

Supplementary Information

Multi-scale techno-economic assessment of nitrogen recovery systems for livestock operations

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S1 Modeling details of nutrient recovery processes

We note that the operating costs reported include the expenditures of running each process, but do not include the amortization of the capital expenses.

S1.1 Anaerobic digestion

The energy requirements of anaerobic digestion (AD) unit, described in Eq 1S, comprise the energy required for substrate warming from ambient temperature, assuming an average annual temperature of 12 °C, which is the average annual temperature in central Spain, to the digestion temperature (40 °C), Eq. 2S, and the energy supplied to offset the digester heat losses, Eq. 3S. The global heat transfer coefficients (U) for the different digester surfaces are collected in Table 1S. The area of different surfaces is estimated through preliminar equipment design, Eqs. 4S to 8S, assuming a maximum digester of 6000 m³ (Fachagentur Nachwachsende Rohstoffe, 2010). The installation of multiple digestion units is considered if the flow of manure to be treated exceeds this capacity.

$$Q_{\text{digester}} = Q_{\text{feedstock}} + Q_{\text{losses}} \quad (1S)$$

$$Q_{\text{feedstock}} = \dot{m}_{\text{manure}} \cdot c_p \cdot (T_{\text{digestion}} - T_{\text{ambient}}) \quad (2S)$$

$$Q_{\text{losses}} = \sum_i U_i A_i (T_{\text{digestion}} - T_{\text{ambient}}), \quad i \in \{\text{roof, walls, floor}\} \quad (3S)$$

Table 1S: Anaerobic digester design parameters. ([Penn State Extension, 2016](#)).

Parameter	Unit	Value
U_{floor}	W/m ² K	2.85
U_{wall}	W/m ² K	0.39
U_{roof}	W/m ² K	0.3
V_{max}	m ³	6000
$V_{\text{freeboard}}$	%	30
HRT	days	21
$D_{\text{digester}}:H_{\text{digester}}$	ratio	1.1
$A_{\text{roof}}:A_{\text{floor}}$	ratio	1.62

$$V_{\text{digester}} = \frac{\dot{m}_{\text{manure}}}{\rho_{\text{manure}}} \cdot HRT \cdot \left(1 + \frac{V_{\text{freeboard}}}{100}\right) \quad (4S)$$

$$V_{\text{digester}} = A_{\text{floor}} \cdot \frac{D_{\text{digester}}}{D_{\text{digester}} : H_{\text{digester}}} \quad (5S)$$

$$A_{\text{floor}} = \pi \cdot \frac{D_{\text{digester}}^2}{4} \quad (6S)$$

$$A_{\text{wall}} = 2\pi \cdot \frac{D_{\text{digester}}}{2} \cdot \frac{D_{\text{digester}}}{D_{\text{digester}} : H_{\text{digester}}} \quad (7S)$$

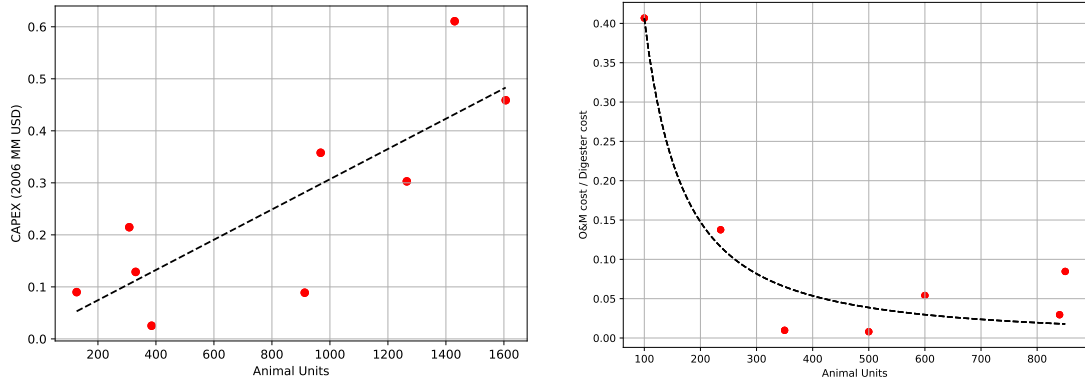
$$A_{\text{roof}} = A_{\text{floor}} \cdot (A_{\text{roof}} : A_{\text{floor}}) \quad (8S)$$

CAPEX and OPEX estimations as a function of animal units based on data reported by the USDA ([Beddoes et al., 2007](#)) (shown in Figure 1S) are shown in Eqs. 9S-11S, where $\dot{m}_{\text{digestate}}$ denotes the digestate flow and AU the number of animal units. It should be noted that the OPEX estimated in Eq. 11S only considers the operating and management (O&M) cost and it does not include the CAPEX amortization.

$$n_{AD} = \left\lceil \frac{\dot{m}_{digestate} \cdot HRT_{AD}}{n_{AD, max}} \right\rceil \quad (9S)$$

$$CAPEX \text{ (MM USD (2019))} = \left(\left(2.9069 \cdot 10^{-4} \cdot \frac{AU}{n_{AD}} + 0.01625 \right) \cdot 1.216 \right) n_{AD} \quad (10S)$$

$$OPEX \left(\frac{\text{MM USD (2019)}}{\text{year}} \right) = \left(\frac{15.858 \cdot 10^3}{1 + \left(\frac{AU}{n_{AD}} \cdot 13.917 \right)^{1.461}} \right) \cdot CAPEX \quad (11S)$$



(a) Cost of AD units as a function of animal units.

(b) O&M costs as a function of animal units.

Figure 1S: Correlations between AD capital and O&M costs, and the number of cattle in the livestock facility. Data from [Beddoes et al. \(2007\)](#).

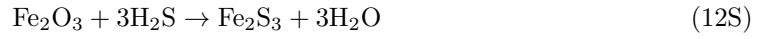
S1.2 Biogas conditioning

S1.2.1 Moisture removal

The biogas produced contains saturated water at operating temperature. The first stage for water removal is the moisture condensation, which can remove residual dust and oil as well. However, the biogas moisture can remain saturated at the cooled down temperature, and additional drying stage can be needed, using adsorbent agents such as silica gel or zeolites in packed bed columns operating at pressures between 6 and 10 bar ([Ryckeboosch et al., 2011](#)).

S1.2.2 H₂S removal

Hydrogen sulphide can be removed during the digestion step, or treating the biogas stream leaving the digester. A review of processes for H₂S removal can be found in [Ryckebosch et al. \(2011\)](#). In this work, in-biogas hydrogen sulphide removal is considered by using a fixed absorbent bed of Fe₂O₃, so that hydrogen sulphide is removed in the form of Fe₂S₃, as shown in Eq 12S. This unit operates at 25-50 °C. Although is sensitive to the presence of water, it is a high efficient process. Therefore, the moisture removal must be previously performed.



The bed can be regenerated using oxygen (air), leading to the formation of elementary sulfur, Eq. 13S.



S1.2.3 Ammonia removal

Ammonia contained in biogas can be removed by using adsorption methods such as fixed beds of zeolites or activated carbon, as well as through water scrubbing processes.

S1.3 Digestate solid-liquid separation

Nutrients contained in the organic materials (either manure or digestate, depending on whether AD is performed or not) are present in both organic and inorganic forms. Organic nutrients are chemically bound to carbon, and they have to be converted into their inorganic forms through a mineralization process to be available for the vegetation to grow. Organic nutrients are mainly contained in the solid phase of the organic materials. Inorganic nutrients are water soluble, and they are mostly present in the liquid phase, or bounded to soluble minerals. They are immediately available to plants, including algae involved in the occurrence of algal blooms. To

recover the inorganic fraction of nutrients, a solid-liquid separation stage is implemented, keeping the inorganic nutrients in the liquid stage, which will be further processed, and the organic nutrients in the solid phase, which can be composted to mineralize nitrogen and phosphorus and be further used as fertilizers.

Based on the evaluation reported by [Møller et al. \(2000\)](#), a screw press is the technology selected to carry out the solid-liquid separation stage since it is the most cost efficient solid-liquid separation equipment. The experimental results reported in this study are used to determine the partition coefficients for the different elements, as shown in Table 2S.

Table 2S: Partition coefficients for solid-liquid manure separation using a screw press unit ([Møller et al., 2000](#))

Element	Solid fraction	Liquid fraction
Total mass	0.08	0.92
Dry matter	0.31	0.69
Org. N	0.09	0.91
Org. P	0.22	0.78

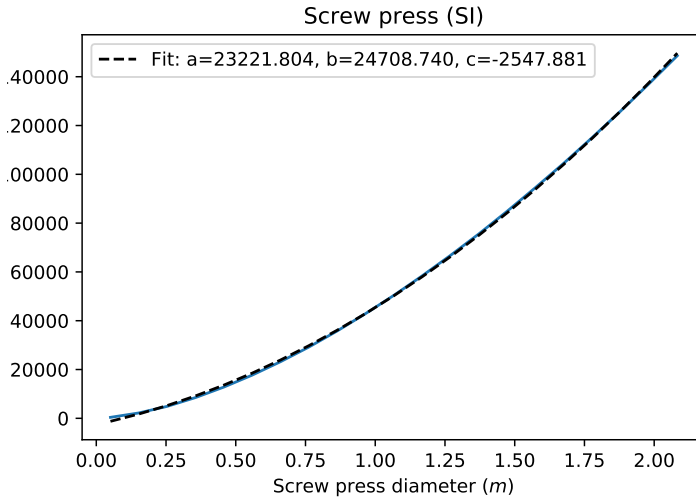


Figure 2S: Estimated screw press investment costs (USD) as a function of the size.

