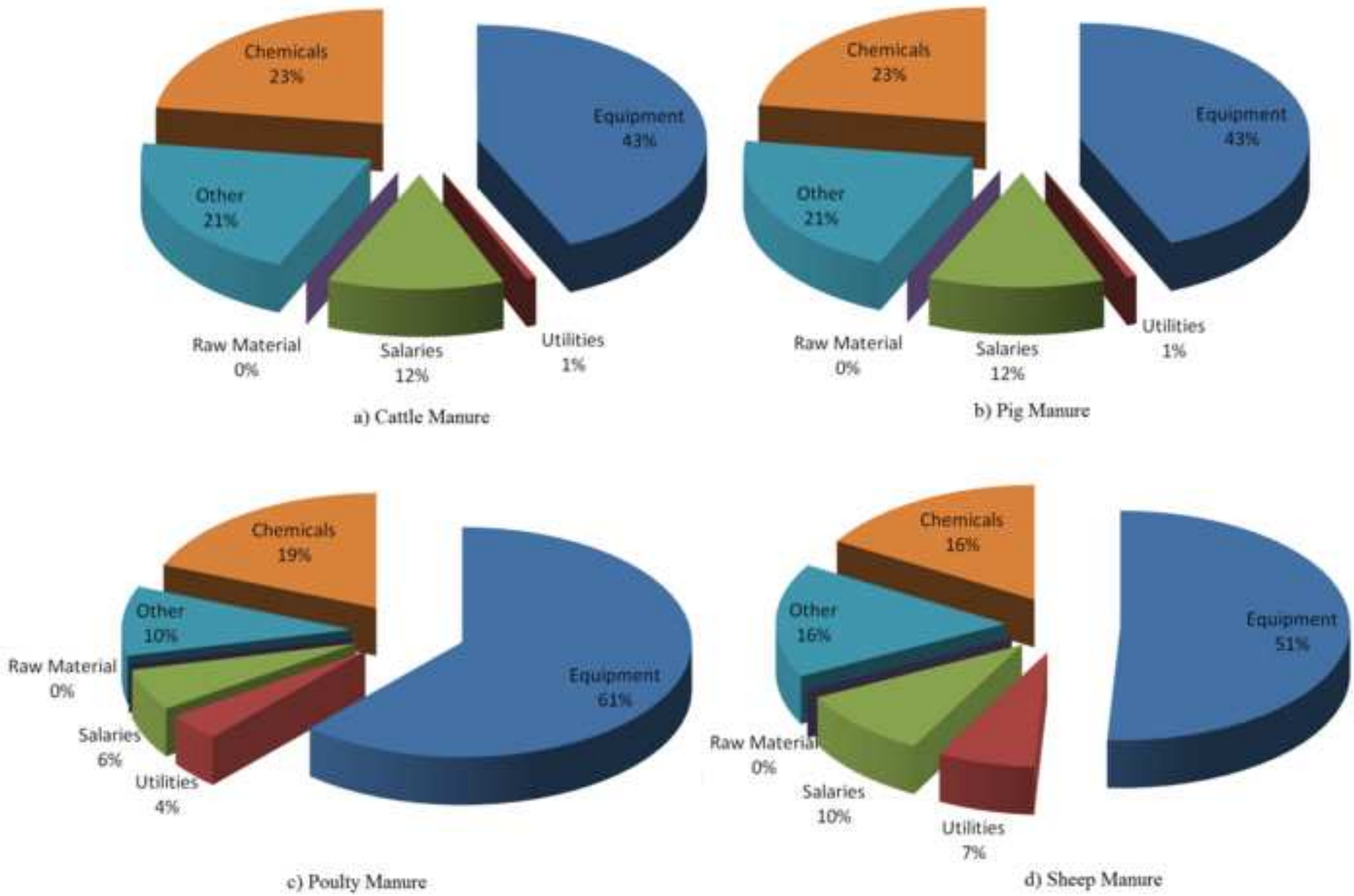

Others
1%
FBR
0%
HX
12%
GT
8%
ST
4%
Digestor
75%
a) Cattle Manure
Others
1%
FBR
1%
HX
11%
GT
8%
ST
4%
Digestor
75%
b) Pig Manure
Others
1%
FBR
0%
HX
10%
GT
20%
ST
9%
Digestor
60%
c) Poultry Manure
Others
1%
FBR
0%
HX
17%
GT
25%
ST
7%
Digestor
50%
d) Sheep Manure

Figure 8


Chemicals
23%
Equipment
43%
Other
21%
Raw Material
0%
Salaries
12%
Utilities
1%
a) Cattle Manure
Chemicals
23%
Equipment
43%
Other
21%
Raw Material
0%
Salaries
12%
Utilities
1%
b) Pig Manure
Chemicals
19%
Other
10%
Raw Material
0%
Equipment
61%
Salaries
6%
Utilities
4%
c) Poulty Manure
Chemicals
16%
Other
16%
Equipment
51%
Raw Material
0%
Salaries
10%
Utilities
7%
d) Sheep Manure

