THEORETICAL AND EXPERIMENTAL STUDIES ON STARTUP STRATEGIES FOR A HEAT-INTEGRATED DISTILLATION COLUMN SYSTEM

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ABSTRACT

Due to their higher efficiency of energy utilisation heat-integrated column systems have been widely used in chemical industry. However, the heat-integration leads to difficulties in startup of such columns, i.e. a long startup time and thus considerable costs will be resulted. In this work, a study consisting of modelling, simulation, optimisation and experimental verification is made for developing optimal operation strategies for heat-integrated columns to reduce the startup time. A pilot two-pressure column system with bubble-cap trays is considered. As a result, more than 35% of the startup time can be reduced in comparison to conventional startup procedure. Heuristics for startup operation of such processes are suggested.

INTRODUCTION

Process intensification has been recently extensively studied and applied in chemical industry with the aim of reducing the costs of equipment and operation. An example of such intensification is the adoption of heat-integrated distillation column systems. Heat integration of two columns is based on the idea of utilising the overhead vapour of one column to provide the latent heat for boiling up the other column. Thus the energy consumption can be reduced while accomplishing the same separation task. However, the heat-integration leads to difficulties in startup operation due to strong coupling of the two columns. A long time period is usually needed for startup of the process. A long startup time costs considerable amount of energy and off-product. Thus optimal strategies are desired from industry practice to shorten the startup period.

In spite of its importance, very few previous work has been done on startup optimisation for distillation processes. Conventionally, the values of control variables corresponding to the steady state values are set to the columns for startup and one just waits for columns running to the desired steady state (so called direct setting

strategy). Empirical startup strategies like total or zero reflux flow and maximum reboiler duty for single columns have been proposed [1, 2, 3] to improve the startup performance. Since a column startup is influenced by many factors such as column structure, type of trays and packings, components in the mixture to be separated as well as top and bottom specifications, these empirical strategies are suitable only for some specific cases. Therefore, systematic approaches concerning these influential factors are needed in order to solve general startup problems for distillation column systems. This calls for methodologies of modelling, simulation and optimisation. In addition, due to their higher complexity, very few studies have been made on startup of heat-integrated columns [4].

The essential difficulty in modelling column startup lies in the fact that it is a complex dynamic process. In most startup models for distillation columns, the "three-phase-model" (discontinuous, semicontinuous and continuous phase) proposed by Ruiz et al. [5] has been used. The discontinuous phase is the time period from an empty cold column to the beginning state of equilibrium. Hangos et al. [6] studied this discontinuous phase with a simplified non-equilibrium model. Wang et al. [7] proposed a model considering the tray-by-tray state transfer from non-equilibrium to equilibrium that takes place at the boiling temperature at the operating pressure. Compared with the other two phases in which the column is in the vapor-liquid equilibrium state, this discontinuous phase is much shorter. In addition, hydrodynamic properties on trays and packings in the column play an important role in modelling startup processes and thus should be considered. The model equations result in a large-scale nonlinear differential algebraic equation (DAE) system.

Based on an established model, simulation can be made by solving the DAE system to study the startup behaviour [7, 8, 9]. For simulation an operating policy during startup has to be defined *a priori*. This means it may be neither optimal (in the sense of minimising the startup time) nor feasible (in the sense of holding the process constraints, e.g. the product specifications at the desired steady state). Thus, a mathematical optimisation has to be employed to search for an optimal as well as feasible operating policy. Optimisation approaches to solving large-scale problems have been proposed in the previous studies [10, 11, 12]. The basic idea of these approaches is to discretize the dynamic system into a large nonlinear programming (NLP) problem so that it can be solved by an NLP solver like sequential quadratic programming (SQP). These dynamic optimisation approaches can be used to solve startup optimisation problems for distillation columns.

In this work, a systematic study on startup strategies for heat-integrated distillation columns is made. A pilot two-pressure heat-integrated column system is considered. The process is modelled with a detailed dynamic model consisting of the MESH equations and hydraulic relations. The model is validated by experimental studies on the pilot plant. Simulation and optimisation are carried out for a preliminary study to gain a rough insight into the dynamic behaviour and to develop the tendency for optimal operating policies with the aim of reducing the startup time. The optimisation approach used is a sequential method with a capability of dealing with large-scale problems. Extensive experiments are conducted to test these results and to adapt them to different column configurations. As a result, significant reduction of the startup time can be achieved by implementing the optimal operating policy. Based on

the results of simulation, optimisation and experimental study, some heuristic rules for optimal startup for heat-integrated column systems are suggested.

PLANT AND MODEL DESCRIPTION

As shown in Fig. 1, the plant considered is a pilot two-pressure column system consisting of a high pressure and a low pressure column both with a diameter of 100 mm. The columns have a central down-comer with 28 and 20 bubble-cap trays, respectively. The overhead vapour from the high pressure column (HP) is introduced as the heating medium to the reboiler of the low pressure column (LP). The plant is so constructed that it is possible to operate the process in downstream, upstream and parallel arrangements. Fig. 1 (right) shows the flow-sheet of the plant with the parallel arrangement. The plant is equipped with temperature, pressure, level and flow rate measurements and electrical valves for flow control. All input/output signals are treated by a process control system. For the experimental study, startup of the plant to a steady state for the separation of a methanol-water mixture is considered. The reboiler duty of HP and reflux flow rate of both HP and LP are the manipulated variables to be optimised for the startup procedure.