To determine the commercial sizes and number of units necessary as a function of the flow to be treated, data from commercial manufacturers are considered ([PWTech, 2018](#)). The feasible configurations in terms of screw press diameter and number of units as a function of the flow

or organic materials treated are shown in Table 3S. Data reported by [Matches \(2014\)](#) for this type of equipment are used to relate the unit diameter and cost, while the operating costs are calculated assuming power consumption reported by the manufacturer for each model, as shown in Fig. 2S and Tables 3S and 4S.

Table 3S: Sizing estimated for screw press units based on commercial data ([PWTech, 2018](#))

Load capacity $\left(\frac{m^3}{day}\right)$	Number of units			
	$\varnothing(m)$ 0.23	$\varnothing(m)$ 0.35	$\varnothing(m)$ 0.42	$\varnothing(m)$ 0.56
< 43	1	-	-	-
43 - 81	-	1	-	-
81 - 190	-	-	1	-
190 - 381	-	-	2	-
381 - 572	-	-	3	-
572 - 708	-	-	-	2
708 - 1090	-	-	-	3
1090 - 1444	-	-	-	4
> 1444	-	-	-	$\left\lceil \frac{\text{Flow } (m^3/day)}{\text{Load Capacity}_{\varnothing 0.56m \text{ unit}}} \right\rceil$

Table 4S: Electrical power of screw press units ([PWTech, 2018](#))

Number of units	Electrical power (kW)			
	$\varnothing(m)$ 0.23	$\varnothing(m)$ 0.35	$\varnothing(m)$ 0.42	$\varnothing(m)$ 0.56
1	0.3	0.45	0.9	-
2	-	-	1.27	3.88
3	-	-	2.01	6.34
4	-	-	-	7.83

S1.4 Struvite production: Multiform system

Struvite production from livestock organic materials has been estimated based on a detailed thermodynamic model for precipitates (struvite and calcium compounds) formation considering the main chemical systems involved in the formation of precipitates, shown in Tables 5S and 6S. pK denotes the thermodynamic equilibrium constant, and pK_{sp} refers to the solubility product. The variability in the organic waste composition is also considered, obtaining surrogate models to predict the formation of struvite based on different waste parameters, including the concentration of magnesium, the concentration of calcium, and the alkalinity level ([Martín-Hernández et al.,](#)

2020). Based on previous studies, the supply of magnesium has been fixed at a $\text{Mg}^{2+}/\text{PO}_4^{3-}$ molar ratio of 2, estimating the formation of struvite as a function of the concentration of Ca^{2+} ions, which compete for phosphate ions hindering the formation of struvite, shown in Eq. 14S. A comprehensive description of the thermodynamic model used for the development of the correlations to estimate the formation of struvite can be found in [Martín-Hernández et al. \(2020\)](#).

Table 5S: Aqueous phase chemical systems considered in the thermodynamic model for estimating the formation of struvite.

Name	Chemical system	pK	Source
Ammonia	$\text{NH}_4^+ \leftrightarrow \text{NH}_3 + \text{H}^+$	9.2	(Bates & Pinching, 1949)
Water	$\text{H}_2\text{O} \leftrightarrow \text{OH}^- + \text{H}^+$	14	(Skoog et al., 2014)
Phosphoric acid	$\text{H}_3\text{PO}_4 \leftrightarrow \text{H}_2\text{PO}_4^- + \text{H}^+$	2.1	(Ohlinger et al., 1998)
	$\text{H}_2\text{PO}_4^- \leftrightarrow \text{HPO}_4^{2-} + \text{H}^+$	7.2	(Ohlinger et al., 1998)
	$\text{HPO}_4^{2-} \leftrightarrow \text{PO}_4^{3-} + \text{H}^+$	12.35	(Ohlinger et al., 1998)
Carbonic acid	$\text{H}_2\text{CO}_3 \leftrightarrow \text{HCO}_3^- + \text{H}^+$	6.35	(Skoog et al., 2014)
	$\text{HCO}_3^- \leftrightarrow \text{CO}_3^{2-} + \text{H}^+$	10.33	(Skoog et al., 2014)

Table 6S: Solids species considered in the thermodynamic model for estimating the formation of struvite.

Name	Chemical system	pK_{sp}	Source
Struvite	$\text{MgNH}_4\text{PO}_4 \cdot 6\text{H}_2\text{O} \leftrightarrow \text{Mg}^{2+} + \text{NH}_4^+ + \text{PO}_4^{3-}$	13.26	(Ohlinger et al., 1998)
K-struvite	$\text{MgKPO}_4 \cdot 6\text{H}_2\text{O} \leftrightarrow \text{Mg}^{2+} + \text{K}^+ + \text{PO}_4^{3-}$	10.6	(Taylor et al., 1963)
Hydroxyapatite	$\text{Ca}_5(\text{PO}_4)_3\text{OH} \leftrightarrow 5\text{Ca}^{2+} + 3\text{PO}_4^{3-} + \text{OH}^-$	44.33	(Brezonik & Arnold, 2011)
Calcium carbonate	$\text{CaCO}_3 \leftrightarrow \text{Ca}^{2+} + \text{CO}_3^{2-}$	8.48	(Morse et al., 2007)
Tricalcium phosphate	$\text{Ca}_3(\text{PO}_4)_2 \leftrightarrow 3\text{Ca}^{2+} + 2\text{PO}_4^{3-}$	25.50	(Fowler & Kuroda, 1986)
Dicalcium phosphate	$\text{CaHPO}_4 \leftrightarrow \text{Ca}^{2+} + \text{HPO}_4^{2-}$	6.57	(Gregory et al., 1970)
Calcium hydroxide	$\text{Ca}(\text{OH})_2 \leftrightarrow \text{Ca}^{2+} + 2\text{OH}^-$	5.19	(Skoog et al., 2014)
Magnesium hydroxide	$\text{Mg}(\text{OH})_2 \leftrightarrow \text{Mg}^{2+} + 2\text{OH}^-$	11.15	(Skoog et al., 2014)

A correlation to estimate the formation of struvite through the fraction of phosphorus recovered as struvite $\left(x_{\text{struvite}(\text{PO}_4^{3-})}\right)$ as a function of calcium concentration in the feedstock is used in this work, Eq. 14S ([Martín-Hernández et al., 2020](#)). Based on the correlation shown in Eq. 14S, the amount of nitrogen recovered is estimated through Eq. 16S.

$$x_{\text{struvite}}(\text{PO}_4^{3-}) = \frac{0.798}{1 + \left(x_{\text{Ca}^{2+}:\text{PO}_4^{3-}} \cdot 0.576\right)^{2.113}} \quad (14\text{S})$$

$$\dot{m}_{\text{struvite recovered}} = \dot{m}_{\text{P in}} \cdot x_{\text{struvite}}(\text{PO}_4^{3-}) \cdot \frac{MW_{\text{struvite}}}{MW_{\text{P}}} \quad (15\text{S})$$

$$\dot{m}_{\text{N recovered}} = \dot{m}_{\text{struvite recovered}} \cdot \frac{MW_{\text{N}}}{MW_{\text{struvite}}} \quad (16\text{S})$$

$$\dot{m}_{\text{N released}} = \dot{m}_{\text{N in}} - \dot{m}_{\text{struvite recovered}} \cdot \frac{MW_{\text{N}}}{MW_{\text{struvite}}} \quad (17\text{S})$$

S1.5 MAPHEX

The mass balances of MAPHEX process are shown in Eqs. 18S-22S, based on experimental data for the full-scale modular units reported by Church et al. (2018).

$$\dot{m}_{\text{nutrients recovered}} = \dot{m}_{\text{nutrients, in}} \cdot \eta_{\text{MAPHEX}}^{\text{nutrients}} \quad \forall \text{ nutrients} \in \{\text{P, N}\} \quad (18\text{S})$$

$$\dot{m}_{\text{nutrients released}} = \dot{m}_{\text{nutrients, in}} \cdot (1 - \eta_{\text{MAPHEX}}^{\text{nutrients}}) \quad \forall \text{ nutrients} \in \{\text{P, N}\} \quad (19\text{S})$$

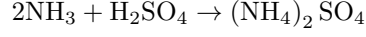
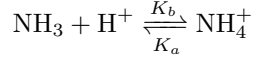
$$\dot{m}_{\text{TS recovered}} = \dot{m}_{\text{TS, in}} \cdot 0.93 \quad (20\text{S})$$

$$\dot{m}_{\text{water recovered}} = \frac{\dot{m}_{\text{TS recovered}}}{0.25} \cdot 0.75 \quad (21\text{S})$$

$$\dot{m}_{\text{water released}} = \dot{m}_{\text{water, in}} - \dot{m}_{\text{water recovered}} \quad (22\text{S})$$

S1.6 Transmembrane chemisorption

Transmembrane chemisorption was modeled adapting the model proposed by Rongwong and Sairiam (2020), considering mass transfer resistances on digestate liquid, membrane, and permeate phases, species distribution of the ammonia system, and the degree of membrane wetting. Mass transfer resistances on the permeate are considered negligible since ammonia is rapidly converted into ammonium sulfate as a consequence of the excess concentration of sulfuric acid in the permeate stream.



