

Modelling and real-time optimisation of a heat-exchanger network [★]

María P. Marcos ^{*,**} José Luis Pitarch ^{*} César de Prada ^{*,**}

^{*} *Systems Engineering and Automatic Control Dept., EII, Universidad de Valladolid. C/Real de Burgos s/n, 47011, Valladolid, Spain.
E-mails: {maria.marcos | jose.pitarch | prada}@autom.uva.es*

^{**} *Institute of Sustainable Processes, C/Real de Burgos s/n, 47011 Valladolid, Spain*

Abstract: This work proposes a hybrid mathematical model and an optimisation-based tool to support the management of a heat-recovery section (formed by several heat exchangers) in a fibre-production factory. The purpose of the network is to heat different products using several hot sources, employed as utilities. Furthermore, concerns about the degradation of the equipment due to fouling are explicitly taken into account. Hence, the goals are to allocate the hot sources to heat exchangers and to suggest which heat exchanger should be cleaned to achieve optimal economic operation. Experimental models for the overall heat-transfer coefficients with respect to the flows have been identified, and production constraints are considered too. The problem is formulated such that it can be solved in a real-time optimisation fashion via mixed-integer non-linear programming. The approach has been tested through plant historical situations.

Keywords: RTO, pulp industry, heat exchangers, fouling, MINLP, decision support

1. INTRODUCTION

Nowadays, the need to play in the global market and the increasingly restrictive environmental regulation force industrial companies to produce as efficient as possible, meaning to reduce the energy and resource consumption in daily operation. One way is coordinating the actions taken at different layers: process control, real-time optimisation (RTO), production/maintenance scheduling and economic planning (Khor and Varvarezos, 2017). In particular, RTO continuously searches for the best operating conditions, linking thus the process control with the scheduling and planning layers of the factory (de Prada and Pitarch, 2017). In the literature one can find many examples of the benefits of using an RTO scheme in process-decision making (Galan et al., 2019; Han et al., 2015; Marcos et al., 2018), integrating it in the information-technology (IT) infrastructure of the plant (Enste, 2019).

In practice, however, the deployment of RTO-based tools is often challenging due to the specific technical issues that arise, e.g. unreliable plant measurements, limited prediction range with data-based models, complex mathematical formulations, demanding computational requirements or difficult integration with the existing control or planning infrastructure. In this paper we face these issues in the development of a decision-support system (DSS) that aims at improving the operation of the heat-recovery section of a viscose fibre production plant.

^{*} This research is funded by the EU's Horizon 2020 research and innovation programme under grant agreement No. 723575, the Spanish MICIU with FEDER funds (PGC2018-099312-B-C31), the Regional Government of Castilla y León and EU-FEDER (UIC 233 and CLU 2017-09). The first author thanks the European Social Fund and the "Consejería de Educación de la Junta de Castilla y León".

A key stage in the proposed methodology is to apply data reconciliation to the raw measurements collected from the plant, in order to obtain a set of data which is consistent with the known basic process physics. In this way, wrong predictions and decisions due to erroneous or unreliable plant data are avoided. Then, prediction models (that will be the basis of the RTO) are build upon such coherent data. In this regard, first-principles models are advisable as they provide a wide range of validity, but some drawbacks like the presence of unknown time-varying parameters or subsystems that are difficult to model limit their usage. In these cases, the combination with a data-based approach is sensible, resulting in *grey-box models* (Pitarch et al., 2019). Here we followed this last approach to model the heat-transfer coefficients in the exchangers with respect to flows and fouling. In addition, when formulating the process model, one likely faces the presence of decisions of discrete nature or logical statements, like the selection of heat sources or the number of process units in operation, which leads to *hybrid models* containing continuous and binary variables. Indeed, the formulation of an RTO has to take into account realistic process operation conditions, in particular the equipment degradation and the need of performing maintenance tasks on the process units. Therefore, both model and optimisation criterion must include these aspects. Finally, the optimisation problem should follow a formulation such that solutions can be gathered in acceptable times with current software and hardware in the plant. In our case, the implementation was done in Pyomo (Hart et al., 2017) and, despite the formulated problem is mixed-integer nonlinear, good nearly-optimal solutions are obtained quick enough.

