

PRELIMINARY DESIGN OF OPTIMIZED HEAT INTEGRATED TWO-STAGE VACUUM EVAPORATION FOR PROCESSING DIGESTATE FROM BIOGAS PLANT

by

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Original scientific paper
<https://doi.org/10.2298/TSCI200401283N>

This work presents a preliminary design of a two-stage vacuum evaporation process converting a diluted liquid digestate into concentrated liquid fertilizers. Digestate is produced in a 1 MW biogas plant during the anaerobic digestion of poultry manure and corn silage. Laboratory experiments showed that in the first evaporation stage, about three-quarters of input digestate can be stripped to a diluted ammonia solution, while the concentrate can be used as phosphorus-potassium PK-fertilizer. After neutralization with H_2SO_4 , the ammonium sulphate solution is concentrated in the second evaporation stage. Feasible operating temperatures that allow heat integration between the two stages were determined in a laboratory environment at 40 °C for the first stage and 60 °C for the second. A preliminary process flow sheet was simulated in Aspen Plus to obtain data for heat integration and optimization of industrial-scale processes. The process was completely integrated by using the waste hot utility available at the site, while the external utilities demand was virtually zero. Optimizing the flow rate of the added sulphuric acid improved overall economic performance. The optimization and heat integration of the two-stage vacuum evaporation process within a biogas plant resulted in a circular and economically viable waste management technology.

Keywords: digestate, biogas, vacuum evaporation, ammonium sulfate, heat integration

Introduction

Waste recovery is a crucial activity for the transition to circular economy, as it reduces waste disposal and contributes to recycling of various precious materials [1]. Waste from livestock production, such as animal manure, can be efficiently used for biogas production by anaerobic digestion in which a digestate is a side product [2]. Liquid digestate, rich in inorganic nitrogen, phosphorus and potassium, can be used directly as fertilizer [3]. The recent European Regulation on Fertilizing Products [4] will enable the use and marketing of digestate as an organic fertilizer and soil improver after the implementation phase.

Recent studies have shown that all digestates cannot be classified directly as organic fertilizers [5], and further quality criteria need to be established in order to put them on the

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market. Moreover, owing to the high percentage of water in digestate, storage and the costs of transport to the fields are high, and the nutrient concentration lower than chemical fertilizers [6]. A variety of technologies is available to concentrate the nutrients in digestate and produce valuable products, including phosphorus precipitation by producing struvite, evaporation of ammonia from the aqueous phase followed by acidic neutralization producing ammonium salts, membrane processes, adsorption on zeolites, resins, and biochar, and the production of biomass [7]. The electrochemical recovery of phosphate as struvite, its electrochemical decomposition, and removal of the ammonia nitrogen by recycling the struvite product were tested to simultaneously recover phosphate and remove the ammonia nitrogen from swine wastewater [8]. Integrated membrane processes, including centrifugation, ultrafiltration, and reverse osmosis, can separate digestate into a solid N,P-fertiliser, a liquid N,K-fertiliser, and dischargeable water, but the high fouling potential of digestate prevents wider use of these processes [9].

One of the most promising technologies for nutrient recovery from digestate is ammonia stripping followed by absorption in sulphuric acid, which produces ammonium sulfate [10]. Alternatively, nitric acid can be used to produce ammonium nitrate [11]. However, nitric acid is more expensive, and the supply of ammonium nitrate to private buyers is restricted because of its explosive properties [12]. Since the stripping process requires high energy consumption, alternatives have been studied, such as solar exploitation for ammonia stripping in a transparent greenhouse [13], and vacuum thermal stripping and acid absorption [14]. In the vacuum process, ammonia is stripped from digestate at a temperature below boiling point [15]. This approach could be a feasible alternative in biogas plants where low temperature waste heat is usually available and can be used for vacuum evaporation, thus improving the environmental sustainability of these plants [16].

