CHEMICAL ENGINEERING TRANSACTIONS Volume 25, 2011
Editors Jiří Jaromír Klemeš, Petar Sabev Varbanov, Hon Loong Lam
Copyright © 2011, AIDIC Servizi S.r.l., ISBN 978-88-95608-16-7 ISSN 1974-9791
DOI: 10.3303/CET1125152

Simultaneous Synthesis of a Biogas Process and Heat Exchanger Network for a Large-scale Meat Company

Rozalija Drobež[1*], Zorka Novak Pintarič[2], Bojan Pahor[3], Zdravko Kravanja[2]

[1]Scientific research centre Bistra Ptuj, Slovenski trg 6, Ptuj, SI-2250 Slovenia
[2]University of Maribor, Faculty of Chemistry and Chemical Engineering, Smetanova ulica 17, Maribor, SI-2000 Slovenia
[3] Perutnina Ptuj d.d, Potrčeva 10, Ptuj SI-2250, Slovenija
rozi.drobez@uni-mb.si

The aim of this contribution is to present an industrial application of mathematical programming for the simultaneous synthesis of a biogas plant and its heat exchanger network (HEN), within an existing meat company. The heat-integrated model recently developed by Drobež et al. (2010) has been extended with Yee & Grossmann simultaneous optimization MINLP model (1990) for the synthesis of heat exchanger networks. After some minor modifications to the model, it was possible to apply it to an industrial application for the simultaneous heat-integrated biogas process and HEN synthesis. The superstructure comprises biogas process alternatives under different anaerobic conditions, with the usage of different raw materials, including animal manure and other organic wastes, the option of a rendering plant for the utilization of slaughterhouse waste, different water supplies and wastewater treatment technologies, and also HEN with alternative process hot and cold streams, in order to consider different heat-exchange opportunities.

1. Introduction

Over the last few years, waste minimization and pollution prevention have become everyday terms throughout the process industry. Rising energy costs, coupled with stringent environmental regulations regarding the accumulation of animal manure and organic matter, have forced industrial sectors towards additional opportunities for the utilization of renewable resource and energy savings, by using process integration.
A diversity of papers in the literature deal with heat exchanger networks because of their impact on the energy recovery of industrial plants. Two major approaches have been developed for the optimal synthesis of heat exchanger networks: the pinch technology and mathematical programming methods. Mathematical optimization methods for HEN synthesis, based on a simultaneous approach, can explicitly handle the trade-offs between the capital and operational costs of the network. One of the most known simultaneous optimization models was developed by Yee & Grossmann (1990). It is based on stage-wise superstructure and formulated as an MINLP model, with the objective of simultaneously minimizing the utilities and capital costs of the network. With slight modification when considering processes stream temperatures and flows as optimization variables, it can be used for simultaneous process and HEN synthesis. However, the application of a superstructure approach, especially when considering the

Please cite this article as: Drobez R., Novak Z., Pahor B. and Kravanja Z., 2011, Simultaneous synthesis of a biogas process and heat exchanger network for a large-scale meat company, Chemical Engineering Transactions, 25, 911-916
DOI: 10.3303/CET1125152

supply of different animal manures for biogas production, has not yet been considered simultaneously during the synthesis of a heat exchanger network.
The work presented in this paper originated from the initiative of a meat company to efficiently transform agro-industrial by-products and animal wastes into biogas. A mathematical model for the simultaneous synthesis of the biogas process and heat exchanger network has been developed in order to accomplish this task optimally and efficiently. This model enables the selection of an optimal biogas production process, the optimization of biogas, electricity, heat production, and the selection of optimal HEN configuration with minimal consumption of utilities, and an optimal area and number of heat exchangers. As a basic formulation, the model recently developed by Drobež et al. (2010) was used, extended with Yee's model (Yee & Grossmann, 1990) modified for the simultaneous process and HEN synthesis.

2. Definition of the superstructure

In this section, Fig. 1 presents the superstructure relevant to this work. The superstructure for the simultaneous synthesis of biogas production presented by Drobež et. al., (2010) was upgraded with alternative process streams, mixers (M_1, M_2 and M_3) for inlet substrates and single-stage HEN superstructure. Note that when necessary process streams are partitioned into several segments rather than postulating a superstructure with many stages in order to decrease the number of alternative HEN units.