The remaining sections show, in this order, the description of the heat-exchanger network (Section 2), the data-based modelling for the overall heat-transfer coefficient (Section 3), the mathematical model describing the network operation and the formulation of the optimisation problem (Section 4) and the results for some of the tests done with plant historical data (Section 5). Finally, the paper ends with some conclusions and a few lines on the ongoing work.

2. DESCRIPTION OF THE NETWORK

The case study is provided by the company Lenzing AG (Austria), the EU's leading factory of man-made viscose fibre production. As in other process industries, some streams have to be heated up to a desired set point. A network of heat exchangers is used for that purpose, recovering the residual heat in waste streams to improve the plant resource efficiency. This network is formed by fifteen heat exchangers that have to heat twelve products using three sources of heat (see Fig. 1).

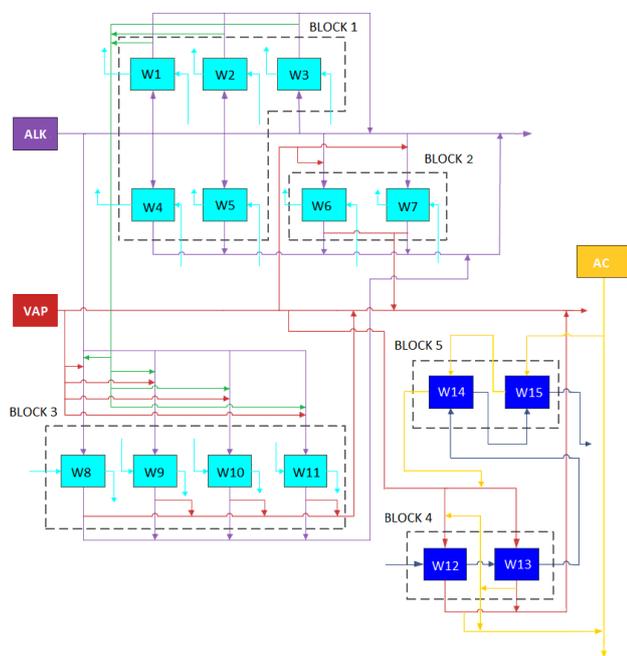


Fig. 1. Heat recovery network at Lenzing AG

However, from the heat exchangers point of view, the available sources of heat are indeed four:

- *alk* which is depicted in purple in Fig. 1
- *ualk* that is the amount of *alk* that has been used previously in some heat exchangers, in green in Fig. 1
- *vap* which is depicted in red in Fig. 1
- *ac* which is depicted in yellow in Fig. 1

In this way, each of these defined hot sources is characterised by its availability and temperature. It is noteworthy that not all the hot sources can pass through all heat exchangers. Depending on the possible physical connections, we can divide the heat exchangers in different groups as shown in Table 1. Note importantly that, although the hot sources are waste water, their use involves a shadow cost due to the fact that they could be used in other parts of the process too.

Table 1. Allowed source connection.

Block	Heat Exchanger	Hot sources
$\mathcal{B}1$	W-1, W-2 W-3 W-4, W-5	<i>alk</i>
$\mathcal{B}2$	W-6, W-7	<i>alk, vap</i>
$\mathcal{B}3$	W-8, W-9 W-10, W-11	<i>alk, vap, ualk</i>
$\mathcal{B}4$	W-12, W-13	<i>vap, ac</i>
$\mathcal{B}5$	W-14, W-15	<i>ac</i>

In the other side there are twelve products to heat up to their corresponding set points. Each heat exchanger depicted in cyan (W1-W11) in Fig 1 always processes the same product. All exchangers in dark blue heat a single product, which flows in serial connection from W12 to W15. It is on the output of W15 where the product temperature set point has to be fulfilled. Hence, the network can also be divided in two subnets: the first one formed by heat exchangers W1-W11, where the hot sources feed the heat exchangers in parallel and each one processes a different product; and other subnet involving W12 to W15 where the same product flows in series through them.

Note that the *ac* flow goes through W12 to W15 in series too, but in reverse way. Therefore, it may be possible to reach the product temperature set point at W15 without feeding the four heat exchangers. In such case, the flow of *ac* can bypass the heat exchangers that are not needed.