The main parameters for membrane modeling are collected in Eqs. 23S-36S. Digestate density (ρ) and (μ) viscosity were assumed equal to those of water. Correlations to estimate these parameters are reported by Hsu and Li (1997). Ammonia Henry's constant (H_{NH_3}) have been taken from Linstrom and Mallard (2001).

$$\rho_{\text{water}} \left(\frac{\text{kg}}{\text{m}^3} \right) = (10^3) \cdot (0.863559 + 1.21494 \cdot 10^{-3} \cdot T(\text{K}) - 2.5708 \cdot 10^{-6} \cdot T(\text{K})^2) \quad (23\text{S})$$

$$\mu_{\text{water}} \left(\frac{\text{kg}}{\text{m} \cdot \text{s}} \right) = (10^{-6}) \cdot \left(\rho \cdot e^{-3.28285 + \frac{456.029}{T(\text{K}) - 154.576}} \right) \quad (24\text{S})$$

$$H_{\text{NH}_3} \left(\frac{\text{kmol}}{\text{atm} \cdot \text{m}^3} \right) = 58 \cdot e^{4100 \cdot \left(\frac{1}{T(\text{K})} - \frac{1}{298.15} \right)} \quad (25\text{S})$$

The cross-sectional area of the lumen area is calculated based on the transverse area of the hollow fiber, and the number of fibers in the membrane module, Eq. 27S. Shell area is calculated as the difference between total and lumen side cross-sectional areas. Velocities of digestate and stripping fluid (i.e., sulfuric acid solution) are computed based on their respective volumetric flow and transverse areas.

$$n_{f_{\text{module}}} = A_{\text{module}} \cdot n_f \quad (26\text{S})$$

$$A_{\text{lumen side}} (\text{m}^2) = \frac{d_i^2}{4} \cdot \pi \cdot n_{f_{\text{module}}} \quad (27\text{S})$$

$$A_{\text{shell side}} (\text{m}^2) = \frac{D_{\text{module}}^2}{4} \cdot \pi - A_{\text{lumen side}} \quad (28\text{S})$$

$$V_{\text{stripping fluid}} \left(\frac{\text{m}}{\text{s}} \right) = \frac{\dot{V}_{\text{stripping fluid}}}{A_{\text{lumen side}}} \quad (29\text{S})$$

$$V_{\text{digestate}} \left(\frac{\text{m}}{\text{s}} \right) = \frac{\dot{V}_{\text{digestate}}}{A_{\text{shell side}}} \quad (30\text{S})$$

Diffusion of ammonia through the membrane is driven by bulk and Knudsen diffusivities, which consider the mean free path of molecules and the pore diameter respectively. Bulk diffusivity of ammonia is calculated based on the kinetic theory, using Eq. 32S, while Knudsen diffusivity depends on the molecular velocity, Eq. 31S, and pore diameter, Eq. 34S Agrahari et al. (2012). Both contributions to ammonia transport are combined in an effective diffusion coefficient, Eq. 35S. Ammonia diffusion in liquid phase has been assumed equal to ammonia diffusion in water, which is calculated using Eq. 36S.

$$\bar{v} \left(\frac{\text{m}}{\text{s}} \right) = \left(\frac{8 \cdot R \left(\frac{\text{J}}{\text{mol} \cdot \text{K}} \right) \cdot T(\text{K})}{MW_{\text{NH}_3} \left(\frac{\text{kg}}{\text{mol}} \right) \cdot \pi} \right)^{1/2} \quad (31\text{S})$$

$$D_{\text{NH}_3} \left(\frac{\text{m}^2}{\text{s}} \right) = \frac{1}{3} \cdot \lambda \cdot \bar{v} \quad (32\text{S})$$

$$\lambda(\text{m}) = \frac{\kappa \left(\frac{\text{J}}{\text{K}} \right) \cdot T(\text{K})}{\pi \cdot \sigma_{\text{NH}_3}^2(\text{m}^2) \cdot P(\text{Pa}) \cdot 2^{1/2}} \quad (33\text{S})$$

$$D_{\text{NH}_3\text{Kn}} \left(\frac{\text{m}^2}{\text{s}} \right) = \frac{d_{\text{pore}}}{3} \cdot \left(\frac{8 \cdot R \left(\frac{\text{J}}{\text{mol} \cdot \text{K}} \right) \cdot T(\text{K})}{\pi \cdot MW_{\text{NH}_3} \left(\frac{\text{kg}}{\text{mol}} \right)} \right)^{1/2} \quad (34\text{S})$$

$$D_{\text{eff}} \left(\frac{\text{m}^2}{\text{s}} \right) = \frac{1}{\frac{1}{D_{\text{NH}_3\text{Kn}}} + \frac{1}{D_{\text{NH}_3}}} \quad (35\text{S})$$

$$D_{\text{NH}_3\text{water}} \left(\frac{\text{m}^2}{\text{s}} \right) = 1.96 \cdot 10^{-9} \cdot \frac{\mu_{\text{water}_{T=298\text{K}}}}{\mu_{\text{water}_T}} \cdot \frac{T(\text{K})}{298} \quad (36\text{S})$$

The distribution of ammonia species is calculated at initial conditions based on the pH of the digestate stream being fed through the tube side of the membrane module, which is assumed equal to 11, as shown in Eqs. 37S-41S.

$$f_{\text{NH}_3} = \frac{1}{10^{(pK_{a_{\text{NH}_3}} - pH)} + 1} \quad (37\text{S})$$

$$pK_{a_{\text{NH}_3}} \left(\frac{\text{kmol}}{\text{m}^3} \right) = 0.0901821 + \frac{2729.92}{T(\text{K})} \quad (38\text{S})$$

$$[\text{NH}_3] = [\text{N}_{\text{inorg}}] \cdot f_{\text{NH}_3} \quad (39\text{S})$$

$$[\text{NH}_4^+] = [\text{N}_{\text{inorg}}] \cdot (1 - f_{\text{NH}_3}) \quad (40\text{S})$$

$$[\text{H}^+] = K_a \cdot \frac{[\text{NH}_4^+]}{[\text{NH}_3]} \quad (41\text{S})$$

A modified Henry's constant considering the speciation of ammonia is used in the model proposed by [Rongwong and Sairiam \(2020\)](#) to evaluate the liquid-solid equilibrium of ammonia, Eq. 42S.

$$H_{\text{NH}_3}^* = H_{\text{NH}_3} \cdot \left(1 + \frac{K_a \cdot [\text{H}^+]}{K_w} \right) \quad (42\text{S})$$

The mass transfer coefficient of the digestate flowing in the tube side, $k_{\text{digestate}}$, is calculated through the Sherwood number of this stream, Eq. 43S.

$$\text{Sh}_{\text{digestate}} = \frac{k_{\text{digestate}} \cdot d_i}{D_{\text{NH}_3\text{water}}} = 1.62 \cdot \left(\frac{d_i}{L} \cdot \text{Re}_{\text{digestate}} \cdot \text{Sc}_{\text{digestate}} \right)^{1/3} \quad (43\text{S})$$

$$\text{Re}_{\text{digestate}} = \frac{d_i \cdot \rho_{\text{digestate}} \cdot V_{\text{digestate}}}{\mu_{\text{digestate}}} \quad (44\text{S})$$

$$\text{Sc}_{\text{digestate}} = \frac{\mu_{\text{digestate}}}{\rho_{\text{digestate}} \cdot D_{\text{NH}_3\text{water}}} \quad (45\text{S})$$

The membrane phase can be divided into non-wetted and wetted membrane sections. Their mass transfers coefficients, k_M and k'_M respectively, are calculated based on the fraction of membrane wetted by the digestate. This is calculated based on a correlation reported by [Rongwong and Sairiam \(2020\)](#), shown in Eq. 46S. Eqs. 48S-49S and 50S-51S shown the equations employed

for the estimation of k_M and k'_M respectively.

$$w_{\text{membrane}} = \frac{0.368}{2.709 + e^{-17.486 \cdot (V_{\text{digestate}} - 0.25)}} \quad (46S)$$

$$d_{\text{interfacial}} (\text{m}) = d_i + w_{\text{membrane}} \cdot (d_o - d_i) \quad (47S)$$

$$k_M \left(\frac{\text{m}}{\text{s}} \right) = \frac{D_{\text{eff}} \cdot \varepsilon}{\tau \cdot \delta_{\text{non-wetted}}} \quad (48S)$$

$$\delta_{\text{non-wetted}} (\text{m}) = \frac{d_o - d_{\text{interfacial}}}{2} \quad (49S)$$

$$k'_M \left(\frac{\text{m}}{\text{s}} \right) = \frac{D_{\text{water}} \cdot \varepsilon}{\tau \cdot \delta_{\text{wetted}}} \quad (50S)$$

$$\delta_{\text{wetted}} (\text{m}) = \frac{d_{\text{interfacial}} - d_i}{2} \quad (51S)$$

The overall mass transfer coefficient is calculated based on the liquid, wetted membrane, and non-wetted membranes mass transfer coefficients, as shown in Eq. 52S. However, high ammonia concentrations result in a decrease of the overall mass transfer (Zhu et al., 2005). Therefore, K_{ov} is modified using a correction factor that considers the concentration of ammonia in digestate.