Different pilot-scale vacuum evaporation systems for treating digestate from biogas plants were tested: single-stage and two-stage, with and without acidification. A two-stage configuration with acidification that transforms digestate into a solid product has proved the most efficient [17]. Among the commercially available industrial evaporators, the most efficient are multi-stage flash evaporators and falling-film evaporators, while forced-circulation evaporators have the highest utility consumption [18]. Vondra *et al.* [19] performed a techno-economic analysis of vacuum evaporation in combination with ammonia scrubbers, stripping and reverse osmosis, while utilizing waste heat from combined heat and power units in a biogas plant. They showed that the feasibility of an evaporation system within a biogas plant depends on transport cost, price of electricity, investment costs and government incentives.

For thermal processes, it is essential to optimize utility consumption, which can be accomplished by using process systems engineering approaches, such as process integration, intensification and optimization [20]. Hernandez and Martin [21] applied a heat integration and optimization approach to waste processing into biodiesel, including nutrient recovery from digestate for algae growing. Ammonia stripping from liquid digestate can also be integrated with thermal processes for processing solid digestate, such as combustion, pyrolysis and gasification [22].

A literature review showed that vacuum evaporation could be a promising alternative for treating digestate and concentrating nutrients. However, the efficiency of this process needs to be maximized in order to use this technology in full scale systems. The aim of this study is to develop the preliminary design of an optimized heat-integrated plant for processing digestate from a 1 MW biogas plant, using two-stage vacuum evaporation technology. The goal was to produce high quality concentrated liquid fertilizers, one rich in phosphorus and

potassium, and another rich in nitrogen in the form of a 36% ammonium sulfate solution. The first could be used as a basal dressing for winter crops, usually applied in autumn, while the other could be used as a top-dressing fertilizer to provide nitrogen for growth [23]. The production of liquid products instead of solids was chosen because of the limited available capacity and because the goal was to apply fertilizers locally within a radius of 10 to 15 km around biogas plant.

The original contribution of this study is its use of process systems optimization to develop new models for: water recovery, through which up to 55% of water usage for biogas plants can be recycled, waste heat recovery for heating evaporation stages, and production of concentrated N-fertilizer and PK-fertilizer for sustainable food production. All three measures contribute to the transition to a circular economy. The study was performed by combining laboratory experiments and process systems engineering methods to obtain a feasible and economically viable preliminary design for industrial plants.

Materials and methods

Laboratory experiments were performed in order to: check the feasibility of two-stage vacuum evaporation for transforming a dilute digestate into concentrated fertilizers, obtain the norms for mass- and energy-balances of industrial-scale processes, and to experimentally determine feasible operating conditions for two-stage vacuum evaporation, including temperature, pressure and pH value. Process systems methods were used to: derive a preliminary flow sheet for an industrial plant with mass and energy balances and perform heat integration, economic evaluation and process optimization in order to achieve an economically viable installation of a vacuum evaporation process for treating waste digestate in a biogas plant.

Laboratory two-stage vacuum evaporation

A laboratory rotary evaporator, fig. 1, was used for testing the evaporation of digestate in two stages and determining the feasibility of certain operating parameters, the temperature, pressure and pH, for producing top quality fertilizers.

In the first stage, 200 mL of liquid digestate was added to a 500 mL round bottomed flask. The evaporation was carried out under vacuum at 40 °C and absolute pressure of 72 mbar. The pH value of the digestate was above 8, which enabled evaporation of aqueous ammonia into the gaseous phase. At higher pH values, the equilibrium between dissolved and gaseous ammonia is shifted towards the gaseous phase [17]. The gaseous ammonia was put in the condenser, followed by neutralization of first-stage condensate by sulphuric acid (H₂SO₄, 10%), which formed the ammonium sulfate with aqueous ammonia. A diluted solution of ammonium sulfate was then introduced into a second evaporation stage, which was performed using the same laboratory rotary evaporator.

This stage was carried out at various temperatures and corresponding pressures (40 °C and 70 mbar, 60 °C and 200 mbar) and pH values 2, 4, and 5. The operating parameters of



Figure 1. Laboratory rotary evaporator (Laborota 4000 Efficient, Heidolph)

both stages were selected with the aim of achieving an energy efficient evaporation process, as shown in section *Heat integration of two-stage vacuum evaporation process*.

Chemical analyses

The pH values were measured in samples of digestate, condensate and concentrate from the first and second stages of evaporation using a Multi 3410 pH meter (WTW) and electrical conductivity, κ , by LF 537 conductometer (WTW).