Apart from the constructive materials a dimensionality factors, the efficiency of the heat exchangers depends also on their operation (Boccardi et al., 2010; Pitarch et al., 2019). The fouling (accumulation of unwanted deposits on the heat-transfer surfaces) is indeed a major issue in many industrial equipment, as it increases the resistance to the heat transfer (Bott, 1995). The fouling state depends also on the operation conditions but it generally increases over time. Consequently, the exchangers must be cleaned from time to time to recover their nominal efficiency. The usual cleaning policy in the factories is exclusively based on operators experience, prioritising first the heat exchangers being more days in operation since last cleaning. In this way it is hard to infer whether the taken decisions really lead to an optimal operation regarding economics.

The goal of this work providing operators and plant managers with an RTO tool to support them in such complex decision-making process: to distribute hot sources among exchangers, suggesting which ones are the more beneficial to clean, for an economically optimal operation.

3. MODELLING THE HEAT-TRANSFER

The network efficiency depends directly on the heat transfer from hot to cold streams. This heat can be computed for each heat exchanger by:

$$Q' := U \cdot A \cdot LMTD \quad (1)$$

Where A is the heat-transfer area, LMTD is the logarithmic mean temperature difference between inlet and outlet streams, and U is the overall heat-transfer coefficient. The transfer areas A are known, and the LMDT can be computed from temperature measurements at the exchanger boundaries. However, the coefficient U is not constant because it depends on the streams densities, viscosities,

flow velocities, etc., i.e., it depends on the operating conditions. Moreover, the fouling in the exchanger surfaces also modifies the conduction coefficient. Since the aim of the proposed RTO is to compute the right flows through each heat exchanger for economic operation, a prediction model for U with respect to these variables is required.

Well established empirical laws based on dimensionless groups could be recalled for such a task. However, this approach requires a precise knowledge on the heat-exchanger dimensions and constructive materials, as well as sensible data on the streams properties (dynamic viscosity, density, etc.), in order to find the more suitable correlation among the Nusselt number with the Prandtl, Reynolds or Peclet ones, plus with other of kinematic and geometric nature (Dorao and Fernandino, 2017; Boccardi et al., 2010). The drawbacks with this approach are twofold:

1. Formulas for the convection coefficient are well known in the case of usual fluids, like water, steam or refrigerants (R-x), but no significant studies are available for other more scarce fluids, mixture of several components at varying concentrations, like the acid spinbaths in our case study. Hence, laboratory experiments are required to collect the relevant data for regression.
2. In “dirty” industrial environments, streams and heat-exchanger features vary with the time and operation conditions (concentrations, fouling, etc.), so the regressions got from the lab data may be quickly outdated if such data is not representative enough for the whole operating region, which barely can be.

In addition, these highly non-convex models rely on empirical data in the end, but the only data we can get online from the plant are flows and temperatures. Therefore, decided to avoid this way and we chose a more flexible data-driven approach. The first step is to build a data-based model for the *clean exchanger*. For such a task, data is gathered from the plant historian in different operating conditions (flow ranges) but always after a cleaning task (Lenzing, 2019). Next, a data reconciliation is ran to correct the raw data according to basic energy balances (Pitarch et al., 2019). If we assume no heat loss to the ambient, the heat lost by the hot water (Q_h) is the one gained by the cold product (Q_c), equation (2), where each heat flow can be computed by (3)¹:

$$Q_h + Q_c = 0 \quad (2)$$

$$Q_k := F_k \rho_k C p_k (T_{out_k} - T_{in_k}) \quad \forall k \in \{h, c\} \quad (3)$$

Here, F is the flow that goes through the heat exchanger, ρ its density, Cp the specific heat, being T_{out} and T_{in} the outlet and inlet stream temperatures respectively.

For example, Fig. 2 compares the reconciled values for the hot flow in an exchanger with the raw data. It can be seen that the corrected values are consistently lower than the measurements. This phenomenon was explained afterwards by the presence of a bypass valve located between the product flowmeter and the heat exchanger.

Once we got reliable data, the next step is to estimate U for each data point matching (1) with (3). Hence, we have *virtual* measurements of U in different operation conditions (F, T). The last step is fitting such data to a polynomial

¹ Densities and specific heats are considered constant values, measured in usual operating conditions.