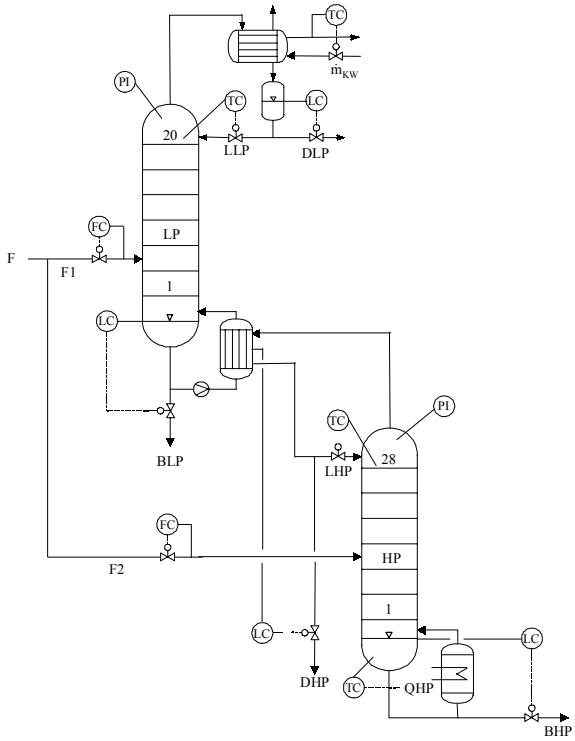

Fig. 1: The two-column system: photo (left) and flow-sheet (right).

To model the pilot plant for startup, two major assumptions (simplifications) are considered. First, the discontinuous phase is not considered in the model, since it is relatively short (about 20min for the pilot plant) in comparison to the other two phases during startup. It means we start to describe the plant at the time point when the vapor reaches the top of the columns, i.e. an equilibrium model beginning with a

pseudo (warm) starting point is used. Second, modelling the build-up of the pressure of HP during startup is a very difficult task. Thus we assume a fixed pressure profile during the startup, which is gained based on experimental studies. Based on these two assumptions, the well-known tray-by-tray model can be used to model the process. The equations of each tray consist of three parts: mass and energy balances, phase equilibrium and tray hydraulics relations. The dynamic balances of mass, component and energy lead to a set of differential equations. The heat capacity of the metal of the column wall and the tray elements are considered in the enthalpy balance. The vapour-liquid equilibrium is described by the Wilson and Antoine equations. The equilibrium state is corrected by Murphree tray efficiencies of stripping and rectifying sections. The tray hydraulics is modelled with the Francis-weir-formula to correlate the tray holdup with the liquid flow. The tray pressure drop is calculated by the gas and liquid fluid dynamics based on the tray geometric sizes.

For the dynamic simulation, collocation on finite elements is applied for the discretization of the differential equations in order to transform the DAE system to a system of nonlinear algebraic equations. This collocation method has the advantage that even with relatively large time intervals an acceptable accuracy can be achieved. Hence the time intervals for the piecewise constant control variables can be identical with those used for discretization. In this study, we chose three collocation points in a time interval with a length of 0.25 h. This simulation is used at first for the validation of the model.

An initial state has to be defined to formulate the dynamic optimisation problem. In the reality, it should be the cold and empty state of the two columns. With the model used, we assume the starting state at the time point when the pressure of HP reaches 2.5 bar, at which both columns just approach the vapour-liquid equilibrium and the heat-integration already takes place. The state variables at this initial state can be computed with help of the measured data in the past experiment. Furthermore, the profile of the pressure increase in the HP from that state on has to be estimated. As a base for estimating this pressure profile, we use the curve resulted from the conventional operating strategy (see Fig. 6) which can be approximated with the following relation:

$$P_{top}^{HP}(t) = (P_0 - P_E) \cdot e^{-ct} + P_E \qquad (1)$$

with P_0 (here 2.5 bar) and P_E (here 4.66 bar) as the initial pressure at which the air in the HP is completely pushed out and the final steady state pressure, respectively. The constant c is validated by experimental data. It has the value of 0.458h^{-1}, approximately.

To verify the model, experimental work was done at first for determining the vapor and liquid load limits. The tray efficiencies and heat losses of the two columns were validated by comparing the simulated and measured data. The following parameters have been achieved by comparing simulated results and measured data: $\eta_R = 0.85$; $\eta_S = 0.80$; Cw = 7.614; ε = 0.5; $\xi_W = 2.3046$, with η_R and η_S as the tray efficiencies of the rectifying and stripping section of both columns, respectively. C_w is the weir constant and ε is the volumetric liquid fraction in the bubble area in defining the holdup relations for each tray. ξ_W is the friction factor needed for calculating the dry

pressure drop caused by the vapour load. These values were fitted to several steady state points within the load limits of the plant.