Sulfate ions (SO_4^{2-}) were determined in samples of condensate and concentrate after evaporation using an ion chromatograph. The concentration of ions was calculated according to the results obtained from standard measurement. The method was in accordance with ISO10304-1:2007. Dry matter content (TS) was determined in samples of digestate. The sample was placed for about 2 hours in a dryer at 105 °C until constant mass. PO_4^{3-} (ISO6878:2004) and NH_4^+ (ISO7150-1:1984) ions were determined spectrophotometrically.

Process simulation and heat integration

Simulation of an industrial-scale two-stage evaporation process was performed using Aspen Plus software. The input data for initial process simulation was taken from the results of the laboratory experiments. The main chemical components within the systems were:

- NH_3 , H_2O , H_2SO_4 , NH_4HSO_4 (salt), $(\text{NH}_4)_2\text{SO}_4$ (salt).
- ions: NH_4^+ , H_3O^+ , OH^- , HSO_4^- , SO_4^{2-} .
- non-volatile compounds were simulated with the component glycerol tristearate ($\text{C}_{57}\text{H}_{110}\text{O}_6$).

An ELECNRTL thermodynamic model was used for predicting the thermodynamic properties of the system because of the presence of electrolytes. The following process modules were used: FLASH2 for evaporation, HEAT for heating and cooling, and MIX for reaction of ammonia with sulphuric acid.

Heat integration options were studied using a Pinch analysis method [20] in which hot and cold process streams within a two-stage vacuum evaporation plant were defined by their supply and target temperatures, while the required heat flow rates were obtained from the results of process simulation. The utilities available at the biogas plant were also identified, together with their temperatures and capacities. The potential for heat integration between process streams and minimum on-site and external utility consumption was determined by constructing composite curves and a grand composite curve.

Economic evaluation and process optimization

Economic evaluation of the fertilizer production plant was performed by evaluating the main economic indicators, as defined by eqs. (1)-(5):

$$P_B = R_{\text{fertilizers}} + S_{\text{water}} - c_{\text{H}_2\text{SO}_4} - c_{\text{utilities}} - D \quad (1)$$

$$P_A = (1 - r_t)P_B \quad (2)$$

$$C_A = P_A + D \quad (3)$$

$$t_{\text{PB}} = \frac{I}{C_A} \quad (4)$$

$$NPV = -I + C_A \frac{(1 + r_d)^n - 1}{r_d(1 + r_d)^n} \quad (5)$$

where c is the cost, C_A – the cash flow, D – the depreciation, I – the capital investment, n – the life time, NPV – the net present value, P_A – the profit after tax, P_B – the profit before tax, R – the revenue, r_d – the discount rate, r_t – the tax rate, S – the saving, and t_{PB} – the payback time.

Table 1. The results of the liquid fraction digestate characterization

Digestate	TS [%]	γ (NH ₄ ⁺) [mg/L]	γ (PO ₄ ³⁻) [mg/L]	γ (K ⁺) [mg/L]
1	6.56	2662	5858	3027
2	6.25	2624	5080	3595
3	6.77	1085	7151	3552

The revenue in eq. (1) originates from the sale of concentrated fertilizers from the first and second evaporation stages, and from savings earned by substituting an amount of fresh water in the biogas plant by the condensate produced within the second evaporation stage. Expenses originate from the purchase of sulphuric acid

and electricity. Capital investment was determined using the Aspen Economic Analyzer, while process optimization was performed with the Aspen Plus Optimization Tool, by defining manipulating variables for maximizing the objective function, *i.e.* the annual profit.

Results and discussion

Laboratory experiments

This section presents the main properties of digestate and the laboratory execution of two evaporation stages, with the addition of sulphuric acid in between. Detailed results are presented elsewhere [24], while the most relevant results are summarized below.

Digestate analyses

Three different digestate samples were taken and analysed to determine their characteristics. Measurements of dry matter content (TS) and concentrations of ammonium, phosphate and potassium ions in liquid digestate are presented in tab. 1. In digestate sample No. 3, the ammonium content was very low, owing to fermenter cleaning; therefore, the steady state was not reached.