Fig. 2. Example of results got by data reconciliation.

candidate form, for instance by SOS constrained regression (Pitarch et al., 2019). In the end, the model for the clean exchanger which gives the best tradeoff between fitness to data and complexity is a third-degree polynomial with regression parameters $\theta = \{a_0, a_1, a_2, a_3\}$:

$$U = a_0 + a_1 F_h + a_2 F_h^2 + a_3 F_h^3 \quad (4)$$

An example of the obtained fitting with this model is shown in Fig. 3. Note that (4) only depends on the hot-source flow. This is because product flow was nearly constant in the dataset and the influence of the temperature changes has resulted negligible. Note also that the values computed for the parameters θ remain valid only for equipment of similar features (size and operating conditions). Consequently, we have developed three different sets of parameters for the heat-recovery network at Lenzing, according to the existing equipment.

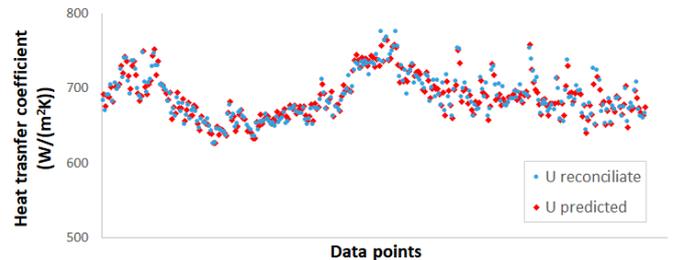


Fig. 3. Goodness-of-fit with model (4).

Remark 1. Note that, by construction, the predictions of (4) will be always higher or equal than the actual U , due to the fouling effect. Assuming the fouling dynamics is much slower than those due to changes in the control variables (F, T), such bias between the prediction of (4) and the estimated U by data reconciliation poses a way to *monitor the fouling* in heat exchangers online (Lenzing, 2019).

Hence, the final adaptive model built for U includes the current fouling state via such monitored bias as a time-varying parameter K , thus closing the gap between predictions and the reality:

$$U = a_0 + a_1 F_h + a_2 F_h^2 + a_3 F_h^3 - K \quad (5)$$

4. MATHEMATICAL REPRESENTATION

Denote by \mathcal{HS} to the set of hot sources and by \mathcal{HE} for the set of heat exchangers. Then, the set \mathcal{HE} can be divided in two ways. On the one hand, according to the product connection $\mathcal{HE} = \mathcal{HE}_s \cup \mathcal{HE}_p$, where \mathcal{HE}_s is the subset of exchangers where heating a single product connected in series, while \mathcal{HE}_p embraces those that process different

products. On the other hand, \mathcal{HE} can be split in by five subsets according to the blocks defined in Table 1.

The decision variables for optimisation are: $X_{h,w} \in \{0, 1\}$ which, if active, denotes that heat exchanger w is using hot source h ; $F_{h,w} \in \mathbb{R}$ which states the flow of h that goes to exchanger w ; $T_{out,s,w} \in \mathbb{R}$ that denote the outlet stream temperatures in exchanger w ; and $Y_w \in \{0, 1\}$ which, if active, considers the heat exchanger w fully clean².

Then, the network is modelled by the following constraints:

- The heat exchangers can only use a single hot source at a time (or none if the exchanger is not being used):

$$\sum_{h \in \mathcal{S}} X_{h,w} \leq 1 \quad \forall w \in \mathcal{HE} \quad (6)$$

- The flow of source ac through those heat exchangers connected in series that are in operation is the same:

$$\sum_{\alpha \in \mathcal{HE}_s \setminus w} F_{ac,\alpha} \leq F_{ac,w} \sum_{\alpha \in \mathcal{HE}_s \setminus w} X_{ac,\alpha} + M(1 - X_{ac,w}) \quad \forall w \in \mathcal{HE}_s \quad (7)$$

Here M is a big enough value, e.g. $M = 3F_{T,ac}$.

- Consequently, inlet temperatures of these heat exchangers linked in series by ac depend on whether the previous exchanger in the chain³ is in operation or not:

$$T_{in,ac,w} = \sum_{j=W15}^{w-1} \left(T_{out,ac,j} X_{ac,j} \prod_{j=W15}^{w-1} (1 - X_{ac,j+1}) \right) + T_{in,ac,W15} \prod_{j=W15}^{w-1} (1 - X_{ac,j}) \quad \forall w \in \mathcal{HE}_s \quad (8)$$

For the sake of clarity, (9) shows the case for the W13 inlet temperature. First $T_{in,ac,W13}$ depends on whether the W14 is connected to ac . If not, it will depend on whether W15 is using ac . Otherwise, $T_{in,ac,W13}$ will be the one at ac .