$$K_{ov} \left(\frac{\text{m}}{\text{s}} \right) = \frac{1}{\left(\frac{1}{k_{\text{digestate}} \cdot d_i} + \frac{1}{k'_M \cdot \delta_{\text{ln wetted}}} + \frac{R \left(\frac{\text{atm} \cdot \text{m}^3}{\text{kmol} \cdot \text{K}} \right) \cdot T(\text{K}) \cdot H_{\text{NH}_3}^*}{k_M \delta_{\text{ln non-wetted}}} \right) \cdot d_o} \quad (52S)$$

$$\delta_{\text{ln wetted}} = \frac{d_{\text{interfacial}} - d_i}{\ln \left(\frac{d_{\text{interfacial}}}{d_i} \right)} \quad (53S)$$

$$\delta_{\text{ln non-wetted}} = \frac{d_o - d_{\text{interfacial}}}{\ln \left(\frac{d_o}{d_{\text{interfacial}}} \right)} \quad (54S)$$

$$K_{ov \text{corrected}} \left(\frac{\text{m}}{\text{s}} \right) = K_{ov} \cdot \left(1 - \frac{0.002728 \cdot [\text{NH}_3] \left(\frac{\text{mg}}{\text{L}} \right)}{100} \right) \quad (55S)$$

The mass balance along the membrane can be performed considering the mass transfer in an infinitesimal element of the membrane, as shown in Eq. 56S. Due to the excess of sulfuric acid in the lumen side that lead to the formation of ammonium sulfate from the ammonia recovered, the concentration of ammonia in the lumen side is assumed negligible. This differential expression can be analytically integrated to estimate the total length of the membrane, Eq. 57S.

$$\begin{aligned}\frac{\dot{n}_{\text{NH}_3}}{dz} &= -n_{f_{\text{module}}} \cdot \pi \cdot d_o \cdot K_{ov_{\text{corrected}}} \cdot \left(\frac{\dot{n}_{\text{NH}_3}}{\dot{V}_{\text{digestate}}} - \frac{H_{\text{NH}_3}^* \cdot p_{\text{NH}_3}}{\dot{V}_{\text{digestate}}} \right) \\ &\approx -n_{f_{\text{module}}} \cdot \pi \cdot d_o \cdot K_{ov_{\text{corrected}}} \cdot \frac{\dot{n}_{\text{NH}_3}}{\dot{V}_{\text{digestate}}}\end{aligned}\quad (56\text{S})$$

$$z \text{ (m)} = \ln \left(\frac{\dot{n}_{\text{NH}_3_{\text{final}}}}{\dot{n}_{\text{NH}_3_{\text{initial}}}} \right) \cdot \frac{\dot{V}_{\text{digestate}}}{-n_{f_{\text{module}}} \cdot \pi \cdot d_o \cdot K_{ov_{\text{corrected}}}} \quad (57\text{S})$$

Characteristics, sizes, and costs of Liquid-Cel[™] Extra Flow membranes are shown in Tables 7S and 8S. All membrane module costs were taken from values reported by manufacturers (SG Projects, 2021; DPC Water Solutions, 2021), except for the Liquid-Cel[™] 14x40 module, which price was estimated based on the cost of the remaining modules, as illustrated in Figure 3S

Table 7S: Main characteristics of Liquid-Cel[™] membrane (Rongwong & Sairiam, 2020; Ulbricht et al., 2013).

Membrane parameters	Unit	Variable	Values
Number of fibers	$\frac{\text{fibers}}{\text{m}^2_{\text{module}}}$	n_f	$2.13 \cdot 10^6$
Fiber outer diameter	m	d_o	$300 \cdot 10^{-6}$
Fiber inner diameter	m	d_i	$240 \cdot 10^{-6}$
Pore diameter	m	a	$0.04 \cdot 10^{-6}$
Porosity	-	ε	0.40
Tortuosity factor	-	τ	2.60
Pressure drop shell side	bar	$\Delta P_{\text{shell side}}$	0.3
Pressure drop lumen side	bar	$\Delta P_{\text{lumen side}}$	2

Table 8S: Liquid-Cel[™] membranes size and cost (3M, 2021; SG Projects, 2021; DPC Water Solutions, 2021).

Membrane model	Flow capacity (m3/h)	Diameter (m)	Lenght (m)	Cost (USD)
2.5×8	0.1 - 0.7	0.067	0.200	5,000
4×13	0.5 - 3.4	0.116	0.242	5,700
8x20	1 - 11	0.219	0.406	14,150
10x28	10 - 48	0.279	0.683	17,000
14x40	16 - 125	0.356	1.129	24,300

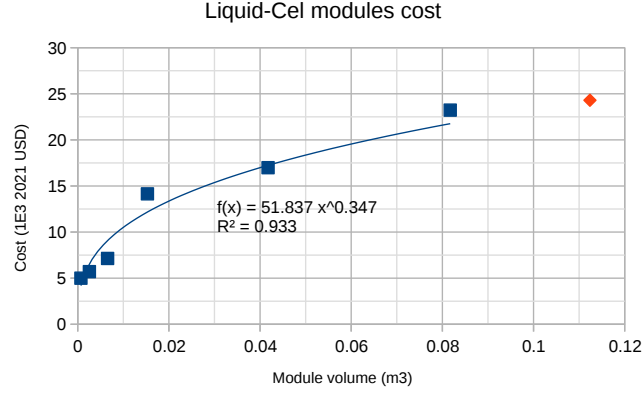


Figure 3S: Estimation of Liquid-CelTM 14x40 membrane module cost.

In the case that the capacity of the largest membrane module is not enough for the treatment of digestate, the installation of several parallel units is considered, as shown in Eq. 58S.

$$n_{\text{parallel}} = \begin{cases} 1 & \text{if } \dot{V}_{\text{digestate}} \leq 125 \frac{\text{m}^3}{\text{h}} \\ \left\lceil \frac{\dot{V}_{\text{digestate}} \left(\frac{\text{m}^3}{\text{h}} \right)}{125} \right\rceil & \text{if } \dot{V}_{\text{digestate}} > 125 \frac{\text{m}^3}{\text{h}} \end{cases} \quad (58\text{S})$$

Membrane module CAPEX is estimated through Eq. 60S, assuming a membrane lifetime (t_{module}) of 10 years (Verrecht et al., 2010), and a plant lifetime (t_{plant}) of 20 years. The use of sulfuric acid as stripping fluid and membrane cleaning is the main contributor to membrane OPEX, as shown in Eq. 61S. Membrane cleaning cost (c_{cleaning}) is reported to be between 2% and 25% of the operating costs (Yu et al., 2020; Verrecht et al., 2010), for which we assume an average value of 13.5%.

$$n_{\text{series}} = \left\lceil \frac{z}{L_{\text{module}}} \right\rceil \quad (59\text{S})$$

$$CAPEX_{\text{membranes}} \left(\frac{\text{USD}}{\text{module}} \right) = (n_{\text{serie}} \cdot n_{\text{parallel}}) \cdot Cost_{\text{module}} \left(\frac{\text{USD}}{\text{module}} \right) \cdot \frac{t_{\text{plant}}}{t_{\text{module}}} \quad (60\text{S})$$

$$OPEX_{\text{membranes}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = \frac{\dot{m}_{\text{H}_2\text{SO}_4} \left(\frac{\text{kg}}{\text{s}} \right)}{\rho_{\text{H}_2\text{SO}_4} \left(\frac{\text{kg}}{\text{m}^3} \right)} \cdot Price_{\text{H}_2\text{SO}_4} \left(\frac{\text{USD}}{\text{m}^3} \right) \cdot \frac{1}{1 - \frac{c_{\text{cleaning}}}{100}} \quad (61\text{S})$$

Pumps to drive the digestate and stripping fluid streams are needed. Their cost is estimated using Eqs. 62S and 63S, assuming that the maximum capacity of one single pump ($\dot{V}_{\text{pump}_{\text{max}}}$) is 1 m³/s (Peters et al., 2003). Pumps operating cost are estimated through Eqs. 64S to 66S, calculating the energy needed to balance the pressure drop of the shell and lumen sides of the membrane, Table 7S. Pump efficiencies (η_{pump_i}) of 0.8 are assumed.

$$n_{\text{pump}_i} = \left\lceil \frac{\dot{V}_i}{\dot{V}_{\text{pump}_{\text{max}}}} \right\rceil \quad (62\text{S})$$

$$CAPEX_{\text{pumps}_i} \text{ (2019 USD)} = \begin{cases} (-50.387 \cdot 10^6 \cdot \dot{V}_i^2 + 1.607 \cdot 10^6 \cdot \dot{V}_i \\ + 5.127 \cdot 10^3) \cdot n_{\text{pump}_i} \cdot 1.055 & \text{if } \dot{V}_i \leq 9 \cdot 10^{-3} \frac{\text{m}^3}{\text{s}} \\ (92.562 \cdot 10^3 \cdot \dot{V}_i^{0.381}) \cdot n_{\text{pump}_i} \cdot 1.055 & \text{if } \dot{V}_i > 9 \cdot 10^{-3} \frac{\text{m}^3}{\text{s}} \end{cases} \quad (63\text{S})$$

$$H_i \text{ (m)} = \frac{\Delta P_i \text{ (Pa)}}{\rho \left(\frac{\text{kg}}{\text{m}^3} \right) g \left(\frac{\text{m}}{\text{s}^2} \right)} \quad (64\text{S})$$

$$\dot{W}_{\text{pump}_i} \text{ (kW)} = \frac{g \left(\frac{\text{m}}{\text{s}^2} \right) \cdot H_i \text{ (m)} \cdot \rho \left(\frac{\text{kg}}{\text{m}^3} \right) \cdot \dot{V}_i \left(\frac{\text{m}^3}{\text{s}} \right)}{\eta_{\text{pump}_i}} \quad (65\text{S})$$

$$OPEX_{\text{pumps}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = \sum_i \dot{W}_{\text{pump}_i} \text{ (kW)} \cdot t_{\text{operation}} \text{ (s)} \cdot 3600 \cdot \text{Price}_{\text{electricity}} \left(\frac{\text{USD}}{\text{kWh}} \right), \quad (66\text{S})$$

$$i \in \{\text{shellside, lumenside}\}$$

Total capital and operating expenses for the recovery of nitrogen from livestock digestate using a transmembrane chemisorption process result from the sum of membrane and pump costs are described in Eqs 67S and 68S.