Calculations of nitrogen content in NH₄⁺, phosphorus in PO₄³⁻ and potassium in K⁺ showed that digestate's NPK ratio is approximately 1:1:1.6 which, together with its large water content, is not favorable for plant growth [25]. Therefore, dewatering and nutrient concentration were performed on a laboratory scale.

First evaporation stage

Evaporation of digestate was performed at 40 °C, yielding a concentrate rich in potassium and phosphorous compounds, and a condensate rich in ammonia. Based on the laboratory experiments, 76% of the digestate mass-flow rate was evaporated and recovered as a condensate, while the concentrated solution amounted to 24%. All percentages in this paper represent mass fractions. The condensate of the first stage was then neutralized with 10% sulphuric acid, which reacts with ammonia, into ammonium sulfate. The volume of added acid varied from 3 mL to 10 mL. The obtained solution contained about 0.4% ammonium sulfate and was fed into the second-stage evaporation.

Second evaporation stage

A diluted solution of ammonium sulfate from the first stage was concentrated in the second stage at 60 °C in order to reach an ammonium sulfate mass fraction of about 36%. For

the second stage, the split fraction of the remaining liquid phase was 8%, while the vapour phase reached 92%. The vapour phase of the second stage was purified water, which can be condensed and re-used as diluting water in biogas plants, thus reducing the consumption of fresh water significantly. The conductivity of water was stable at 27 $\mu\text{S}/\text{cm}$.

The conductivity and mass fraction of ammonium sulfate in the concentrate after the second evaporation stage at 60 °C and pH = 5 are presented in tab. 2. As the volume of concentrate decreased, the mass fraction of ammonium sulfate increased. It reached 35% after 60 minutes, when the experiment was stopped because further concentration was not desired. However, this result should be used with caution due to the very low volume available for analysis. For this reason, measurement of conductivity was not performed on this sample.

Table 2. The results of the concentrate after the second stage of evaporation

T [min]	κ [μScm^{-1}]	ω [(NH_4) $_2$ SO $_4$] [%]
0	6490	0.61
15	9680	0.75
30	13140	1.18
45	31300	2.53
60	–	35.62

Design, heat integration and optimization of industrial-scale process

A biogas plant of 1 MW_e capacity was assumed for the industrial case study. Biogas would be produced by anaerobic digestion of poultry manure and corn silage. Annual input capacity of the biogas plant would amount to 21000 tonnes per year of raw materials, of which about 70% consists of poultry manure and 30% corn silage. 25,000 m³ per year of fresh water would be consumed for dilution of input substrates. Production of biogas would amount to around 4 million m³ per year containing about 50% to 60% methane. It was assumed that 20000 tons of digestate would be produced annually.

Simulation of an industrial-scale two-stage vacuum evaporation process

First, mass balances of an industrial two-stage vacuum evaporation process for treating 20000 tonnes per year of digestate were derived using a simulation in Aspen Plus software. The simulation process flow sheet consisted of the main steps as shown in fig. 2.

Digestate (stream DIGEST, 20000 tonnes per year, 40 °C) is fed into the first evaporator stage simulated by the three blocks: heater (EVAPIH), flash (EVAPIF) and condenser (EVAPIC). The mass fraction of nonvolatile compounds in digestate is assumed to be 7.0% and ammonia 0.3%.

In the flash unit (EVAPIF), a vapour phase is separated from the liquid phase. The vapour phase, containing 0.4% ammonia (stream S2, 15200 tonnes per year), is condensed in a condenser (EVAPIC, stream S3). The liquid outlet stream is a concentrated solution (stream FERT-PK, 4800 tonnes per year) containing 29% nonvolatile compounds and 71% of water, while ammonia is present in trace amounts (< 0.02%). This solution, rich in phosphorus and potassium, could be used as a PK-fertilizer.

The condensed stream is mixed with technical sulphuric acid (146 tonnes per year, 80%) in a stirred tank reactor (M). The flow rate of sulphuric acid is determined in such a way that the pH of the final product is between 5 and 6. Note that for the industrial process, 80% acid was chosen. If 10% acid were used as in laboratory simulations, larger volumes and thus larger processing units would be required. Figure 3 presents the concentration of free ammonia

in the reactor effluent (stream S4) and the pH value of the solution, depending on the H_2SO_4 mass-flow rate. At the assumed acid flow rate (146 tonnes per year or 17 kg/h), the effluent's pH value was about 8.1.