$$T_{in,ac,W13} = T_{out,ac,W14} X_{ac,W14} + T_{out,ac,W15} X_{ac,W15} (1 - X_{ac,W14}) + T_{in,ac,W15} (1 - X_{ac,W15}) (1 - X_{ac,W14}) \quad (9)$$

Production constraints must be also accomplished:

- Product streams have to reach temperature set points⁴:

$$T_{out,c,w} \geq SP_w \quad \forall w \in \mathcal{HE}_p \cup W15 \quad (10)$$

- There are some impossible connections of exchangers to hot sources (see allowed connections \mathcal{M} in Table 1):

$$X_{h,w} = 0 \quad \forall \{w, h\} \notin \mathcal{M} \quad (11)$$

- The total flow taken from each hot source by the heat exchangers has to be lower than the maximum available:

$$\sum_{w \in \mathcal{HE}} F_{h,w} \leq F_{Th} \quad \forall h \in \{alk, ualk, vap\} \quad (12)$$

In particular for the source ac , the flow used in each exchanger has to be lower than the total available.

$$F_{ac,w} \leq F_{T,ac} \quad \forall w \in \mathcal{HE}_s \quad (13)$$

- There is a flow limit per heat exchanger \bar{F}_w . Furthermore, if a hot source is not linked to a heat exchanger, the flow through it must be zero:

² In practice, $Y_w = 1$ will mean that exchanger w should be cleaned.

³ The sequence from the source side is W15→W14→W13→W12.

⁴ A single set point at W15 outlet is enough for the group \mathcal{HE}_s .

$$F_{h,w} \leq \bar{F}_w X_{h,w} \quad \forall w \in \mathcal{HE} \quad (14)$$

- Energy balances in each heat exchanger:

$$\sum_h^{\mathcal{HS}} Q_{h,w} + Q_{c,w} = 0 \quad \forall w \in \mathcal{HE} \quad (15)$$

Heats $Q_{s,w}$ are computed by (3). Note that, if a hot source is not linked to a heat exchanger, $F_{h,w}$ is set to zero and, consequently, $Q_{h,w}$ will be zero too.

- The hot streams provide the amount of heat corresponding to the heat transfer defined in (1):

$$Q'_{h,w} = Q_{h,w} \quad \forall h \in \mathcal{HS} \quad \forall w \in \mathcal{HE} \quad (16)$$

Where the data-driven model (5) can be used in the formula to compute Q' for each heat exchanger. However, as providing cleaning recommendations was one of the aims for the RTO design, (5) is slightly modified with decision variables Y_w as follows:

$$U_{h,w} = (a_0 - K_w(1 - Y_w)) X_{h,w} + a_1 F_{h,w} + a_2 F_{h,w}^2 + a_3 F_{h,w}^3 \quad \forall h \in \mathcal{S}, \quad \forall w \in \mathcal{HE} \quad (17)$$

In this way, independently of the actual fouling state, a value $Y_w = 1$ means that a more beneficial economic operation would be achieved with the exchanger w clean.

Remark 2. Note that $(a_0 - K_w(1 - Y_w))$ has been also multiplied by $X_{h,w}$ in order to fulfil the constraint (16) for the cases when $Q_{h,w} = 0$ (source h not connected to exchanger w). In such cases, (17) makes $Q'_{h,w} = 0$ feasible.

Optimisation problem. Provided the above constraints, the cost of operating the network is computed by the hot-sources usage in real time and the cleaning costs. For hot-sources consumption, the cost is computed as the total used flow in each source times a given price P_h , representing the costs of pumping, maintenance of electrical drives, etc. For the heat exchangers connected in parallel this cost is computed by J_p in (20), while for the ones connected in series by ac , as the same flow goes through them, the cost is just this flow times P_{ac} . However, due to the fact that some of these heat exchangers could not be connected to ac , we will compute such flow via the dummy decision variable F_{ac} and the linear constraints (18)-(19), giving J_s in (20).

$$F_{ac} \leq F_{T,ac}; \quad F_{ac} \leq \sum_{w \in \mathcal{HE}_s} F_{ac,w} \quad (18)$$

$$F_{ac} \geq F_{ac,w} \quad \forall w \in \mathcal{HE}_s \quad (19)$$

$$J_p := \sum_{h \in \mathcal{S}} \sum_{w \in \mathcal{HE}_p} P_h F_{h,w}; \quad J_s := P_{ac} F_{ac} \quad (20)$$