Tables

Table 1: Manure composition and properties

Manure / Element	Dry matter (% wt)		N (% dry mass)		P (% dry mass)		K (% dry mass)		VS (% dry mass)		V_{biogas} (m^3_{gas}/kg_{VS})		Density (kg/m^3)
	max	min	max	min	max	min	max	min	max	min	max	min	
Cattle	10	2	8	4,7	1.3	0.8	10	3.3	0.8	0.8	0,3	0,2	1041.2
Pig	6	2	15	13	2.2	1.9	8.3	3.9	0.8	0.7	0,5	0,25	1000.0
Poultry	60	30	5.4	5.4	1.7	1.7	1,2	2,3	0.8	0.8	0,6	0,35	1009.2
Sheep	28	28	2,9	2,9	0.78	0.78	2.9	2.9	0.8	0.8	0,61	0,37	1009.2

Table 2: Set of components

Components set							
Number of component	**Component**	**Number of component**	**Component**	**Number of component**	**Component**	**Number of component**	**Component**
1	Wa	**12**	O	**23**	K2O	**34**	Cl
2	CO2	**13**	N	**24**	CaCO3	**35**	Struvite
3	CO	**14**	Norg	**25**	FeCl3	**36**	KStruvite
4	O2	**15**	P	**26**	Antifoam	**37**	MgCl2_CSTR
5	N2	**16**	K	**27**	Fe2SO4_3	**38**	NaOH_CSTR
6	H2S	**17**	S	**28**	Al2SO4_3	**39**	Mg_CSTR
7	NH3	**18**	Rest	**29**	AlCl3	**40**	Cl_CSTR
8	CH4	**19**	Cattle_slurry	**30**	MgCl2	**41**	Struvite_CSTR
9	SO2	**20**	Pig_slurry	**31**	NaOH	**42**	KStruvite_CSTR
10	C	**21**	Poultry_slurry	**32**	Struvite_seeds	**43**	FeCl3_Coag
11	H	**22**	P2O5	**33**	Mg		

Table 3. Recovered P and N yield for different filter media.

Media/Nutrient	P (% recovered)	N (% recovered)
Polonite	96.7[a]	18.0[c]
Filtra_P	98.2[a]	50.0[e]
Wollastonite	51.1[a]	70.0[d]
Dolomite	44.0[b]	50.0[e]
Metal_slag	85.6[a]	67.0[f]

a: Gustafsson et al., 2008.
b Pant et al., 2001.
c: Kietlinska and Renman, 2005.
d: Lind et al., 2000.
e: Aziz et al., 2004.
f: Yang et al., 2009

Table 4.- Operating data of the optimal configuration for each raw material.

		T (ºC)	P (bar)	Extractions
Cattle	**Bioreactor**	55	1	--
	Gas Turbine	2430 (In) 1205 (Out)	8.2 (In) 1 (Out)	--
	Steam Turbine	1000 (T_1) 568 (T_2) 442 (T_3) 41.8 (T_4)	125 (P_1) 11 (P_2) 5 (P_3) 0.08 (P_4)	6.7% to HX7
	FBR	25	1	--

		T (ºC)	P (bar)	Extractions
Pig	**Bioreactor**	55	1	--
	Gas Turbine	2430 (In) 1205 (Out)	8.2 (In) 1 (Out)	--
	Steam Turbine	1000 (T_1) 568 (T_2) 442 (T_3) 41.8 (T_4)	125 (P_1) 11 (P_2) 5 (P_3) 0.08 (P_4)	6.7% to HX7
	FBR	25	1	--

		T (ºC)	P (bar)	Extractions
Poultry	**Bioreactor**	55	1	--
	Gas Turbine	2430 (In) 1205 (Out)	8.2 (In) 1 (Out)	--
	Steam Turbine	1000 (T_1) 568 (T_2) 442 (T_3) 41.8 (T_4)	125 (P_1) 11 (P_2) 5 (P_3) 0.08 (P_4)	6.7% to HX7
	FBR	25	1	--

		T (ºC)	P (bar)	Extractions
Sheep	**Bioreactor**	55	1	--
	Gas Turbine	2337 (In) 896 (Out)	15.6 (In) 1 (Out)	--
	Steam Turbine	769.6 (T_1) 439.1 (T_2) 329.6 (T_3) 73.0 (T_4)	95 (P_1) 11 (P_2) 5 (P_3) 0.35 (P_4)	2.9% to HX7
	FBR	25	1	--

Table 5.- Process optimization results for considered manures

Manure	Power (kW)	Comp Biogas (CH4/CO2 ratio)	Digestate treatment technology	Product recovered	Biogas/Manure ratio	Digestate/Manure ratio
Cattle	2,612	0.816	FBR struvite	Struvite	0.0208	0.9794
Pig	2,612	0.816	FBR struvite	Struvite	0.0208	0.9794
Poultry	31,349	0.818	FBR struvite	Struvite	0.2499	0.7526
Sheep	14,106	0.818	FBR struvite	Struvite	0.1217	0.8795

Table 6.- Biogas composition for considered manures

Manure	CH_4 (%wt)	CO_2 (%wt)	Water (%wt)	O_2 (%wt)	N_2 (%wt)
Cattle	0.385	0.470	0.120	0.006	0.020
Pig	0.385	0.470	0.120	0.006	0.020
Poultry	0.385	0.470	0.120	0.006	0.020
Sheep	0.385	0.470	0.120	0.006	0.020

Table 7: Electricity production cost and NPV for the facility considering different raw materials

Raw material	**Annual Production costs (M€/y)**	**Electricity production cost (€/kWh)**	**NPV**
Cattle manure	12.04	0.45	-1.93E+07
Pig manure	12.07	0.45	-1.96E+07
Poultry manure	25.51	0.03	2.85E+08
Sheep manure	15.53	0.10	5.46E+07