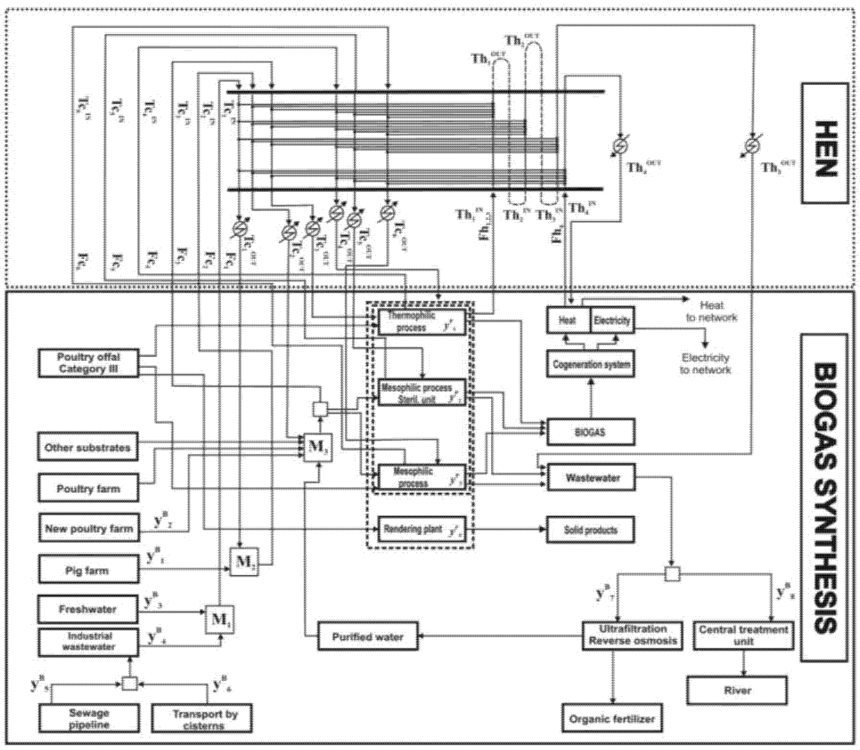

Figure 1: Superstructure for the simultaneous synthesis of biogas production and HEN.

3. Mathematical model

The mathematical model proposed in this contribution introduces an extension of our earlier work (Drobež et al., 2010) and therefore, only a brief description of the applied extensions for the simultaneous synthesis of the biogas process and heat exchanger network is presented in the following subsection: a definition of mixer mass and heat balances, definition of heat capacity flow-rates, and a set of linking equations between the process and its HEN. Note that the following assumptions were made on the modified model: a) the heat capacity flow-rates, and the supply and target temperatures of the streams are optimization variables, b) the hot and cold streams have constant heat capacities ($c_p^{\mathrm{IS}} = 4{,}186\ \mathrm{kJ/kgK}$ of liquids and $c_p^{\mathrm{BG}} = 2.293\ \mathrm{kJ/kgK}$ of gases), and c) only one hot and one cold utility is available.

3.1 Process synthesis model

Overall mass balances for mixer units ($q_{\mathrm{m}_j}^{\mathrm{M}}$ /(kg/d)) is given by:

$$q_{\mathrm{m}_j}^{\mathrm{M}_1} = \sum_{i \in I_4} q_{\mathrm{m}_{i,j}} + \sum_{i \in I_5} q_{\mathrm{m}_{i,j}} \qquad \forall j \in J_1 \tag{1}$$

$$q_{\mathrm{m}_j}^{\mathrm{M}_2} = \sum_{i \in I_2} q_{\mathrm{m}_{i,j}} + q_{\mathrm{m}_j}^{\mathrm{M}_1} \qquad \forall j \in J_1 \tag{2}$$

$$q_{\mathrm{m}_j}^{\mathrm{M}_3} = \sum_{i \in I_3} q_{\mathrm{m}_{i,j}} + \sum_{i \in I_7} q_{\mathrm{m}_{i,j}} + \sum_{l \in I_8} q_{\mathrm{m}_{j,l}}^{\mathrm{RWW}} \qquad \forall j \in J_1 \tag{3}$$

where $q_{\mathrm{m}_{i,j}}$ /(kg/d) denotes the mass flow-rate of substrate *i* in process *j* and $q_{\mathrm{m}_{j,l}}^{\mathrm{RWW}}$ /(kg/d) recirculated wastewater.