The cost P_C of performing a cleaning task is fixed, so we normalise it in (21) with respect to the time t_w that the equipment has been in operation since the last cleaning. Hence, it is comparable to sources-usage costs (20).

$$J_{clean} := \sum_{w \in \mathcal{HE}} \frac{Y_w P_C}{t_w} \quad (21)$$

Accordingly, the objective function to minimise is defined as the tradeoff existing between the above defined costs:

$$J := J_p + J_s + J_{clean} + \alpha \sum_{w \in \mathcal{HE}_s} \sum_{h \in \mathcal{S}} X_{w,h} \quad (22)$$

HE	Product stream				Hot source streams				Cleaning	Operation time (days)	Fouling [W/(m ² ·K)]			
	flow (m ³ /h)	inflow (°C)	outflow SP (°C)	outflow (°C)	alk flow (m ³ /h)	ualk flow (m ³ /h)	vap flow (m ³ /h)	ac flow (m ³ /h)				inflow (°C)	outflow (°C)	
B1	W1	15	35.0	60.0	67.0	40.92	-	-	-	70.0	58.2	✗ 0	10.00	20.00
	W2	60	25.0	40.0	40.7	36.29	-	-	-	70.0	43.9	✗ 0	15.00	30.00
	W3	50	30.0	50.0	50.0	49.62	-	-	-	70.0	49.7	✗ 0	5.00	10.00
	W4	140	47.0	55.0	55.0	54.43	-	-	-	70.0	49.3	✗ 0	0.00	0.00
	W5	130	45.0	55.0	55.0	61.93	-	-	-	70.0	48.9	✗ 0	3.00	5.00
B2	W6	50	45.0	60.0	60.0	0.00	-	32.27	-	75.0	51.6	✗ 0	3.00	6.00
	W7	25	40.0	60.0	60.0	0.00	-	30.68	-	75.0	58.6	✓ 1	17.00	35.00
B3	W8	60	20.0	40.0	40.0	0.00	131.67	0.00	-	50.0	40.9	✓ 1	22.00	45.00
	W9	60	20.0	45.0	45.0	0.00	0.00	50.95	-	75.0	45.4	✗ 0	6.00	12.00
	W10	60	25.0	40.0	40.0	0.00	76.58	0.00	-	50.0	38.2	✓ 1	12.00	25.00
	W11	45	17.0	35.0	35.0	0.00	34.95	0.00	-	50.0	26.7	✗ 0	1.00	3.00
B4	W12	50	25.0	-	45.0	-	-	21.86	0.00	75.0	29.0	✗ 0	12.00	24.00
	W13	50	45.0	-	45.0	-	-	0.00	0.00	0.0	0.0	✗ 0	11.00	22.00
B5	W14	50	45.0	-	45.0	-	-	0.00	0.00	0.0	0.0	✗ 0	18.00	36.00
	W15	50	45.0	45.0	45.0	-	-	0.00	0.00	0.0	0.0	✗ 0	9.00	18.00

Fig. 4. Designed concept for the operator dashboard. Results computed in simulation with historical data.

Note that an additional term ($\alpha > 0$ user-defined weight) is added in (22) to penalise connections of heat exchangers that are negligible to reach the temperature set points. This is to avoid an unnecessary fouling in the equipment.

The optimisation problem is then to minimise (22) subject to (6)-(19). Data about heat-exchange surfaces, product flows and inlet temperatures, temperature set points, stream densities and specific heats, temperatures at the hot sources⁵, as well as prices in (22) are known values to feed the optimisation in real time.

As non-convexities in the energy balances could not be avoided, several nonlinear constraints remain in the formulation, so the problem needs to be solved via mixed-integer nonlinear programming. Nonetheless, as it is of medium size (210 decision variables, 50 of them binaries), it is solved within Pyomo with BONMIN (Bonami et al., 2008) in half a minute over an Intel® i7-7700 CPU machine.

Note that predictive-maintenance scheduling (Palacín et al., 2018) is out of the scope of this work, so the solution obtained may not be the optimal in the long term. However, our formulation can trivially include a constraint limiting the number cleaning suggestions allowed in every execution, coping thus with issues of limited resources (e.g. available personnel) for cleaning operations.