$$CAPEX_{\text{transmembrane chemisorption}} = CAPEX_{\text{membranes}} + \sum_{\substack{i \in \\ \{\text{shellside,} \\ \text{lumenside}\}}} CAPEX_{\text{pumps}_i} \quad (67\text{S})$$

$$OPEX_{\text{transmembrane chemisorption}} = OPEX_{\text{membranes}} + OPEX_{\text{pumps}} \quad (68\text{S})$$

S1.7 Ammonia evaporation

The following assumptions have been made for the modeling of digestate drying:

- $c_{p\text{digestate}_{\text{in}}} \approx c_{p\text{water}}$
- $P_{\text{water}} + P_{\text{NH}_3} \approx P_{\text{water}}$
- $T_{\text{outair}} = T_{\text{outdigestate}}$
- $\alpha = \text{constant}$

The mass and energy balances for digestate drying in a belt dryer unit are shown in Eqs. [69S-71S](#) and [72S-81S](#) respectively

$$\dot{m}_{\text{NH}_3 \text{ evap}} = \dot{m}_{\text{NH}_3 \text{ in}} \cdot \eta_{\text{belt dryer}}^{\text{NH}_3} \quad (69\text{S})$$

$$\frac{\dot{m}_{\text{water evap}}}{\sum_j \frac{\dot{m}_j}{MW_j}} \cdot P_{\text{total}} \leq P_{v\text{water}}(T), \quad j \in \{\text{air}, \text{H}_2\text{O}, \text{NH}_3\} \quad (70\text{S})$$

$$\frac{\dot{m}_{\text{NH}_3 \text{ evap}}}{\sum_j \frac{\dot{m}_j}{MW_j}} \leq P_{v\text{NH}_3}(T), \quad j \in \{\text{air}, \text{H}_2\text{O}, \text{NH}_3\} \quad (71\text{S})$$

$$\dot{Q}_{\text{Belt Dryer}} = \dot{H}_{\text{latent}} + \dot{H}_{\text{sensible}} = \dot{H}_{\text{air}_{\text{in}}} - \dot{H}_{\text{air}_{\text{out}}} \quad (72\text{S})$$

$$\dot{H}_{\text{latent}} = \dot{m}_{\text{water evap}} \cdot \lambda_{\text{evap}_{\text{water}}}(T) + \dot{m}_{\text{NH}_3 \text{ evap}} \cdot \lambda_{\text{evap}_{\text{NH}_3}}(T) \quad (73\text{S})$$

$$\dot{H}_{\text{sensible}} = \dot{H}_{\text{digestate}_{\text{out}}} - \dot{H}_{\text{digestate}_{\text{in}}} + \dot{H}_{\text{sensible}_{\text{gases}}} \quad (74\text{S})$$

$$\dot{H}_{\text{sensible}_{\text{gases}}} = \left(\dot{m}_{\text{water evap}} + \dot{m}_{\text{NH}_3 \text{ evap}} \right) \cdot c_{p\text{water (liquid)}} \cdot T_{\text{air}_{\text{out}}} \quad (75\text{S})$$

$$\dot{H}_{\text{digestate}_{\text{in}}} = \dot{m}_{\text{digestate}_{\text{in}}} \cdot c_{p\text{water}} \cdot \Delta T_{\text{digestate}_{\text{in}}} \quad (76\text{S})$$

$$\dot{H}_{\text{digestate}_{\text{out}}} = \dot{m}_{\text{digestate}_{\text{out}}} \cdot c_{p\text{digestate}} \cdot \Delta T_{\text{digestate}_{\text{out}}} \quad (77\text{S})$$

$$c_{p\text{digestate}} (kg/kJ \cdot K) = 4.19 - 0.0275 \cdot \left(\frac{\dot{m}_{\text{TS}_{\text{out}}}}{\dot{m}_{\text{digestate}_{\text{out}}}} \cdot 100 \right) \quad (78\text{S})$$

$$\dot{H}_{\text{airin}} = \dot{m}_{\text{air}} \cdot c_{p\text{air}} \cdot \Delta T_{\text{airin}} \quad (79\text{S})$$

$$\dot{H}_{\text{airout}} = \left(\dot{m}_{\text{air}} \cdot c_{p\text{air}} + \dot{m}_{\text{water evap}} \cdot c_{p\text{water (gas)}} + \dot{m}_{\text{NH}_3 \text{evap}} \cdot c_{p\text{NH}_3 \text{(gas)}} \right) \cdot \Delta T_{\text{airout}} \quad (80\text{S})$$

$$\Delta T_i = T_i - T_{\text{ref}} \quad \forall i \in \{\text{airin}, \text{airout}, \text{digestatein}, \text{digestateout}\} \quad (81\text{S})$$

Gaseous ammonia is further recovered through acidic scrubbing, which is described below.

CAPEX estimation for belt dryer units is based on the energy required for ammonia evaporation. A drying efficiency ($\eta_{\text{belt dryer}}$) of 0.6 is assumed from the experimental work reported by [Awiszus et al. \(2018\)](#). Belt dryer scale-up is based on the correlation proposed by [Towler and Sinnott \(2012\)](#), as shown in Eq. 85S. The reference values and scale factor used in this correlation are taken from costs and capacities reported by [Turley et al. \(2016\)](#), as well as the maximum belt dryer capacity used to compute the number of dryer units needed ($n_{\text{belt dryer}}$), as shown in Eq. 83S. These data are also used to estimate the scale-up factor, which is estimated equal to 0.7. Belt dryer operating costs are due to electrical consumption, which was estimated in 0.099 kW of electricity per kW of thermal energy used by the unit, as shown in Eq. 86S ([Awiszus et al., 2018](#)).

$$\dot{Q}_{\text{belt dryer}}^{\text{real}} = \frac{\dot{Q}_{\text{belt dryer}}^{\text{design}} (\text{kW})}{\eta_{\text{belt dryer}}} \quad (82\text{S})$$

$$n_{\text{belt dryer}} = \left\lceil \frac{\dot{Q}_{\text{belt dryer}}^{\text{real}} (\text{kW})}{1000} \right\rceil \quad (83\text{S})$$

$$\dot{Q}_{\text{belt dryer}}^{\text{design}} = \frac{\dot{Q}_{\text{belt dryer}}^{\text{real}}}{n_{\text{belt dryer}}} \quad (84\text{S})$$

$$CAPEX_{\text{belt dryer}} (2019 \text{ USD}) = \left(n_{\text{belt dryer}} \cdot 214,997 \cdot \left(\frac{\dot{Q}_{\text{belt dryer}}^{\text{design}} (\text{kW})}{500} \right)^{0.7} \right) \cdot 1.216 \quad (85\text{S})$$

$$OPEX_{\text{belt dryer}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = n_{\text{belt dryer}} \cdot \dot{Q}_{\text{belt dryer}}^{\text{design}} (\text{kW}) \cdot 0.099 \left(\frac{\text{kW}_e}{\text{kW}_t} \right). \quad (86\text{S})$$

$$t_{\text{operation}} (\text{s}) \cdot 3600 \cdot Price_{\text{electricity}} \left(\frac{\text{USD}}{\text{kWh}} \right)$$

S1.8 Stripping in packed tower

Nitrogen recovery using packed tower stripping has been modeled using the number of transfer units (NTU) method (Metcalf & Eddy, 2014). The pressure drop of tower packing is estimated through the correlation proposed by Kister and Gill (1991), as shown in Eq. 87S.

$$P \left(\frac{\text{inch H}_2\text{O}}{\text{ft}} \right) = 0.115 \cdot (F_P \text{ (ft}^{-1}\text{)})^{0.7} \quad (87\text{S})$$

The tower diameter is calculated through the tower flooding capacity. A tower flooding capacity correlation considering the packing pressure drop has been developed by using ALAMO (Wilson & Sahinidis, 2017) based on the flooding curves developed by Strigle (1994), as shown in Eq. 88S and Figure 3. Two-inch (0.051 m) Intalox packing is considered (Strigle, 1994), which packing factor (F_P) is assumed to be 18 ft⁻¹ (59 m⁻¹) (Geankoplis, 2003). The operating line considered, defined as the ratio of gas and liquid volumetric flows, can be computed through Eqs. 88S to 91S, where P is in inch H₂O/ft, v_G denotes the superficial gas velocity in ft/s, ρ_G the gas density in lb/ft³, ρ_L the liquid density in lb/ft³, ν the kinematic viscosity in centistokes, G_G the gas mass velocity in lb/ft² · s, and G_L the liquid mass velocity in lb/ft² · s (Strigle, 1994).