A special property of the model is the heat-integration between the two columns, i.e. the heat exchange between the overhead vapour of HP and the liquid of the bottom of LP. The heat delivery Q_{Con}^{HP} from the overhead vapour of the HP can be calculated as follows:

$$Q_{Con}^{HP} = h_{Con}^{V} \cdot V - \frac{d(HU \cdot h^{L})}{dt} - (1+\nu) \cdot D \cdot h^{L} \qquad (2)$$

with h_{Con}^{V} as the enthalpy of the vapour entering into the heat exchanger, V as the molar flow of this vapour, HU as the liquid holdup in the HP section of the heat exchanger, h^{L} as the enthalpy of the liquid leaving the heat exchanger, ν as the reflux ratio and D as the distillate flow rate of the HP. In the ideal case, the energy flow supplied by the top of the HP totally goes to the reboiler in the bottom of the LP. In the reality, however, this heat exchange causes a heat loss Q_{Loss} to the environment, which is not negligible. Thus the reboiler duty of the LP Q_{B}^{LP} is:

$$Q_{B}^{LP} = Q_{Con}^{HP} - Q_{Loss} \qquad (3)$$

The value of Q_{Loss} strongly depends on the specific construction. A value of 2.5 KW was approximated with help of experimental results of several steady state points. The temperature difference between the two outflows of the heat-integrator is set to be constant at 8°C which conforms with the most experimental cases. In addition, reflux temperature of the LP is controlled with a subcooling for 5°C, which is described in the model.

As a result, a complex large-scale DAE system is formulated with 468 state variables and corresponding equations which can correctly describe the dynamic behavior. A detailed explanation of this model can be found in [13]. Investigations on product changeover of this pilot plant were accomplished based on this model [14, 15]. This model includes the relation between startup performances and the influential factors. These consist of the plant data (e.g. column structure, type of trays), the component properties in the mixture and the impact of control variables (reflux flow and reboiler duty). Therefore the model provides a basis for the subsequent study on startup optimisation for the column system.

STARTUP OPTIMIZATION

Since the minimization of startup time is desired in most industrial practice, we consider the time-optimal problem for the startup of the heat-integrated column system. The end of the startup period is defined as the time when the column system has arrived at the desired steady state. The time point can be defined as the objective function, which is to be minimised, subject to the steady state conditions, which mean that all time derivatives of the state variables approach to zero. Since

this kind of formulation may cause numerical expenses, the following formulation can be used as an alternative to describe this optimisation problem

$$\min \int_{t_0}^{t_f} \left[\omega_1 \left(x_D^{HP} - x_D^* \right)^2 + \omega_2 \left(x_D^{LP} - x_D^* \right)^2 + \omega_3 \left(x_B^{HP} - x_B^* \right)^2 + \omega_4 \left(x_B^{LP} - x_B^* \right)^2 \right] dt \qquad (4)$$

where x_D^* and x_B^* are the distillate and bottom product specifications, respectively, for the light component. t_0 is the initial time point and t_f the final time point considered for the startup optimisation. In this study, the pilot plant with the parallel arrangement to separate a mixture of methanol and water is considered, i.e. the total feed flow is splitted into two parallel flows to the two columns and both have top and bottom product. This arrangement represents a typical operation case of such processes in chemical industry. The input constraints are the limitations of the control variables, i.e. the reflux flow of both HP and LP and the reboiler duty of HP:

$$0 \le L^{HP}(t) \le L_{\max}^{HP} \qquad (5)$$

$$0 \le L^{LP}(t) \le L_{\max}^{LP} \qquad (6)$$

$$0 \le Q^{HP}(t) \le Q_{\max}^{HP} \qquad (7)$$

For the startup optimization problem, the feed condition (flow, composition and temperature) is given. It is determined such that it will lead to a proper vapour and liquid load for the two columns. Now a large-scale dynamic nonlinear optimization problem is formulated. To solve this problem, the sequential optimization method proposed by Li et al. in [12] is used. In this approach, the entire computation is divided into one layer for optimization and one layer for simulation. The model equations are integrated in the simulation layer, so that the state variables and their sensitivities can be computed by given controls. The control variables, defined as piecewise constants, are computed in the optimization layer by SQP as the only decision variables. Collocation on finite elements is applied for the discretization. A detailed derivation of the optimization approach can be found in [12].

An initial state has to be defined for the dynamic simulation and optimization. As mentioned in the previous section, we generate the initial state from the time point at which the pressure of the top of HP has reached the value of 2.5 bar. The temperatures of the product flows as well as the other measured data at this time point are used for the simulation, through which all other state variables can be computed and thus the initial state is determined.

To deal with the time-optimal problem according to equation (4) a sufficiently large time horizon has to be chosen for the integration, in order to ensure the plant to reach the steady state. The whole time horizon is divided into a certain number of time intervals of equal length in which the control variables defined in (5) – (7) are kept as piecewise constant. Specifically for this case we chose a time horizon of 5h which is divided into 20 time intervals. The final optimisation results are illustrated in Fig. 2 showing the optimal trajectories of the reboiler duty for HP and the reflux flow for both columns. The corresponding results are illustrated in Fig. 3 and Fig. 4 with the profiles of the product compositions and the top and bottom temperatures.

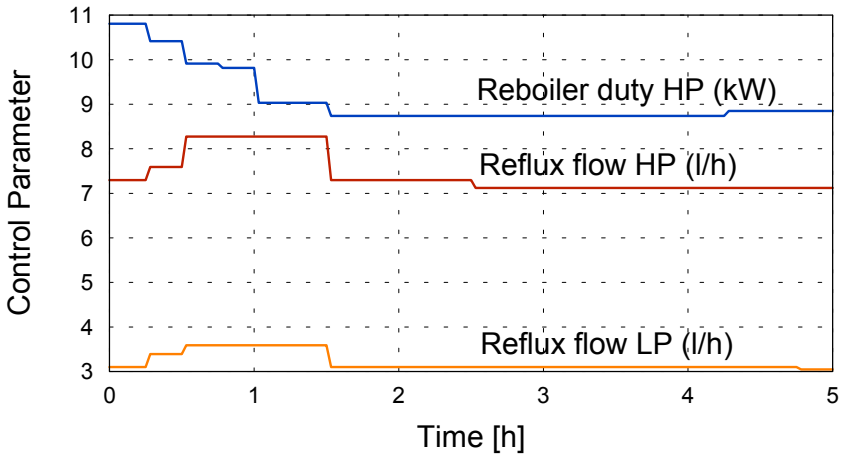

Fig.2: Optimal trajectories of the control variables.