The heat capacity flow-rates ($Fc_{i,j}^{\mathrm{S}}$ /(kW/K)) of water supplies (freshwater or industrial wastewater) and inlet substrates from the pig, poultry, and new poultry farms, are similarly defined as follows:

$$\sum_{i \in I} Fc_{i,j}^{\mathrm{S}} = \sum_{i \in I} q_{\mathrm{m}_{i,j}} \cdot c_p^{\mathrm{IS}} \cdot f_{\mathrm{d}} \qquad \forall j \in J_1 \tag{4}$$

where c_p^{IS} /(kJ/(kg·K)) is the specific heat capacity during the liquid phase, and f_{d} /(d/s) the time conversion factor.

The heat capacity flow-rates of the recirculated wastewater ($Fc_{j,l}^{\mathrm{RWW}}$ /(kW/K)) is given by:

$$\sum_{l \in L_8} Fc_{j,l}^{\mathrm{RWW}} = \sum_{l \in L_8} q_{\mathrm{m}_{j,l}}^{\mathrm{RWW}} \cdot c_p^{\mathrm{IS}} \cdot f_{\mathrm{d}} \qquad \forall j \in J_1 \tag{5}$$

The heat capacity flow-rate relating to the heating requirements of anaerobic digestion:

$$Fc_j^{\mathrm{AD}} = \sum_{i \in I} q_{\mathrm{m}_{i,j}} \cdot (\Phi_{\mathrm{loss},j}^0 / q_{\mathrm{m}_{i,j}}^0) \qquad \forall j \in J_1 \tag{6}$$

where $\Phi_{\mathrm{loss},j}^0$ /(kW/K) is the energy loss during anaerobic fermentation and $q_{\mathrm{m}_{i,j}}^0$ /(kg/d) is the base-case daily mass flow-rate of the substrates.

Heat capacity flow-rates of the mixer outlet streams:

$$Fc^{\mathrm{M}} = \sum_{j \in J_1} q_{\mathrm{m}_j}^{\mathrm{M}} \cdot c_p^{\mathrm{IS}} \cdot f_{\mathrm{d}} \tag{7}$$

Overall heat balance of the mixer units is calculated as:

$$Fc^{\mathrm{M}_1} \cdot T^{\mathrm{M}_1,\mathrm{OUT}} = \sum_{i \in I_4} \sum_{j \in J_1} Fc_{i,j}^{\mathrm{S}} \cdot T_{i,j}^{\mathrm{IN}} + \sum_{i \in I_5} \sum_{j \in J_1} Fc_{i,j}^{\mathrm{S}} \cdot T_{i,j}^{\mathrm{IN}} \tag{8}$$

$$Fc^{\mathrm{M}_2} \cdot T^{\mathrm{M}_2,\mathrm{OUT}} = \sum_{i \in I_2} \sum_{j \in J_1} Fc_{i,j}^{\mathrm{S}} \cdot T_{i,j}^{\mathrm{IN}} + Fc^{\mathrm{M}_1} \cdot T^{\mathrm{M}_2,\mathrm{IN}} \tag{9}$$

$$Fc^{\mathrm{M}_3} \cdot T^{\mathrm{M}_3,\mathrm{OUT}} = \sum_{i \in I_3} \sum_{j \in J_1} Fc_{i,j}^{\mathrm{S}} \cdot T_{i,j}^{\mathrm{IN}} + \sum_{i \in I_7} \sum_{j \in J_1} Fc_{i,j}^{\mathrm{S}} \cdot T_{i,j}^{\mathrm{IN}} + \sum_{j \in J_1} \sum_{l \in L_8} Fc_{j,l}^{\mathrm{RWW}} \cdot T_{j,l}^{\mathrm{IN}} + Fc^{\mathrm{M}_2} \cdot T \tag{10}$$

where $T^{\mathrm{M}_2,\mathrm{OUT}}$, $T^{\mathrm{M}_3,\mathrm{IN}}$/(K) are the outlet and inlet temperatures for the mixer units, and $T_{i,j}^{\mathrm{IN}}$/(K) the inlet temperature of the inlet substrates.