$$Y = -0.25 \cdot X + 0.22 \cdot \ln(P) - 0.78 \cdot 10^{-1} \cdot P^2 + 0.19 \cdot 10^{-1} \cdot X^3 - 0.39 \cdot X \cdot P \quad (88\text{S})$$

$$+ 0.49 \cdot 10^{-2} \cdot (X \cdot P)^3 + 0.89$$

$$Y = v_G \left(\frac{\rho_G}{\rho_L - \rho_G} \right)^{0.5} F_P^{0.5} \nu^{0.05} \quad (89\text{S})$$

$$X = \log_{10} \left(\frac{G_L}{G_G} \left(\frac{\rho_G}{\rho_L} \right)^{0.5} \right) \quad (90\text{S})$$

$$\frac{\dot{V}_G}{\dot{V}_L} = 2688 \frac{\text{m}_{\text{gas}}^3}{\text{m}_{\text{liquid}}^3} \quad (91\text{S})$$

$$G_{G_{\text{design}}} = 0.7 G_G = 0.7 \cdot v_G \cdot \rho_G \quad (92\text{S})$$

The liquid mass velocity is a known parameter since it corresponds to the digestate being

processed. The gas velocity is estimated by combining Eqs. 87S, 88S and 91S. The design gas mass velocity considered is 0.7 time the gas mass velocity, Eq 92S, while the liquid design mass velocity is computed by combining Eqs. 92S and 91S. Tower diameter is computed based on the design liquid mass velocity and liquid mass flow. However, design restrictions reported by Branan (2005) have been considered in the sizing of the packed tower. Tower height is estimated through the height and number of transfer units, Eqs 93S and 94S. Therefore, the design diameter is computed as shown in Eqs. 95S to 97S. Additionally, the number of packed towers installed in-parallel arrangement is estimated through Eq. 98S.

$$D_{\text{design}} \leq 1.2 \text{ m} \quad (93\text{S})$$

$$\frac{H_{\text{design}}}{D_{\text{design}}} \leq 30 \quad (94\text{S})$$

$$A_{\text{stripping tower}} = \frac{\dot{m}_L}{G_{L\text{design}}} \quad (95\text{S})$$

$$D_{\text{stripping tower}} = \left(\frac{A_{\text{stripping tower}}}{\pi} \right)^{0.5} \quad (96\text{S})$$

$$D_{\text{stripping tower}}^{\text{design}} (\text{m}) = \begin{cases} D_{\text{stripping tower}} & \text{if } D_{\text{stripping tower}} \leq 1.2 \text{ m} \\ 1.2 & \text{if } D_{\text{stripping tower}} > 1.2 \text{ m} \end{cases} \quad (97\text{S})$$

$$n_{\text{stripping tower}}^{\text{parallel}} = \left\lceil \frac{D_{\text{stripping tower}}^{\text{design}} (\text{m})}{1.2} \right\rceil \quad (98\text{S})$$

The height of the stripping packed towers is estimated through the height and number of transfer units, Eq. 99S and 100S respectively (Metcalf & Eddy, 2014). The liquid overall mass transfer coefficient value assumed for the ammonia air system is 2 h^{-1} (Larsen et al., 2013). The design restriction shown in Eq. 94S is considered to compute the design height of the stripping towers, and the number of units that must be installed in series, Eq. 103S.

$$HTU = \frac{\frac{\dot{m}_L}{n_{\text{stripping tower}}^{\text{parallel}} \cdot \rho_{\text{digestate}}}}{K_L a \cdot \pi \cdot \left(\frac{D_{\text{stripping tower}}}{2} \right)^2} \quad (99S)$$

$$NTU = \frac{S}{S-1} \ln \left(\frac{1}{1 - \eta_{\text{stripping tower}/100}} \cdot (S-1) + 1 \right) \quad (100S)$$

$$S = \frac{H_{\text{NH}_3}}{P_{\text{stripping tower}}} \cdot \frac{\dot{m}_G}{\dot{m}_L} \quad (101S)$$

$$H_{\text{stripping tower}} = NTU \cdot HTU \quad (102S)$$

$$n_{\text{stripping tower}}^{\text{series}} = \left\lceil \frac{\frac{H_{\text{stripping tower}}}{D_{\text{stripping tower}}^{\text{design}}}}{30} \right\rceil \quad (103S)$$

$$H_{\text{stripping tower}}^{\text{design}} (\text{m}) = \begin{cases} H_{\text{stripping tower}} & \text{if } \frac{H_{\text{stripping tower}}}{D_{\text{stripping tower}}} \leq 30 \\ \frac{\frac{\dot{m}_L}{n_{\text{stripping tower}}^{\text{series}} \cdot n_{\text{stripping tower}}^{\text{parallel}} \cdot \rho_{\text{digestate}}}}{K_L a \cdot \pi \cdot \left(\frac{D_{\text{stripping tower}}}{2} \right)^2} & \text{if } \frac{H_{\text{stripping tower}}}{D_{\text{stripping tower}}} > 30 \end{cases} \quad (104S)$$

The number of stripping units needed ($n_{\text{stripping tower}}$) is calculated as the product of the number stripping units in-series arrangement to satisfy the packed towers height limit ($n_{\text{stripping tower}}^{\text{series}}$) and the number stripping units in-parallel arrangement to process the amount of organic materials generated in the livestock facility under evaluation ($n_{\text{stripping tower}}^{\text{parallel}}$), as shown in Eq. 105S. CAPEX of stripping packed towers is estimated based on the columns volume using a correlation based on data from [Turton, R. \(2010\)](#), as shown Eq. 106S. Additionally, capital expenses of compressor units are estimated based on the correlation reported by [Almena and Martín \(2016\)](#), Eq. 108S. Operating expenses of scrubbing are mainly due to the compression cost, Eq. 109S.

$$n_{\text{stripping tower}} = n_{\text{stripping tower}}^{\text{series}} \cdot n_{\text{stripping tower}}^{\text{parallel}} \quad (105S)$$

$$CAPEX_{\text{stripping tower}} (2019 \text{ USD}) = 1.216 \cdot n_{\text{stripping tower}} \cdot$$

$$\left(-6.157 \cdot (V_{\text{stripping tower}} (\text{m}^3))^2 + 1276.035 \cdot V_{\text{scrubber}} (\text{m}^3) + 4007.619 \right) \quad (106S)$$

$$V_{\text{stripping tower}} = \frac{\left(D_{\text{stripping tower}}^{\text{design}} \right)^2 \cdot \pi}{4} \cdot H_{\text{stripping tower}} \quad (107S)$$

$$CAPEX_{\text{compressor}} (2019 \text{ USD}) = 1.216 \cdot \left(335.27 \cdot \dot{W}_{\text{compressor}} (\text{kW}) + 36211 \right) \quad (108S)$$

$$OPEX_{\text{compressor}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = \dot{W}_{\text{compressor}} (\text{kW}) \cdot t_{\text{operation}} (\text{s}) \cdot 3600 \cdot Price_{\text{electricity}} \left(\frac{\text{USD}}{\text{kWh}} \right) \quad (109S)$$

S1.9 Acidic scrubbing

Mass balances of the scrubber unit consider the water transferred to the gas stream, assuming that saturation is reached, Eq. 111S. Ammonia recovery efficiency ($\eta_{\text{scrubber}}^{\text{NH}_3}$) of full-scale ammonia scrubbers was reported in the range of 40% to 99% (Melse & Ogink, 2005). A typical $\eta_{\text{scrubber}}^{\text{NH}_3}$ of 96% is selected based on the work of Melse and Ogink (2005).

$$\frac{P_{\text{water}}}{Pv_{\text{water}}} = 1 \quad (110S)$$

$$Pv_{\text{water}} = \frac{\frac{\dot{m}_{\text{water out}}^{\text{gas stream}}}{MW_{\text{water}}}}{\sum_j \frac{\dot{m}_j^{\text{gas stream}}}{MW_j}} \cdot P_{\text{in}}^{\text{gas stream}} \quad (111S)$$

$$\dot{m}_{\text{NH}_3 \text{ out}}^{\text{gas stream}} = \dot{m}_{\text{NH}_3 \text{ in}}^{\text{gas stream}} \cdot \left(1 - \eta_{\text{scrubber}}^{\text{NH}_3} \right) \quad (112S)$$

The water flow needed to perform the scrubbing operation is computed from the operating line of the unit (L/G), Eq. 113S. Following the rules of thumb for scrubbing units, the design operating line is assumed as twice the minimum operating line, Eq. 114S. Y_{NH_3} and X_{NH_3} denote the ammonia free basis molar fractions in gas and liquid streams respectively. The amount of sulfuric acid supplied to make-up the sulfate used for ammonium sulfate formation, Eq. 116S, is slightly larger than the stoichiometric amount of the precipitation reaction, 3.5 kg_{H₂SO₄} per

kg_{NH₃} recovered (Bolzonella et al., 2018).