Figure 3 indicates that there is still some room for optimization of the acid flow rate, because free ammonia can be further reduced, leading to a higher yield of ammonium sulfate. However, as shown in fig. 3, the flow rate of H_2SO_4 should not exceed approximately 24 kg/h; otherwise, the pH value of the final product would be too low. Information derived from fig. 3 would be used as limiting values for process optimization in section *Optimization of the heat integrated process flow sheet*.

The neutralized solution (stream S4) at 24 °C and a flow rate of 15346 tonnes per year was fed into the second evaporator comprising three blocks: heater (EVAP2H), flash (EVAP2F) and condenser (EVAP2C). The vapour outlet stream (stream S6) contains mostly water with traces of ammonia (0.12%). After cooling and condensation in EVAP2C, the water stream can be re-used in biogas digestion. The liquid outlet stream (stream S7) is a 36% aqueous solution of ammonium sulfate and stream (FERT-N) is the final product at a flow rate of 439 tonnes per year.

The results of the simulation showed that unwanted precipitation of solid crystals from the final liquid product occurs if the mass fraction of ammonium sulfate increases above 47%. This would occur at a very small temperature increase within the second stage, from 60 °C to 62 °C. As this might cause serious operating problems, efficient temperature control during the second evaporation stage is of great importance.

The results of the simulation, summarized in tab. 3, indicate that treating 20000 tonnes per year of digestate would produce 4800 tonnes per year of phosphorus- and potassium-rich fertilizer in the first evaporation stage and 439 tonnes per year of liquid nitrogen-rich fertilizer (ammonium sulfate, 36%) in the second stage. 14907 tonnes per year of water would be produced in the second stage evaporator, which could be reused in a biogas digestion process. In this way, the consumption of fresh water would decrease by almost 60%.

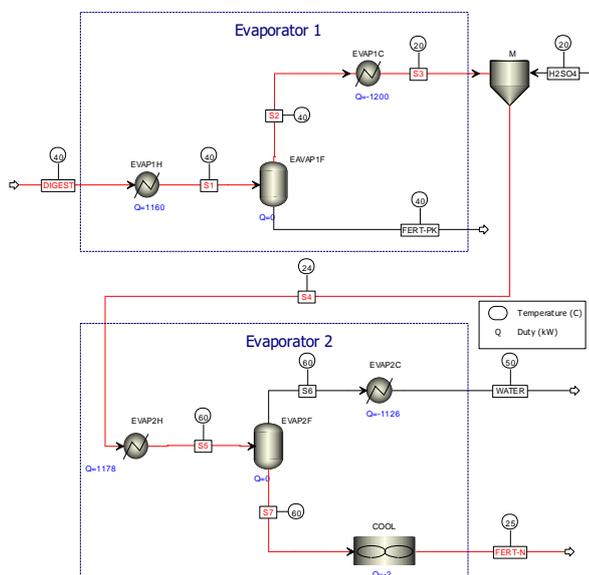


Figure 2. Process flow sheet of the two-stage vacuum evaporation in Aspen Plus

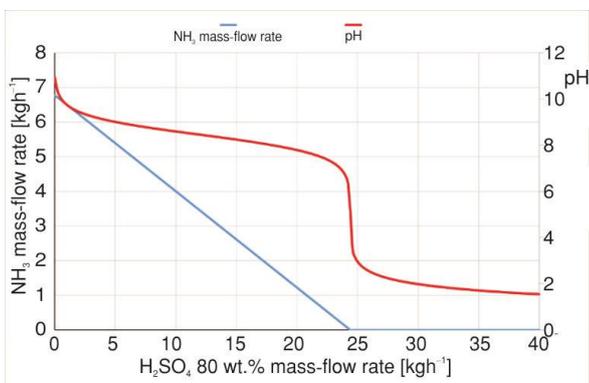


Figure 3. Process flow sheet of the two-stage vacuum evaporation in Aspen Plus

These results served as an input for process integration, economic evaluation and optimization.