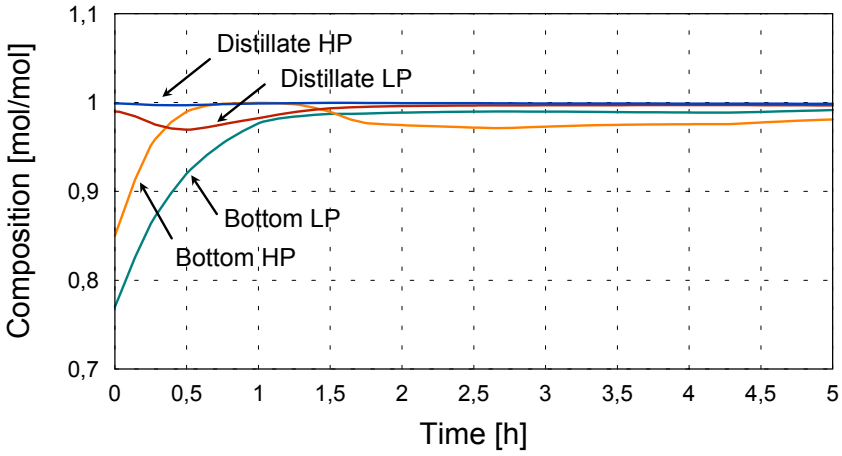

Fig. 3: Optimal composition profiles during startup.

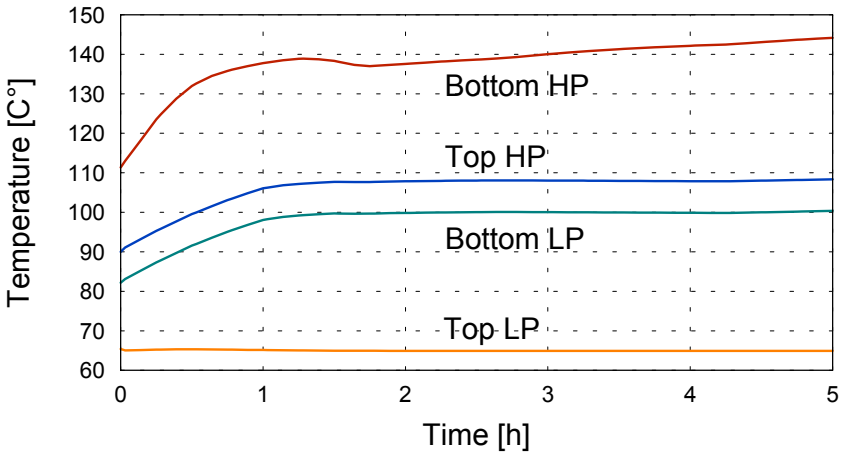

Fig. 4: Optimal Temperature profiles during startup.

As shown in Fig. 2 the results of the numerical optimisation illustrate that at the beginning a high value of the reboiler duty should be chosen since it is necessary to increase especially the bottom temperatures and thus the purity of the products. With a slight time delay, the two refluxes need to be increased in order for the two columns to approach the desired product purities as fast as possible, as shown in Fig. 3 and 4. However, as soon as the column system reaches close to the steady state, the control parameters need to be step by step decreased down to their steady state level. It should be noted that the optimiser is based on the model which reflects the steady state points of the pilot plant fairly well and also depicts roughly the shape of the experimentally measured profiles, but there is still a large mismatch concerning the time delay.

However, a startup rule for practical purpose can be obtained from the optimisation results: It can be seen in Fig. 2 that there is one certain time point at which the control parameters have to be decreased fairly drastically. The decrease for the reboiler duty has to be made, when the temperature of the bottom in the LP has reached the value of approximately 98°C. Since this temperature is the most major concern for startup operation of this plant, this value can be used as a switching criteria. Furthermore, it can be noted that the ratio between the maximum value and the steady state value of the control is approximately 1.2. To transfer the numerical optimisation results to a practical operation policy the maximum strategy can be derived: the plant should be first operated with the maximum value for all control parameters until the bottom of the LP reaches its switching temperature and then all the control parameters should be gradually switched to their steady state values. This strategy has been realised on the pilot plant. The experimental results are presented in the following section.

EXPERIMENTAL RESULTS

The parallel arrangement shown in Fig. 1 (right) was selected for experimental study to verify the above theoretical startup investigation. For comparison of startup time, different operating policies were implemented in different startup runs. A feed flow of a methanol-water mixture ($F = 34.5\ \text{l/h}, x_f = 0.3\ \text{mol/mol}, T_f = 60\ °\text{C}$) is to be separated. The feed flow of the two columns are set as constant with their steady-state value ($F_1 = 19\ \text{l/h}, F_2 = 15.5\ \text{l/h}$). The feed tray is set at 10$^{\text{th}}$ tray for HP and 6$^{\text{th}}$ tray for LP. The operating pressure at steady state is 4.7 bar for HP and atmospheric for LP.