$$(L/G)_{\min} = \frac{Y_{\text{NH}_3 \text{in}} - Y_{\text{NH}_3 \text{out}}}{X_{\text{NH}_3 \text{out}} - X_{\text{NH}_3 \text{in}}} \quad (113\text{S})$$

$$(L/G) = (L/G)_{\min} \cdot 2 \quad (114\text{S})$$

$$\dot{m}_{\text{water}_{\text{in}}}^{\text{liquid stream}} = \left((L/G) \cdot \dot{n}_{\text{total}_{\text{in}}}^{\text{liquid stream}} - \dot{n}_{\text{NH}_3 \text{in}}^{\text{liquid stream}} \right) \cdot MW_{\text{water}} \quad (115\text{S})$$

$$\dot{m}_{\text{H}_2\text{SO}_4 \text{in}}^{\text{liquid stream}} = \dot{n}_{\text{NH}_3 \text{in}}^{\text{gas stream}} \cdot \eta_{\text{scrubber}}^{\text{NH}_3} \cdot 3.5 \quad (116\text{S})$$

$$\dot{m}_{(\text{NH}_4)_2\text{SO}_4 \text{out}}^{\text{liquid stream}} = \frac{\dot{n}_{\text{NH}_3 \text{in}}^{\text{gas stream}} \cdot \eta_{\text{scrubber}}^{\text{NH}_3}}{MW_{\text{NH}_3}} \cdot MW_{(\text{NH}_4)_2\text{SO}_4 \text{out}} \quad (117\text{S})$$

The scrubber diameter, Eq. 119S, is based on the gas velocity in the equipment, which is assumed equal to 1.75 m/s (Melse & Ogink, 2005). The number of units is set by the maximum diameter of scrubbers, Eq. 120S, which is assumed equal to 1.2 m according to the rules of thumb for packed columns (Branan, 2005).

$$D_{\text{scrubber}} (\text{m}) = \frac{\dot{V}_{\text{gas stream}} \left(\frac{\text{m}^3}{\text{s}} \right)}{1.75 \left(\frac{\text{m}}{\text{s}} \right)} \quad (118\text{S})$$

$$D_{\text{scrubber}}^{\text{design}} (\text{m}) = \begin{cases} D_{\text{scrubber}} & \text{if } D_{\text{scrubber}} \leq 1.2 \text{ m} \\ 1.2 & \text{if } D_{\text{scrubber}} > 1.2 \text{ m} \end{cases} \quad (119\text{S})$$

$$n_{\text{scrubbers}} = \left\lceil \frac{\dot{V}_{\text{gas stream}} \left(\frac{\text{m}^3}{\text{s}} \right)}{1.75 \left(\frac{\text{m}}{\text{s}} \right) \cdot \frac{1.2^2 (\text{m}^2)}{4} \pi} \right\rceil \quad (120\text{S})$$

The height of scrubbing units is estimated through the height and number of transfer units, Eq. 121S, as described by Couper et al. (2005). The number of transfer units is determined by the Kremser shortcut method, as shown in Eq. 122S (Seader et al., 2004), while the height of each transfer unit is calculated in Eq. 126S. The gas film overall mass transfer coefficient value for the ammonia air system is 272474.56 $\frac{\text{mol}}{\text{h} \cdot \text{m}^3 \cdot \text{atm}}$ (Branan, 2005). The scrubbing operation involves the use of compressors for pumping the stripping gas stream. The energy required for compression is estimated though Eq. 127S, assuming a compressor efficiency ($\eta_{\text{compressor}}$) of 0.85 and a polytropic coefficient (k) of 1.4. The pressure drop of a typical scrubber for ammonia

capture assumed is 200 Pa (Melse & Ogink, 2005).

$$H_{\text{scrubber}} = NTU \cdot HTU \quad (121S)$$

$$NTU = \frac{\ln \left(\left(1 - m \cdot \frac{\dot{n}^{\text{gas stream}}}{\dot{n}^{\text{liquid stream}}} \right) \frac{x_{\text{NH}_3 \text{ in}}^{\text{gas stream}} - x_{\text{NH}_3}^*}{x_{\text{NH}_3 \text{ out}}^{\text{gas stream}} - x_{\text{NH}_3}^*} + m \cdot \frac{\dot{n}^{\text{gas stream}}}{\dot{n}^{\text{liquid stream}}} \right)}{\ln \left(m \cdot \frac{L}{V} \right)} \quad (122S)$$

$$x_{\text{NH}_3 k}^{\text{gas stream}} = \frac{\dot{n}_{\text{NH}_3 k}^{\text{gas stream}}}{\dot{n}^{\text{gas stream}}}, \forall k \in \{in, out\} \quad (123S)$$

$$x_{\text{NH}_3}^* = 0 \quad (124S)$$

$$m = \frac{P_{\text{NH}_3}}{P} \quad (125S)$$

$$HTU = \frac{\dot{n}^{\text{gas stream}} \left(\frac{\text{mol}}{\text{h}} \right)}{\pi \cdot \frac{D_{\text{scrubber}}^{\text{design}^2}}{4} (\text{m}^2) \cdot k_{GA} \left(\frac{\text{mol}}{\text{h} \cdot \text{m}^3 \cdot \text{atm}} \right) \cdot P (\text{atm})} \quad (126S)$$

$$\dot{W}_{\text{compressor}} (\text{kW}) = \frac{k \cdot R \left(\frac{\text{J}}{\text{K} \cdot \text{kmol}} \right) \cdot T_{in}^{\text{gas stream}} (\text{K}) \cdot \dot{m}_{in}^{\text{gas stream}} \left(\frac{\text{kg}}{\text{s}} \right)}{\eta_{\text{compressor}} \cdot (k - 1) \cdot MW_{gas}} \cdot \left(\left(\frac{P_{in}^{\text{gas stream}}}{P_{out}^{\text{gas stream}}} \right)^{\frac{k-1}{k}} - 1 \right) \quad (127S)$$

Operating expenses of scrubbing are mainly related to the use of sulfuric acid, Eq. 128S, and the compression cost, Eq. 129S.

$$OPEX_{\text{scrubbers}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = \frac{\dot{m}_{\text{H}_2\text{SO}_4} \left(\frac{\text{kg}}{\text{s}} \right)}{\rho_{\text{H}_2\text{SO}_4} \left(\frac{\text{kg}}{\text{m}^3} \right)} \cdot Price_{\text{H}_2\text{SO}_4} \left(\frac{\text{USD}}{\text{m}^3} \right) \quad (128S)$$

$$OPEX_{\text{compressor}} \left(\frac{2019 \text{ USD}}{\text{year}} \right) = \dot{W}_{\text{compressor}} (\text{kW}) \cdot t_{\text{operation}} (\text{s}) \cdot 3600 \cdot Price_{\text{electricity}} \left(\frac{\text{USD}}{\text{kWh}} \right) \quad (129S)$$

S2 Capital and operating expenses

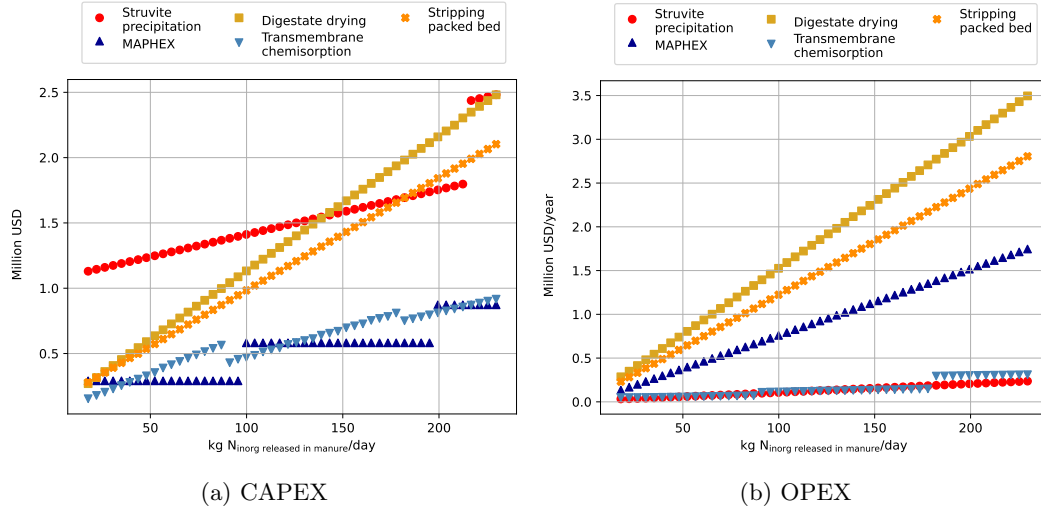


Figure 4S: CAPEX and OPEX of the assessed nitrogen recovery technologies, including the cost of anaerobic digestion stage.

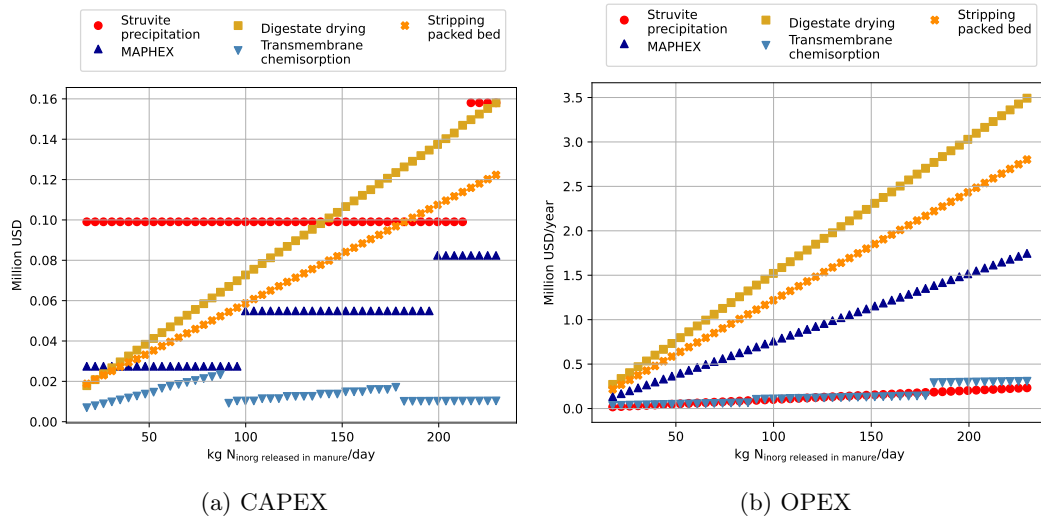


Figure 5S: CAPEX and OPEX of the assessed nitrogen recovery technologies, excluding the cost of anaerobic digestion stage.