The heat capacity flow-rate of the wastewater (Fh^{WW}/(kW/K)) is defined as:

$$Fh^{\mathrm{WW}} = q_{\mathrm{m}_{j,l}}^{\mathrm{WW}} \cdot c_p \cdot f_{\mathrm{d}} \qquad j = 1,\ l \in L_7 \tag{11}$$

where is $q_{\mathrm{m}_{j,l}}^{\mathrm{WW}}$/(kg/d) mass flow-rate of wastewater.

Heat produced by a cogeneration system (Φ^{CHP}/kW) is calculated as:

$$\Phi^{\mathrm{CHP}} = \sum_{j \in J_1} q_{\mathrm{v}_j}^{\mathrm{BG}} \cdot \eta^{\mathrm{T}} \cdot e_j^{\mathrm{BG}} \cdot f_{\mathrm{d1}} \qquad \forall j \in J_1 \tag{12}$$

$$\Phi^{\mathrm{CHP}} = \Phi^{\mathrm{USE}} + \Phi^{\mathrm{SOLD}} \tag{13}$$

where $q_{\mathrm{v}_j}^{\mathrm{BG}}$ /(m^3/d) is the volume flow-rate during biogas production, η^{T} is the efficiency of heat generation, e_j^{BG}/(kWh/m^3) the heating value of the biogas, f_{d1}/(d/h) the time conversion factor, and Φ^{USE}, Φ^{SOLD}/(kW) represent the heat which is used in the plant and that which is sold, respectively.

The heat capacity flow-rate of the cogeneration system (Fh^{CHP}/(kW/K)) is defined as:

$$Fh^{\mathrm{CHP}} = \Phi^{\mathrm{USE}} / \left(T_{\mathrm{IN}}^{\mathrm{CHP}} - T_{\mathrm{OUT}}^{\mathrm{CHP}}\right) \qquad \forall j \in J_1 \tag{14}$$

where $T_{\mathrm{IN}}^{\mathrm{CHP}}$ and $T_{\mathrm{OUT}}^{\mathrm{CHP}}$/(K) are the inlet and outlet temperatures of the hot water, respectively.

3.2 Linking the process synthesis and HEN models

Both models are linked through the appropriate mapping of heat capacity flow-rates, and the supply and target temperatures of the streams as defined in the process model, with those in the HEN.

Hot streams:

$$\left(Fh^{\mathrm{WW}},\ Fh^{\mathrm{CHP}}\right) = \left(Fh_h, h \in H\right) \tag{15}$$

$$\left(T_j^{\mathrm{WW,IN}}, j{=}1\right) = \left(Th_1^{\mathrm{IN}}\right),\ \left(Th_{h-1}^{\mathrm{OUT}}\right) = \left(Th_h^{\mathrm{IN}}, h = 2,3\right) \text{ and } \left(T^{\mathrm{CHP,IN}}\right) = \left(Th_4^{\mathrm{IN}}\right) \tag{19}$$

$$\left(Th_2^{\mathrm{IN}}\right) \geq \left(Th_2^{\mathrm{OUT}}\right),\ \left(Th_j^{\mathrm{WW,OUT}}, j = 1\right) = \left(Th_3^{\mathrm{OUT}}\right) \text{ and } \left(T^{\mathrm{CHP,OUT}}\right) = \left(Th_4^{\mathrm{OUT}}\right) \tag{20}$$