The startup of the pilot plant from a cold empty state (at atmospheric temperature and pressure) to the desired steady state ($x_D^* = 0.99,\ x_B^* = 0.01\ \text{mol/mol}$ for both columns) was considered. Based on simulation as well as experiment results, the bottom and top temperatures of the two columns corresponding to the steady state purity specifications are $T_{Bss}^{HP} = 144\ °\text{C}, T_{Tss}^{HP} = 108\ °\text{C}$ and $T_{Bss}^{LP} = 99\ °\text{C}, T_{Tss}^{LP} = 65\ °\text{C}$, respectively. The corresponding values of the control variables at this steady state are $Q_{ss}^{HP} = 9.54\ \text{kW}, L_{ss}^{HP} = 26\ \text{l/h}, L_{ss}^{LP} = 11.2\ \text{l/h}$. During startup, the level and flow control loops shown in Fig. 1 (right) are active, while the temperature control loops (except for the cooling water temperature control) are set to open loop in order to manually implement different startup policies.

The conventional **direct setting strategy** was first implemented for startup, i.e. the steady state control values were set during the run. Fig. 5 shows the measured temperature profiles of the two columns. It can be seen that there is a time delay of about 0.8 h for the top temperature of LP (T_T^{LP}) to begin to increase after the top temperature of HP rises (T_T^{HP}). This is because the driving force of LP is from the latent heat of the overhead vapour from HP. However, despite the delay, T_T^{LP} first reached its steady state value. It took 7.75 h for all temperatures to approach their steady state. This means that to reduce the startup time for the plant, a control policy should be designed to accelerate the increase of T_B^{HP}, T_T^{HP} and T_B^{LP}. Moreover, it is shown that the warm-up time of the two columns (i.e. the discontinuous phase) was about 10 and 20min, respectively.

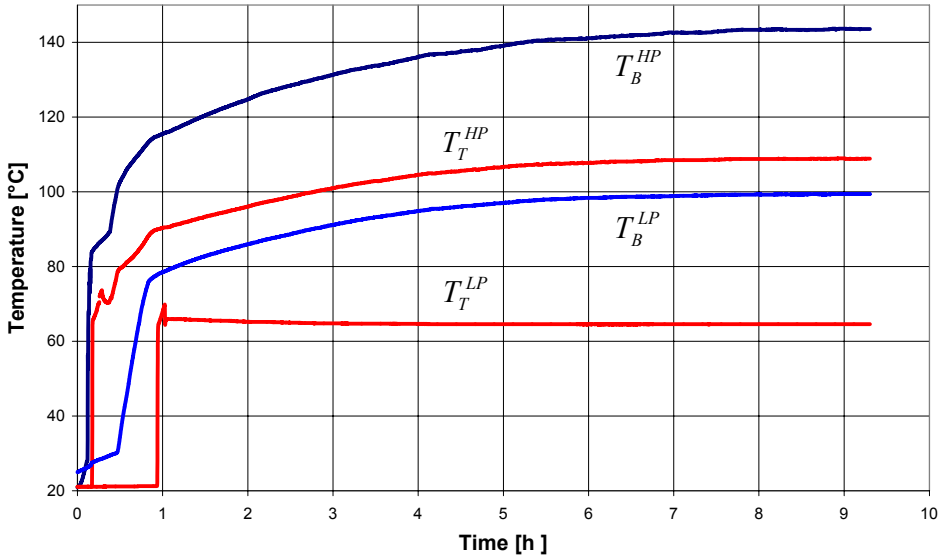

Fig. 5: Measured temperature profiles by direct setting strategy.

The second strategy implemented was the **total reflux strategy**. The basic idea of this strategy is the employment of total reflux for the top temperature to quickly reach its desired value. The reboiler duty is set at the steady state value during the startup. A switching time from total reflux to the steady-state reflux flow has to be defined. This can be made by observing the bottom temperature of both columns. At the beginning they will be rising for a period and cease to increase at a certain time point due to the total reflux. This time point was determined for the switching. The experimental result shows, however, that the improvement of startup time by the total reflux strategy is marginal, compared with that of direct setting. This is because the main problem in startup of this plant is not the top but the bottom temperature which has a large time constant. The minor improvement of startup time from total reflux lies in the quicker increase of the pressure in HP due to the higher reflux flow, as shown in Fig. 6.

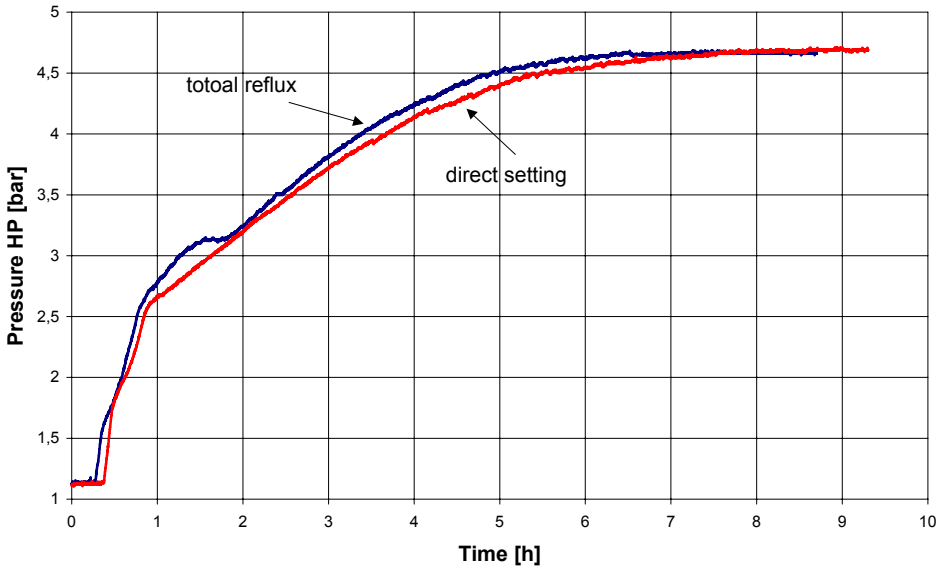

Fig. 6: Measured pressure profiles by direct setting and total reflux strategy.