References

- 3M. (2021). *3m™liqui-cel™data sheets* (Tech. Rep.). Author.
- Agrahari, G. K., Shukla, S. K., Verma, N., & Bhattacharya, P. K. (2012). Model prediction and experimental studies on the removal of dissolved nh3 from water applying hollow fiber membrane contactor. *Journal of Membrane Science*, *390*, 164–174.
- Almena, A., & Martín, M. (2016). Technoeconomic analysis of the production of epichlorohydrin from glycerol. *Industrial & Engineering Chemistry Research*, *55*(12), 3226–3238.
- Awiszus, S., Meissner, K., Reyer, S., & Müller, J. (2018). Utilization of digestate in a convective hot air dryer with integrated nitrogen recovery. *Landtechnik*, *73*(4), 106–114.
- Bates, R. G., & Pinching, G. (1949). Acid dissociation constant of ammonium ion at 0–50 °C, and the base strength of ammonia. *Journal of Research of the National Bureau of Standards*, *42*, 419–430.
- Beddoes, J. C., Bracmort, K. S., Burns, R. T., & Lazarus, W. F. (2007). *An Analysis of Energy Production Costs from Anaerobic Digestion Systems on U.S. Livestock Production Facilities* (Tech. Rep.). US Department of Agriculture (USDA).
- Bolzonella, D., Fatone, F., Gottardo, M., & Frison, N. (2018). Nutrients recovery from anaerobic digestate of agro-waste: Techno-economic assessment of full scale applications. *Journal of Environmental Management*, *216*, 111–119.
- Branan, C. R. (Ed.). (2005). *Contents* (Fourth Edition ed.). Burlington: Gulf Professional Publishing. doi: <https://doi.org/10.1016/B978-075067856-8/50000-1>
- Brezonik, P., & Arnold, W. (2011). *Water chemistry an introduction to the chemistry of natural and engineered aquatic systems*. Oxford University Press.
- Church, C. D., Hristov, A. N., Kleinman, P. J., Fishel, S. K., Reiner, M. R., & Bryant, R. B. (2018). Versatility of the MAnure PHosphorus EXtraction (MAPHEX) System in Removing Phosphorus, Odor, Microbes, and Alkalinity from Dairy Manures: A Four-Farm Case Study. *Applied Engineering in Agriculture*, *34*(3), 567–572. Retrieved 2019-09-13, from <http://elibrary.asabe.org/abstract.asp?AID=48976&t=3&dabs=Y&redir=&redirType=> doi: 10.13031/aea.12632
- Couper, J. R., Penney, W. R., Fair, J. R., & Walas, S. M. (2005). *Chemical process equipment:*

- selection and design*. Gulf Professional Publishing.
- DPC Water Solutions. (2021). *Debubblers* (Tech. Rep.). DPC Water Solutions.
- Fachagentur Nachwachsende Rohstoffe. (2010). *Biogas Guide. From production to use*. (Tech. Rep.).
- Fowler, B., & Kuroda, S. (1986). Changes in heated and in laser-irradiated human tooth enamel and their probable effects on solubility. *Calcif. Tissue. Int.*, *38*, 197-208.
- Geankoplis, C. J. (2003). *Transport processes and separation process principles:(includes unit operations)*. Prentice Hall Professional Technical Reference.
- Gregory, T., Moreno, E., & Brown, W. (1970). Solubility of $\text{CaHPO}_4 \cdot 2\text{H}_2\text{O}$ in the system $\text{Ca}(\text{OH})_2 - \text{H}_3\text{PO}_4 - \text{H}_2\text{O}$ at 5, 15, 25 and 37.5 °C. *J. Res. Natl. Bur. Stand.*, *74*, 461-475.
- Hsu, C.-H., & Li, M.-H. (1997). Densities of aqueous blended amines. *Journal of Chemical & Engineering Data*, *42*(3), 502-507.
- Kister, H., & Gill, D. (1991). Predict flood point and pressure drop for modern random packings. *Chemical Engineering Progress*, *87*(2), 32-42.
- Larsen, T., Udert, K., & Lienert, J. (2013). *Source separation and decentralization for wastewater management*. Iwa Publishing.
- Linstrom, P. J., & Mallard, W. G. (2001). The nist chemistry webbook: A chemical data resource on the internet. *Journal of Chemical & Engineering Data*, *46*(5), 1059-1063.
- Martín-Hernández, E., Ruiz-Mercado, G. J., & Martín, M. (2020). Model-driven spatial evaluation of nutrient recovery from livestock leachate for struvite production. *Journal of Environmental Management*, *271*, 110967.
- Matches. (2014). *Matches' process equipment cost estimates*. <https://matche.com/equipcost/Default.html>. ([Online; accessed 29-March-2018])
- Melse, R. W., & Ogink, N. (2005). Air scrubbing techniques for ammonia and odor reduction at livestock operations: Review of on-farm research in the netherlands. *Transactions of the ASAE*, *48*(6), 2303-2313.
- Metcalf, & Eddy. (2014). *Wastewater engineering: Treatment and resource recovery*. McGraw-Hill, New York.
- Møller, H., Lund, I., & Sommer, S. (2000). Solid-liquid separation of livestock slurry: efficiency and cost. *Bioresource Technology*, *74*, 223-229.

- Morse, J. W., Arvidson, R. S., & Lüttge, A. (2007). Calcium carbonate formation and dissolution. *Chemical Reviews*, 107, 342-381.
- Ohlinger, K. N., Young, T., & Schroeder, E. (1998). Predicting struvite formation in digestion. *Water Research*, 32(12), 3607-3614.
- Penn State Extension. (2016). *Agricultural Anaerobic Digesters: Design and Operation* (Tech. Rep.). Penn State University.
- Peters, M. S., Timmerhaus, K. D., West, R. E., et al. (2003). *Plant design and economics for chemical engineers* (Vol. 4). McGraw-Hill New York.
- PWTech. (2018). *Volute Dewatering Press* (Tech. Rep.). PWTech Process Wastewater Technologies LLC.
- Rongwong, W., & Sairiam, S. (2020). A modeling study on the effects of pH and partial wetting on the removal of ammonia nitrogen from wastewater by membrane contactors. *Journal of Environmental Chemical Engineering*, 8(5), 104240.
- Ryckebosch, E., Drouillon, M., & Vervaeren, H. (2011). Techniques for transformation of biogas to biomethane. *Biomass and Bioenergy*, 35(5), 1633-1645.
- Seader, J., Seider, W. D., & Lewin, D. R. (2004). *Product and process design principles: synthesis, analysis and evaluation*. Wiley.
- SG Projects. (2021). *Gas transfer membranes* (Tech. Rep.). SG Projects.
- Skoog, D., West, D., Holler, F., & Crouch, S. (2014). *Fundamentals of analytical chemistry*. Cengage Learning.
- Strigle, R. F. (1994). *Packed tower design and applications: random and structured packings*. Gulf Pub Co.
- Taylor, A., Frazier, A., & Gurney, E. (1963). Solubility products of magnesium ammonium and magnesium potassium phosphates. *Transactions of the Faraday Society*, 59, 1580-1584.
- Towler, G., & Sinnott, R. (2012). *Chemical engineering design: principles, practice and economics of plant and process design*. Elsevier.
- Turley, D., Hopwood, L., Burns, C., & Di Maio, D. (2016). *Assessment of digestate drying as an eligible heat use in the renewable heat incentive* (Tech. Rep.). NNFCC.
- Turton, R. (2010). *CAPCOST software to accompany: Analysis, synthesis, and design of chem-*

- ical processes* (Tech. Rep.). Upper Saddle River, N.J: Prentice Hall PTR.
- Ulbricht, M., Schneider, J., Stasiak, M., & Sengupta, A. (2013). Ammonia recovery from industrial wastewater by transmembranechemisorption. *Chemie Ingenieur Technik*, 85(8), 1259–1262.
- Verrecht, B., Maere, T., Nopens, I., Brepols, C., & Judd, S. (2010). The cost of a large-scale hollow fibre mbr. *Water Research*, 44(18), 5274–5283.
- Wilson, Z. T., & Sahinidis, N. V. (2017). The alamo approach to machine learning. *Computers & Chemical Engineering*, 106, 785–795.
- Yu, H., Li, X., Chang, H., Zhou, Z., Zhang, T., Yang, Y., ... Liang, H. (2020). Performance of hollow fiber ultrafiltration membrane in a full-scale drinking water treatment plant in china: a systematic evaluation during 7-year operation. *Journal of Membrane Science*, 613, 118469.
- Zhu, Z., Hao, Z., Shen, Z., & Chen, J. (2005). Modified modeling of the effect of ph and viscosity on the mass transfer in hydrophobic hollow fiber membrane contactors. *Journal of Membrane Science*, 250(1-2), 269–276.