Cold streams:

$$\left(Fc^{\mathrm{M}},\ Fc_j^{\mathrm{AF}}\ j \in J\right) = \left(Fc_c, c \in C\right) \tag{16}$$

$$\left(T^{\mathrm{M_m,OUT}}, \mathrm{m}{=}1,2,3\right) = \left(Tc_c^{\mathrm{IN}}, c = 1,2,3\right) \text{ and } \left(\sum_{j \in J_1} T_j^{\mathrm{AP,IN}} \cdot y_j^{\mathrm{P}}\right) = \left(Tc_c^{\mathrm{IN}}, c = 4,5,6\right) \tag{17}$$

$$\left(T^{\mathrm{M_{m+1},IN}}, \mathrm{m}{=}1,2\right) = \left(Tc_c^{\mathrm{OUT}}, c = 1,2\right),\ \left(\sum_{j \in J_1} T_j^{\mathrm{AD,IN}} \cdot y_j^{\mathrm{P}}\right) = \left(Tc_c^{\mathrm{OUT}}, c = 3\right) \text{ and}$$

$$\left(\sum_{j \in J_1} T_j^{\mathrm{AP,OUT}} \cdot y_j^{\mathrm{P}}\right) = \left(Tc_c^{\mathrm{OUT}}, c = 4,5,6\right) \tag{18}$$

3.3 Objective function

In order to consider an economic trade-off between the product income, cost for raw materials, operating cost, and the process and HEN investment, the objective function was formulated as a maximization of the net present worth (NPW) by:

$$F_{\mathrm{obj}} = -I + \left[\left(1 - r_{\mathrm{t}}\right) \cdot \left(R - E\right) + r_{\mathrm{t}} \cdot D \right] \cdot \left[\frac{(1 + r_{\mathrm{d}})^{t_{\mathrm{D}}} - 1}{r_{\mathrm{d}}(1 + r_{\mathrm{d}})^{t_{\mathrm{D}}}} \right] \qquad (21)$$

where I /(€) represents the investment of processes and HEN, r_{t} the tax rate, R /(€/y) the revenues or incomes, E /(€/y) the expenditures for raw materials and operating costs including the costs for hot and cold utilities, for the sterilization of category III slaughterhouse waste, D /(€/y) depreciation, r_{d} the discount rate, and t_{D} /(a) the depreciation period. For further details of the objective function, please refer to Drobež et al. (2010).

4. Industrial case study and results

The mathematical model for simultaneous synthesis of the biogas process and HEN with 4 hot and 6 cold stream alternatives was applied to an existing meat company. Firstly the hot stream was split into three segments (H_1, H_2 and H_3), where $\left(Th_{h-1}^{\mathrm{OUT}}\right) = \left(Th_h^{\mathrm{IN}}, h = 2,3\right)$. The industrial case-study was implemented in GAMS (Brooke et al., 2005). The MINLP problem was solved using DICOPT (Viswanathan & Grossmann, 1990) with CPLEX as a MIP solver, and BARON (Sahinidis & Tawarmalani, 2005) as a NLP solver on a PC machine (2.53G Hz, 2GB RAM). The optimal process scheme of the problem is shown in Fig.2 and the results are given in Table 1. By-products and animal manures from the meat company were converted into biogas under thermophilic conditions, and subsequently, the biogas was converted into electricity and heat in the CHP unit. The solution utilized freshwater and a closed water network using technologies for wastewater re-usage, and the production of organic fertilizer. The net present worth (NPW) is 10.362 M€.

Table 1: Results of HEN for hot and cold streams.

Streams ($h \in H$)	Fh_h (kWK^{-1})	Φ_{C_k} (kW)	$\Phi_{H,C,M}$ (kW)	$Th_{H,M-1}$ (°C)	$Th_{H,M-2}$ (°C)	Th_H^{IN} (°C)	Th_H^{OUT} (°C)
1	20.255	0.00	405.10	55	35	55	35
2	20.255	0.00	0.00	35	35	35	35
3	20.255	0.00	0.00	35	35	35	35
4	3.920	0.00	98.00	85	60	85	60
Streams ($c \in C$)	Fc_c (kWK^{-1})	Φ_{h_k} (kW)	$\Phi_{H,C,M}$ (kW)	$Tc_{H,M-1}$ (°C)	$Tc_{H,M-2}$ (°C)	Tc_C^{IN} (°C)	Tc_C^{OUT} (°C)
1	0.657	0.00	0.00	10	10	10	10
2	0.657	0.00	0.00	10	10	10	10
3	21.378	100.60	405.10	50	31	31	55
4	98.010	0.00	0.00	55	54	54	55
5	0.000	0.00	0.00	35	34	34	35
6	0.000	0.00	0.00	35	34	34	35