The third strategy studied was the **total distillate strategy** proposed in [2]. In a beginning period the condensed liquid from both columns was drawn out as distillate (zero reflux) and then the operation was switched to the reflux flow for the steady state value. The reboiler duty remained constant during startup. This strategy has the advantage to accelerate the rise of the bottom temperature of both columns. The switching point from zero reflux to normal reflux was decided by the minimum point of the following function

$$M_T = \sum_{i=1}^{N} \left| T_i - T_{ss,i} \right| \qquad (6)$$

which is the sum of discrepancies of measured temperatures on all trays (there is a temperature sensor on each tray for the pilot plant) and their desired values at the steady state. The value of this function was on-line computed and observed, through which the switching time point can be determined.

Fig. 7 shows the measured temperature profiles caused by the total distillate strategy. Due to zero reflux before the switching, the temperature increased with a fast speed. An over-shooting of the top temperature of both columns was resulted. Because of the heat-integration, the bottom temperature of LP is stagnated without over-shooting. Switching the reflux flow to their steady state value led to a decrease of the top temperature of both columns. The total startup time for all temperatures to their steady-state value was reduced to 6.72 h. Fig. 8 shows the profiles of the function values of (6) for the two columns. It indicates that this startup strategy causes a rapid approaching to the steady-state before the switching but a slow approaching after that.

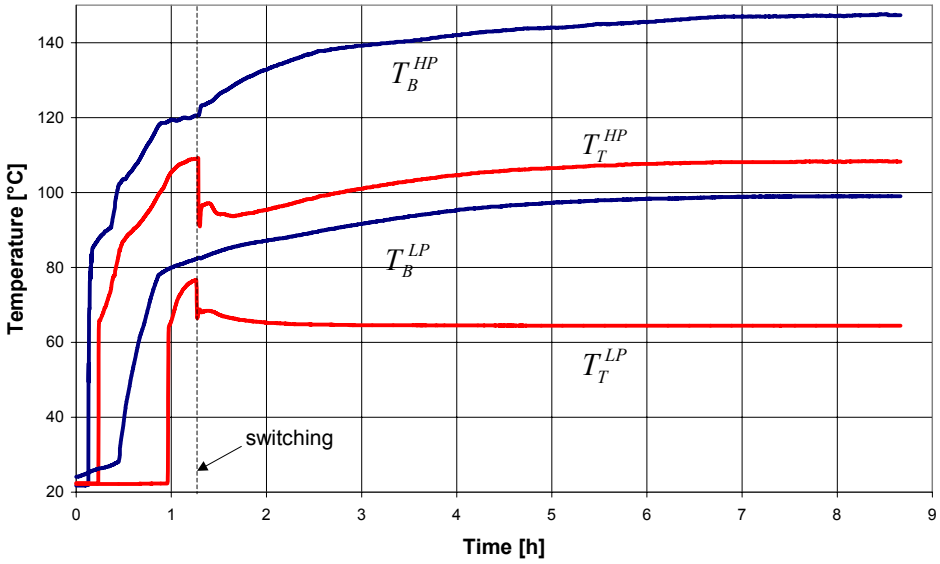

Fig. 7: Measured temperature profiles by total distillate strategy.

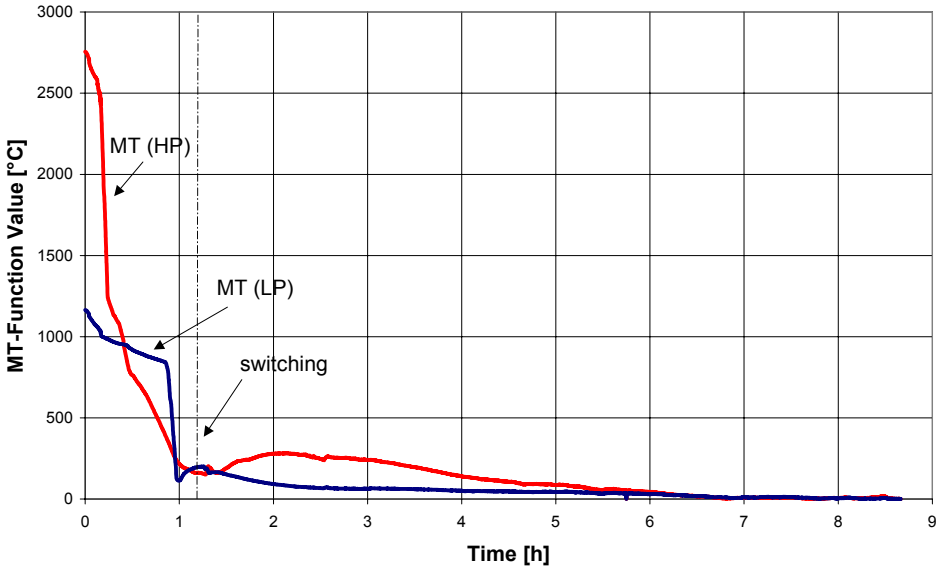

Fig. 8: M_T – Function profiles of HP and LP by total distillate strategy.