Table 3. Results of simulation in Aspen Plus

	DIGEST	FERT-PK	S3	H2SO4	S4	S6	S7	FERT-N
Flow rate [t/y]	20000	4800	15200	146	15346	14907	439	439
Flow rate [kg/h]	2283	547.9	1735.1	16.7	1751.8	1701.7	50.1	50.1
Flow rate [t/y]	20000	4800	15200	146	15346	14907	439	439
Water mass fraction [%]	92.7	70.8	99.6	20.0	98.8	99.8	64.1	64.1
T [°C]	40	40	20	20	23.8	60	60	20
p [bar]	1	0.0738	0.0738	1	0.0738	0.1742	0.1742	0.1742
Phase	liquid	liquid	liquid	liquid	liquid	vapour	liquid	liquid

Heat integration of two-stage vacuum evaporation process

Process simulation provided heat flow rates for evaporation and condensation in the two-stage evaporation process. Table 4 summarizes the data for hot and cold streams. Note that a 1 K temperature difference was assumed for evaporation and condensation in order to determine heat capacity flow-rates, CF , as ratios between heat flow-rates and temperature difference. In the first stage, 1160 kW of hot utility was required for evaporation at 40 °C (EVAP1H), and 1200 kW of cold utility for condensation (EVAP1Ca) and cooling to 20 °C (EVAP1Cb) of the vapour stream containing 0.4% ammonia. Second stage evaporation required 1191 kW for preheating and evaporation (EVAP2Ha and EVAP2Hb), while condensation of the vapour water stream at 60 °C and cooling to 50 °C released cumulatively 1136 kW of heat (EVAP2Ca and EVAP2Cb). Waste hot water at 80 °C generated during the cooling of an engine within a cogeneration unit was available at the biogas plant as a free hot utility.

Without heat integration, cold streams would require 2351 kW of hot utilities, and hot streams 2336 kW of cold utilities. The goal was to determine the maximum extent of heat that can be exchanged between process streams and minimum utility consumption. The Composite Curves, fig. 4, were constructed for a minimum temperature difference of 10 K, yielding a pinch temperature at 30 °C to 40 °C. Minimum consumption of hot utility was 1200.7 kW and cold utility 1185.7 kW. The process streams can exchange 1150.3 kW. Most of it (1136 kW) would be exchanged between the second-stage condenser (at 60 °C) and first-stage evaporator (at 40 °C), while 14.3 kW could be transferred from the first-stage condenser (at 40 °C) to preheating the stream entering the second stage evaporator (from 23 °C to 30 °C). From a practical point of view, only the first match would be reasonable, since the load of the second match is too small for practical applications.

Table 4. Heat exchange stream data for the two-stage vacuum evaporation process

	T_s [°C]	T_t [°C]	ϕ [kW]	CF [kWK ⁻¹]
Cold streams				
EVAP1H	39	40	1160.0	1160.0
EVAP2Ha	23	59	73.6	2.04
EVAP2Hb	59	60	1117.4	1117.4
Hot streams				
EVAP1Ca	40	39	1161.5	1161.5
EVAP1Cb	39	20	38.5	2.02
EVAP2Ca	60	59	1118.1	1118.1
EVAP2Cb	59	50	17.9	2.00

The utilities were selected by constructing a Grand Composite Curve, fig. 5. Heating duty could be provided by hot water at 80 °C, which would cool down to 68 °C (these temperatures are shown shifted down by 10/2 = 5 K in fig. 5). The mass-flow rate of hot water was determined to be approximately 85 m³/h. Cooling duty of the first-stage condenser would be provided within a closed loop by a glycol-water mixture, which would release the absorbed heat in an air-cooled heat exchanger. Operating temperatures of the glycol/water solution were assumed to range from 10 °C at the inlet to 30 °C at the outlet (shown shifted up by 5 K in fig. 5). The estimated flow rate of the glycol-water mixture was 56 t/h.