5. VERIFICATION RESULTS

For the test we used data reflecting a plant snapshot (i.e. the actual state from a time instant) where the product flows F_c range from 15 to 140 m³/h, their inlet temperatures from 17 to 47 °C and the set points to achieve at the outlets are between 35 to 60 °C. The temperatures at the hot sources are 70°C for *alk*, 75 °C for *vap* and 85 °C for *ac*, and their availability is 300 m³/h for *alk* and *vap*, and 150 m³/h for *ac*. Values about exchange surfaces, stream densities, specific heats and prices for the hot-sources usage are omitted due to confidentiality

⁵ The inlet temperature at *ualk* is an average value computed with the *alk* temperatures and flows leaving **B1**.

reasons with the company⁶. The upper bound \bar{F}_w is set to 300 m³/h for all the heat exchangers.

By the time we performed this test, the plant historian did not record data about the equipment fouling state (values for K_w) because the method for online fouling monitoring was not implemented, yet a relative good approximation for verification purposes is to set, for instance, $K_w = 2t_w$.

The solution shown in Fig. 4 (got with zero relative optimality gap) is built upon these data. The values in blue are the data to feed the optimisation, meanwhile the values in black are the computed suggestions and predictions. The physically impossible connections between hot sources and heat exchangers are indicated with a dash.

In this test, as there is enough flow availability at the hot sources, the DSS tries to link each heat exchanger to the cheaper hot source within the permitted connections. We can observe that, for the set **B3** (W8 to W11), the DSS tries to connect as many exchangers as possible to *ualk*, as the cost of this source is zero, even suggesting for that to use more *alk* than what would be strictly necessary to reach the temperature set points in heat exchangers W1 and W2. For the heat exchangers connected in series (W12 to W15), the DSS shows that the product temperature set point can be achieved with just one exchanger in operation.

Concerning the cleaning suggestions provided by the tool, one recommendation is to clean W7 and W8 (the dirtiest), which is sensible and matches with the actual policy at Lenzing. Nevertheless, the other suggestion is to clean W10, that is dirty too but out of the top three of the dirtiest. Moreover, two of the heat exchangers suggested to clean are using *ualk*, which may look a suboptimal decision since this source is “cost free” and, consequently, the decrease of flow obtained when cleaning does not impact directly on the objective function. However, as the total availability of *ualk* depends on the use of *alk* in **B1** (exchangers that are taking more flow than the needed if clean), keeping these heat exchangers dirty provides enough flow for those using *ualk* in **B2**, saving

⁶ For better interpreting the results, we can disclose that $P_{alk} = 0.9P_{vap}$ and $P_{ac} = 1.2P_{vap}$. Note that P_{ualk} is always zero.

thus the corresponding cleaning costs plus the cost that the switching of either W8 or W10 to *vap* would incur. The last exchanger in the dirtiest top three (W14) is not suggested to be cleaned simply because it is not needed, so its cleaning does not provide any *instantaneous* saving.

6. CONCLUSIONS AND FURTHER STEPS

In this work we addressed the optimal management in real time of a heat-recovery network in a fibre-production plant. The different heat exchangers need to be allocated to the available utilities in order to heat several products. In addition, the equipment suffers from fouling in the heat-transfer surfaces, which decreases their efficiency, and cleaning operations need to be scheduled too. Our proposed DSS, based on a hybrid model, a suitable method for fouling monitoring and mathematical optimisation, helps operators and plant managers in such complex decision-making process.

From the results got in the verification stage, we proved that the current cleaning policy in the plant (dirtiest first) is not always the most economically optimal. Indeed, the potential benefits of performing a cleaning task depend on many different factors. Therefore, it is not easy for an operator/manager to infer which heat exchanger shall be clean and the right time for it. Consequently, the developed tool is very helpful to reach nearly-optimal solutions in practice while reducing the plant personnel workload. Nevertheless, it is important to remark that this tool is an RTO run, for instance, every half an hour. So it does not consider any prediction on the effects due to the fouling dynamics. Therefore the suggested maintenance actions might be suboptimal in the long term.