The **optimal strategy** developed by the optimisation was implemented in the fourth experimental run. From the optimisation results shown in Fig. 2, the control variables (reflux flow for both columns and reboiler duty for HP) should be first set at an optimised maximal value for a time period and then decreased gradually to their steady state value. From the practical point of view, it is desired to implement a simple startup policy. Therefore, we simply tailored the numerically optimised operating policy shown in Fig. 2 into a two-stage strategy. The optimised maximal values of the control variables ($Q_{opt}^{HP} = 11.14$ kW, $L_{opt}^{HP} = 29$ l/h, $L_{opt}^{LP} = 13.5$ l/h) were taken in the first period and the steady-state values were set in the second period to the

pilot plant. The switching time point is chosen at the time when the bottom temperature of LP reaches 98°C that is almost approaching its steady state value.

Fig. 9 shows the top and bottom temperature profiles of both columns by the optimal strategy. With the enhanced values of the control variables, T_B^{HP}, T_T^{HP} and T_B^{LP} increased fast to their steady state temperatures before the switching. After the switching the two columns continued running to the desired steady state. It can be seen that HP ran a little bit over before the switching in order to provide enough energy to LP, so that both columns could approach the desired steady state as quickly as possible. Compared with the results of the other startup strategies (Fig. 5 and 7), the optimal strategy really resulted in the best startup performance and thus took the shortest startup time.

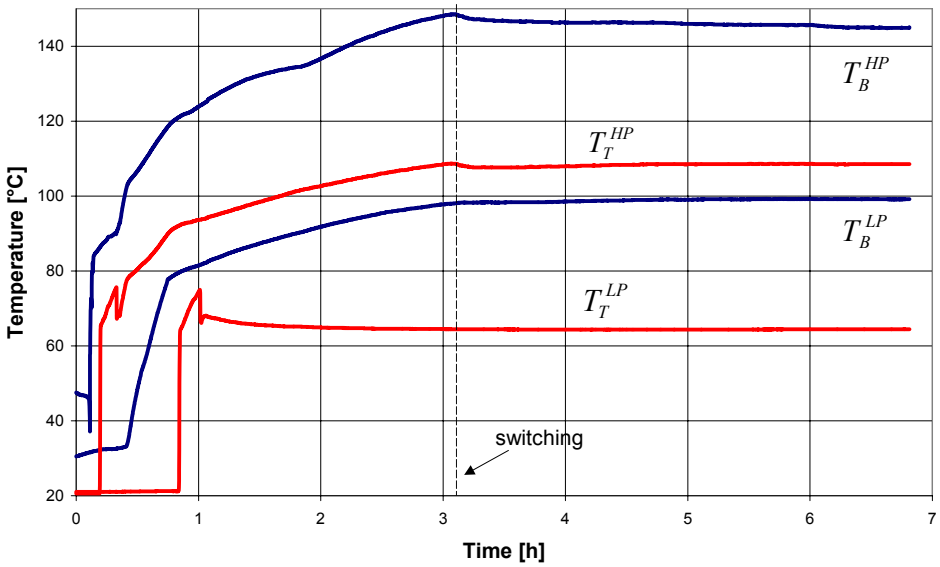

Fig. 9: Measured temperature profiles by optimal strategy.

To gain a further comparison, an additional run with *only* the optimised maximal value for the reboiler duty of HP was carried out. The reflux flow of both columns remained constant at their steady-state value. The switching from Q_{opt}^{HP} to Q_{ss}^{HP} was also determined as the time point when $T_B^{LP} = 98\,°\text{C}$. We call this strategy **enhanced reboiler duty strategy**. It is obvious that this strategy is easier to implement than the optimal strategy. Fig. 10 shows the measured temperature profiles. Compared with Fig. 9, the temperature increase caused by this strategy before switching was, due to fewer reflux, faster than that by the optimal strategy. It is interesting to note the slight decrease of the bottom temperature of HP before the switching. This was caused by the rising composition of water at the top of HP which led to a significantly improved heat exchange between the condenser of HP and the reboiler of LP. The enlarged condensation of the vapor of HP resulted in a pressure drop, as shown in Fig. 11, and thus led to a decrease for the bottom temperature. The pressure incresed again after the reboiler duty was switched to Q_{ss}^{HP}. From this analysis, an oversupplied reboiler duty will lead to a slower startup and thus should be avoided.