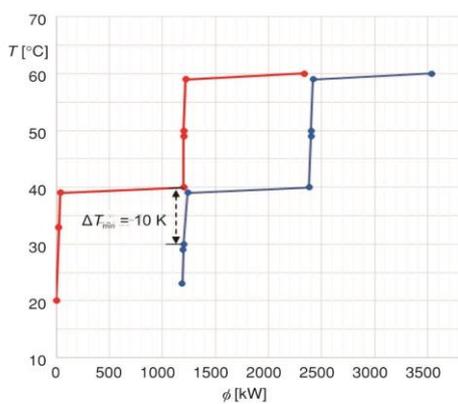


Figure 4. Composite curves of streams in the two-stage vacuum evaporation process

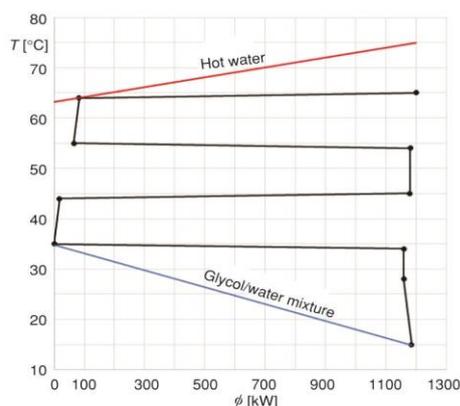


Figure 5. Grand composite curve with utilities for the two-stage vacuum evaporation process

Preliminary process flow sheet and economic assessment

A preliminary flow sheet of the heat integrated process is presented in fig. 6. It was assumed that liquid digestate entering the system would mix with the preheated digestate stream and then evaporate in the first evaporator. The vapour phase would be condensed in HE1 by a glycol-water mixture which circulates in a closed loop and releases the absorbed heat in an air-cooled chiller.

The diluted liquid ammonia stream is then mixed with technical H₂SO₄ (80%) in three neutralization basins, and the resulting ammonium sulfate solution enters the second evaporation stage. In this stage, evaporation takes place at 60 °C, heat is provided in HE2 by hot process water generated in the cogeneration engine. The liquid product is a concentrated nitrogen fertilizer, while vapours are condensed in HE3, releasing heat to the first stage evaporator and producing clean process water. The latter is recycled and re-used for diluting digestion input.

Economic evaluation of this flow sheet comprised approximate sizing of equipment, and evaluation of capital investment, revenue and expenses.

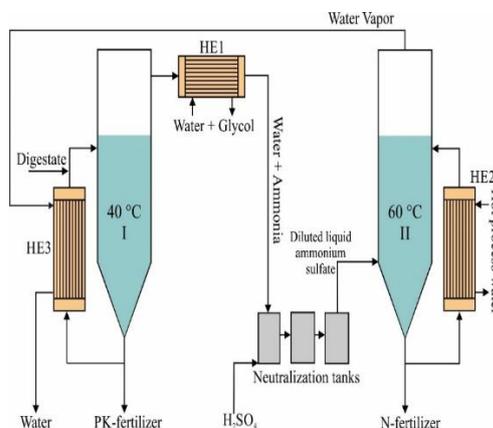


Figure 6. Flow sheet of the heat-integrated two-stage vacuum evaporation process

Investment. The capital investment was determined using the Aspen Process Economic Analyser, tab. 5. Installed capital cost amounted to 407000 €. This amount was increased by 30% in order to account for indirect costs, yielding a total capital investment of 529100 €.

Table 5. Capital cost estimation

	Size of process unit	Number of units	Installed cost per unit [€]	Total capital cost [€]
Evaporator	5 m ³	2	100000	200000
Heat transfer units	50 m ²	3	15000	45000
Pumps	25 kW	4	3000	12000
Neutralization basins	5 m ³	3	50000	150000
Sum of installed cost				407000
Indirect cost (30 %)				122100
Total capital investment				529100

Revenues and expenses. The main revenue was assumed to be from the sale of fertilizers, and the saving in water consumption was treated as a positive cash flow. The costs were composed of acid purchase and electricity costs for pumps, tab. 6. Straight-line depreciation over 10 years was assumed.