Future work may be pointed to formulate the problem taking into account such fouling dynamics, providing an scheduling-fashion formulation where the operation of the network is optimised for a chosen time horizon. With this formulation we would be able to really unlock the potential savings existing in a better coordination between the operation and planning layers. However, to be able to formulate the problem in this way we need a dynamic model of the fouling state with respect to the operation time, as well as reliable predictions on the product loads (inlet flows and temperatures) and utilities availability (hot sources). Furthermore, the computational complexity of this problem will increase substantially, so ways to reformulate non-convex constraints into a more computationally tractable forms need to be explored.

In addition, this heat-recovery network is part of a plant section (spinbath recovery) where other two systems have been already modelled and optimised (Marcos et al., 2018; Kalliski et al., 2019). The end goal is to join together these local RTO problems in order to solve them in a coordinated fashion, searching for the global optimum for the whole section. The integration of the other two problems was already addressed in (Marcos et al., 2019), so next step is to add the one proposed here to such integral framework.

REFERENCES

- Boccardi, G., Celata, G., Lazzarini, R., Saraceno, L., and Trinchieri, R. (2010). Development of a heat transfer correlation for a scraped-surface heat exchanger. *Applied Thermal Engineering*, 30(10), 1101 – 1106.
- Bonami, P., Biegler, L.T., Conn, A.R., Cornuéjols, G., Grossmann, I.E., Laird, C.D., Lee, J., Lodi, A., Margot, F., Sawaya, N., et al. (2008). An algorithmic framework for convex mixed integer nonlinear programs. *Discrete Optimization*, 5(2), 186–204.
- Bott, T.R. (1995). *Fouling of heat exchangers*. Elsevier.
- de Prada, C. and Pitarch, J.L. (2017). Real-time optimization (RTO) systems. *Resource Efficiency of Processing Plants: Monitoring and Improvement*.
- Dorao, C.A. and Fernandino, M. (2017). Dominant dimensionless groups controlling heat transfer coefficient during flow condensation inside pipes. *International Journal of Heat and Mass Transfer*, 112, 465 – 479.
- Enste, U. (2019). D5.4 prototype of the CoPro integration framework. Outcomes of the CoPro Project. URL www.spire2030.eu/copro.
- Galan, A., de Prada, C., Gutierrez, G., Sarabia, D., Grossmann, I.E., and Gonzalez, R. (2019). Implementation of rto in a large hydrogen network considering uncertainty. *Optimization and Engineering*, 1–30.
- Han, J., Forman, G.S., Elgowainy, A., Cai, H., Wang, M., and DiVita, V.B. (2015). A comparative assessment of resource efficiency in petroleum refining. *Fuel*, 157, 292–298.
- Hart, W.E., Laird, C.D., Watson, J.P., Woodruff, D.L., Hackedbeil, G.A., Nicholson, B.L., and Siirola, J.D. (2017). *Pyomo-optimization modeling in python*, volume 67. Springer Science & Business Media, second edition.
- Kalliski, M., Pitarch, J.L., Jasch, C., and de Prada, C. (2019). Apoyo a la toma de decisión en una red de evaporadores industriales. *Revista Iberoamericana de Automática e Informática industrial*, 16(1), 26–35.
- Khor, C.S. and Varvarezos, D. (2017). Petroleum refinery optimization. *Optimization and Engineering*, 18(4), 943–989.
- Lenzing (2019). D2.5 final report on equipment degradation modelling. Outcomes of the CoPro Project. URL www.spire2030.eu/copro.
- Marcos, M.P., Pitarch, J.L., Jasch, C., and de Prada, C. (2019). Optimal distributed load allocation and resource utilisation in evaporation plants. In *Computer Aided Chemical Engineering*, volume 46, 979–984. Elsevier.
- Marcos, M.P., Pitarch, J.L., de Prada, C., and Jasch, C. (2018). Modelling and real-time optimisation of an industrial cooling-water network. In *2018 22nd International Conference on System Theory, Control and Computing (ICSTCC)*, 591–596.
- Palacín, C.G., Pitarch, J.L., Jasch, C., Méndez, C.A., and de Prada, C. (2018). Robust integrated production-maintenance scheduling for an evaporation network. *Computers & Chemical Eng.*, 110, 140 – 151.
- Pitarch, J.L., Sala, A., and de Prada, C. (2019). A systematic grey-box modeling methodology via data reconciliation and SOS constrained regression. *Processes*, 7(3).