* $c^{\mathrm{CU}} = 0.005$ €/kWh, $c^{\mathrm{HU}} = 0.05$ €/kWh, $c^{\mathrm{LP}} = 0.05$ €/kWh, $h^{\mathrm{HU}} = 5$ kW/m^{2}K, $h^{\mathrm{CU}}, h_{H,C} = 1$ kW/m^{2}K, $\beta = 1$, $c_f = 46$ k€, $c_v^{\mathrm{HE,CU,HU}} = 2.742$ k€/m^2, $\Delta_{\mathrm{HRAT}}T = 1$ K, $T^{\mathrm{HU,IN}} = 157$ °C, $T^{\mathrm{CU,IN}} = 15$ °C

The heat exchanger network consisted of 3 heat exchangers (one heater, and two heat exchangers). The corresponding total area is 209.207 m^2, with heat of 100.6 kW. The capital cost for HEN and annual operating costs would be 0.758 M€ and 0.043 M€/y, respectively. The capital cost for the process is 16.684 M€ and total annual operating cost is 0.772 M€/y.

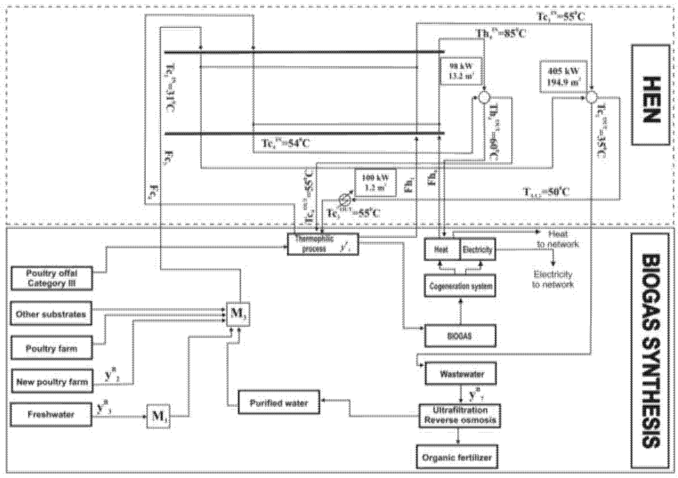

Figure 2: Optimal process scheme for synthesis of biogas production and HEN.

5. Conclusion

The large scale meat industrial application of an MINLP mathematical model is presented, where the synthesis of biogas process is performed simultaneously with the synthesis of heat-integrated HEN. This model developed by Yee & Grossmann (1990) was slightly modified to enable simultaneous process and HEN synthesis for specific industrial circumstances. The proposed optimization/synthesis model in this study can be used to test different options in order to support decision making for future biogas production investment.

References

Drobež R., Novak Pintarič Z., Pahor B., and Kravanja Z., 2010, Simultaneous Heat Integration and the Synthesis of Biogas Processes from Animal Waste, Asia-Pac. J. Chem. Eng. doi: 10.1002/apj.504.

Yee T.F., and Grossmann I.E., 1990, Simultaneous Optimization Model for Heat Integration – II: Heat Exchanger Network Synthesis, Comp. Chem. Eng. 14, 1165-1184.

Yee, T.F., Grossmann, I.E., and Kravanja, Z., 1990, Simultaneous Optimization Model for Heat Integration – III: Process and Heat Exchanger Network Optimization, Comput. Chem. Eng. 14 1185-1200.

Brooke A., Kendrick D., Meeraus A., and Raman R., 2005, GAMS: A Users Guide. Washington D.C., USA: GAMS Development Corporation.

Viswanathan J., and Grossmann I.E., 1990, A Combined Penalty Function and Outer Approximation Method for MINLP Optimization, Comp. Chem. Eng. 14, 769-782.

Sahinidis N.V., and Tawarmalani M., 2005, Accelerating Branch-and-Bound Through a Modeling Language Construct for Relaxation-Specific Constraints, J.Global. Optim. 32, 259-280.