Table 6. Revenue, expenses and depreciation estimation

	Annual amount	Price per unit	Revenue/cost [€/y]
Revenue			
Fertilizer 1 st stage	4800 t/y	50 €/t	240000
Fertilizer 2 nd stage	439 t/y	180 €/t	79020
Water saving	14907 m ³ /y	1 €/m ³	14907
Expenses			
Sulphuric acid (80 %)	146 t/y	200 €/t	29200
Electricity for pumps	876000 kWh/y	0.07 €/kWh	61320
Depreciation			52910

Economic indicators. Based on the above results, the indicators of economic viability were assessed by assuming a 10-year life-time, 20% tax rate and 10% discount rate. Positive net present value indicated good economic potential of the proposed process (tab. 7, column entitled *Initial process*). Payback time under 3 years is in accordance with another study [19].

Table 7. Economic indicators of preliminary process flow sheet

	Initial process	Optimized process
Profit before tax [€/y]	190497	204533
Profit after tax [€/y]	152398	163627
Cash flow [€/y]	205308	216537
Payback time [y]	2.6	2.4
Net Present Value [€]	732426	801424
Internal Rate of Return [%]	37.2	39.5

Optimization of the heat integrated process flow sheet

In the final step, the heat integrated process flow sheet was optimized in Aspen Plus by maximizing profit before tax, as defined in eq. (1). It was established that the mass-flow rate of sulphuric acid had a large influence on the quality of the final fertilizer product, as well as on economic performance. This parameter was therefore defined as a manipulative variable varying between 15 and 24 kg/h as determined in fig. 3. The optimum flow rate was established at 20.4 kg/h or 178.8 tonnes per year (146 tonnes per year in a non-optimized process). Production of fertilizer increased

from 439 tonnes per year to 554 tonnes per year, while pH value was still within the allowed limits of 5 to 6. The amount of water for reuse was slightly lower. The economic indicators of the optimized process are presented in tab. 7, in the column *Optimized process*. The NPV is almost 10% higher than in the non-optimized flow sheet.

Conclusion

We present here the preliminary design of an industrial two-stage vacuum evaporation process for processing digestate from biogas plants into liquid fertilizers and clean water. We combined laboratory experiments to verify the feasibility of the procedure, a systems approach for design, heat integration and optimization of industrial-scale processes. Feasible operating conditions on a laboratory scale were determined at 40 °C and pH = 5 for the first evaporation stage, which produces a phosphorus and potassium rich concentrate, and 60 °C for the second stage, which produces an ammonium sulfate solution. Simulation of an industrial-scale plant processing 20000 tonnes per year digestate was accomplished with Aspen Plus. Heat integration showed that vapour from the second stage condensation could provide sufficient heat for the first stage evaporation, while the second stage could be heated by waste low-temperature process water available at the biogas plant. Cooling of the first condenser was accomplished using a water-glycol solution circulating in a closed loop. In this way, a two-stage evaporation unit would not require any external utilities, thus improving its economic performance. The main benefits of the evaporation process are the production of several valuable products from waste digestate: liquid fertilizer rich in phosphorus and potassium, a 36% solution of ammonium sulphate and clean water that can be re-used for diluting input substrates to the biogas digester. The optimized process flow sheet showed positive net present value, a high internal rate of return (39.5%) and a payback time of 2.4 years. Installation of two-stage vacuum evaporation within biogas production plants contributes to the transition to a circular economy; moreover, it could be feasible and economically viable.

Acknowledgment

The authors acknowledge financial support from the Slovenian Research Agency (research program P2-0032 and project J7-1816).

Nomenclature

C	– annual cost [€/y]	r_t	– tax rate [1]
C_A	– cash flow [€/y]	S	– saving [€/y]
CF	– heat capacity flow-rate [kW/K]	t_{PB}	– payback time [year]
D	– depreciation [€/y]	t	– time [min]
I	– capital investment [€]	TS	– total solids [mg/L]
m	– mass [g]	V	– volume [mL]
n	– life time [y]		
NPV	– net present value [€]		
P_A	– profit after tax [€/y]	<i>Greek symbols</i>	
P_B	– profit before tax [€/y]	γ	– mass concentration [mg/L]
R	– revenue [€/y]	κ	– electrical conductivity [mS/cm]
r_d	– discount rate [1]		

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