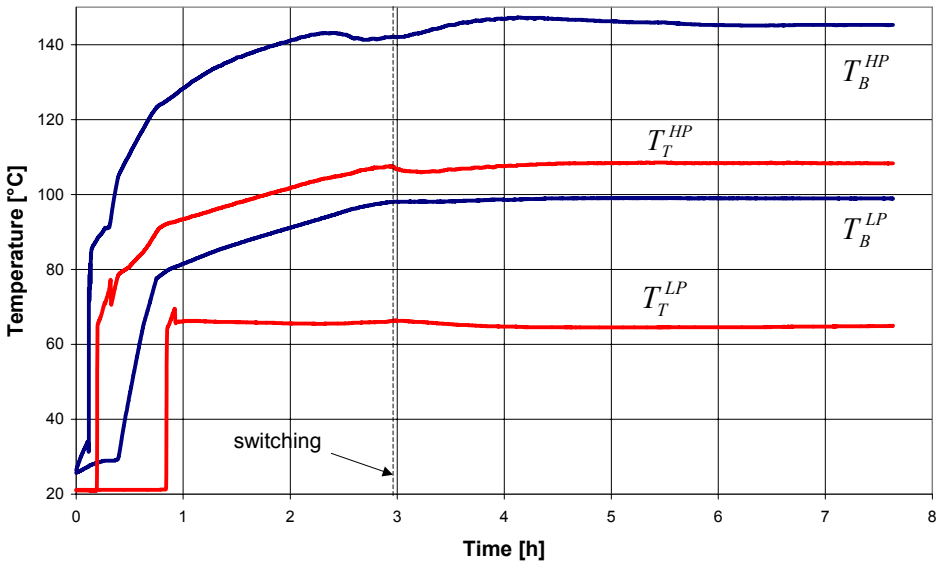

Fig. 10: Measured temperature profiles by enhanced reboiler duty strategy.

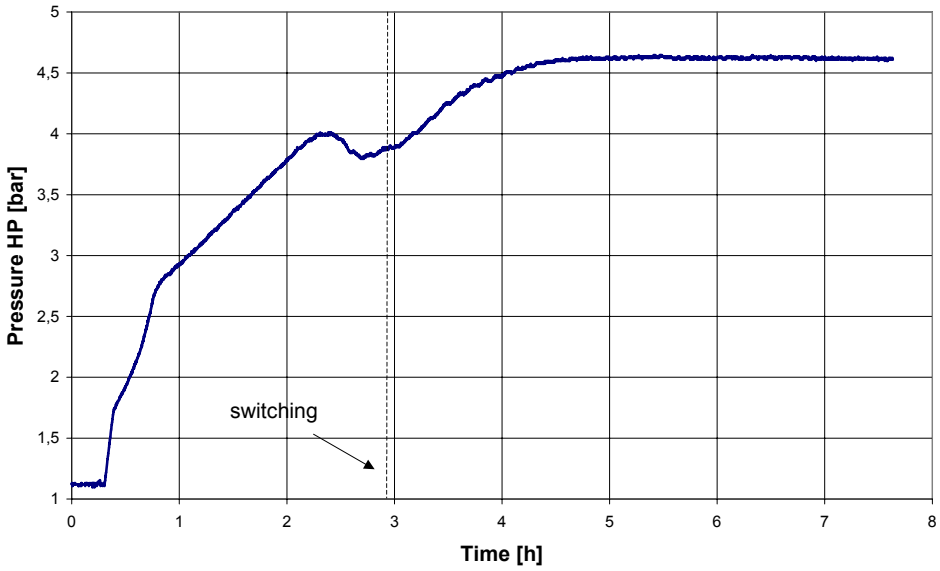

Fig. 11: Measured pressure profile by enhanced reboiler duty strategy.

The experimental results by the strategies described above are summarized in Table 1 showing the time points at which the measured temperatures (T_B^{HP}, T_T^{HP}, T_B^{LP}, T_T^{LP}) reached their steady state values (T_{Bss}^{HP}, T_{Tss}^{HP}, T_{Bss}^{LP}, T_{Tss}^{LP}). It can be seen that by all strategies the reboiler temperature of both columns was more sluggish than the top temperature. With the optimal strategy, the total startup time was about 5h, which is 64% of the time needed by the direct setting strategy.

Table 1: Time period (h) of reaching the steady-state values of the temperatures

	Direct setting	Total reflux	Total distillate	Optimal	Enhanced reboiler duty
T_T^{LP}	3.15	3.45	2.88	3.5	3.83
T_B^{LP}	7.73	6.38	6.72	4.78	6.05
T_T^{HP}	7.53	6.52	6.1	4.6	4.5
T_B^{HP}	**7.75**	**7.53**	**6.72**	**4.95**	**5.83**
%	100%	97%	87%	64%	75%

CONCLUSIONS

Due to its feature of multiple variables and complex dynamics the startup of heat-integrated distillation columns represents a difficult operation. Besides the control variables, factors of both column structure and component properties of the mixture have impact on the startup performance. Therefore, it is impossible to derive general rules for startup operation of common heat-integrated column systems. Experiences can bring about improvement for startup but for an optimal operation a systematic investigation is required.

In this work, a pilot two-pressure column system was considered for a theoretical and experimental study on startup strategies. A model concerning the different influential factors was used. A mathematical optimisation was carried out to develop optimal operating policies for the columns system. In the experimental study, different startup strategies were tested and the results verified the optimality of the optimal strategy.

From this work, the following simple operating policy can be suggested for starting up heat-integrated column systems in parallel operating arrangement. The feed splitting to the two columns can be set constant at the desired steady state value. An enhanced value (i.e. approximately 1.2 time of the steady state value) for the reflux flow of both columns and the reboiler duty of HP can be implemented in the first period to accelerate the startup. When the reboiler temperature of LP almost reaches its steady state value, all thee control variables should be switched to their steady state value, allowing the columns running to the desired steady state.

It should be noted that this strategy is derived mainly to minimize the startup time. Due to the maximum values the energy consumption could be larger than necessary. If we chose the minimization of the energy consumption as the objective function in the optimization problem, different trajectories of the control variables can be expected, most likely a strategy reflecting a trade-off between the startup time and the reboiler duty of the HP. In addition, the comparison of results of minimizing off-spec products, energy consumption and startup time can be made. These can be work in the future study.

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