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# Advanced Process Control Design for a Distillation Column Using UniSim Design

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***Abstract*—The paper addresses implementation of advanced predictive control (APC) for a distillation column. The APC controller was designed using Profit Design Studio software. The distillation column was modeled and the closed-loop control was implemented in UniSim Design environment. The distillation column was handled as a multiple-inputs and multiple-outputs system. Moreover, constraints on the controlled and manipulated variables were considered. APC controller ensured good control performance.**

## I. Introduction

Distillation columns are key devices in the petroleum industry, as they are frequently used separation processes. The disadvantage is that distillation columns are energy demanding devices and therefore it is necessary to implement advanced control strategies. Moreover, industrial implementations of distillation column control have to ensure a quality of products, yields, safety operation, environmental limits, maximal profit, see e.g. [1], [2], [3, pp.1820–1854], and references therein.

Model predictive control (MPC) represents the state of art in model-based control strategies, as it enables to optimize control performance in the presence of constraints. Industrial implementations of MPC begun back in the 70's especially for processes in the petroleum industry, see e.g. [4]. The principles of MPC were implemented under the original name Dynamic Matrix Control (DMC) [5]. MPC attracted the high interest of the researchers in past three decades, and the significant progress was reached, see [6]. Widely-used receding horizon policy enabled to reduce the *process-model mismatch*, see e.g. [4]. An extensive survey on industrial application of MPC is in [7].

Advanced process control (APC) usually represents an optimal control strategy implemented in addition to basic low-level process controllers. The basic controllers ensure set-point tracking and disturbance rejection. APC supervises the basic controllers, optimizes set-points for them and helps to decouple and minimize the side effects of interactions between multiple process variables [8].

The main benefits of APC are listed in [9]. Compared to other well-known strategies, APC increases the yields of products, decreases the operation costs, and improves the process safety. Moreover, the common payback period of APC implementation is less than six months [9].

Optimal control of refinery units in the petrochemical industry represents a challenging task due to the complex interactive problems. In [10] the model of a de-propanizer unit was derived and dynamic control based on the non-linear optimization problem was designed. Decentralized adaptive control of a distillation column was designed in [11]. In [12] the laboratory distillation column was controlled in a real-time framework using MPC based on the state estimation and disturbance modeling. The authors of [1] implemented the regionless explicit MPC for the same laboratory distillation column.

There are various free and commercial software tools to simulate the behaviour of complex systems, e.g., MATLAB/Simulink environment by MathWorks [13], Octane Software by Crunchbase, ASPEN by Aspen Technology [14], UniSim Design by Honeywell [15], etc. We used UniSim Design environment to implement an advanced control strategy, as this software is widely used in industry and academia and provides high-quality libraries for simulating the complex dynamics of various process units.

In this paper, APC of a distillation column is presented. This work extends the master thesis [16]. The controlled plant was a de-propanizer unit. The APC controller was designed using Profit Design Studio software. The distillation column was modeled and the closed-loop control was implemented in UniSim Design environment. The distillation column was handled as a multiple-inputs and multiple-outputs system. Moreover, constraints on the controlled and manipulated variables were considered.

The paper is organized as follows. Section II presents the considered control plant. Section III describes the APC design problem. Particularly, it formulates the optimization problem, sets the control conditions, and presents the identified model of the distillation column that is used for predictions of the future system behavior. Section II discusses simulation results of the closed-loop control performance. The conclusions are summarized in Section V.

### A. Notation

The following notation has been used in the paper:

1) $\mathbb{R}^n$ denotes the $n$-dimensional space of real-valued vectors, $\mathbb{R}^{n\times m}$ represents the $(n \times m)$-dimensional space of real-valued matrices.
2) For a real-valued matrix $A$, $A^\top$ denotes its transposition and $A^{-1}$ denotes its inverse, if exists.
3) For a real-valued vector $x$ and positively defined matrix $A$, $\| x \|_A^2 = x^\top A x$.
4) For a real-valued time-varying vector $y$, $y(k+p|k)$ denotes the value of vector $y$ in $(k+p)$-th control step

predicted in $k$-th control step. Analogous notation holds for $u(k+p|k)$.

## II. Distillation Column

The controlled process was the de-propanizer unit. The scheme of the considered distillation column equipment setup is depicted in Fig. 1. It was the multi-component distillation column used for separation of the ten-component mixture of methane, ethane, propane, i-butane, n-butane, i-pentane, n-pentane, n-hexane, n-heptane, and n-octane. In Tab. I are summarized the mole fractions of all components in the feed. The *light key* and the *heavy key* of the de-propanizer unit were propane in the bottom product and i-butane in the distillate, respectively.

The distillation column had 10 trays, a condenser, and a reboiler. The trays were numbered from the top to the bottom, and the feed entered into the 5th tray. The steady-state mass balance of the de-propanizer unit was

$$\dot{m}_{\text{in}} = \dot{m}_{\text{dis}} + \dot{m}_{\text{btm}}, \tag{1a}$$
$$\dot{m}_{\text{vap}} = \dot{m}_{\text{dis}} + \dot{m}_{\text{rfx}}, \tag{1b}$$
$$\dot{m}_{\text{liq}} = \dot{m}_{\text{vap}} + \dot{m}_{\text{btm}}, \tag{1c}$$

where $\dot{m}_{\text{in}}$ is the mass flow of the feed, $\dot{m}_{\text{dis}}$ is the mass flow of the distillate, $\dot{m}_{\text{btm}}$ is the mass flow of the bottom product, $\dot{m}_{\text{vap}}$ is the mass flow of the vapour stream, $\dot{m}_{\text{rfx}}$ is the mass flow of the reflux, and $\dot{m}_{\text{liq}}$ is the mass flow of the liquid stream to the reboiler. The properties of the distillation column's streams in a steady state are summarized in Tab. II, where $\dot{m}$ denotes the mas flow, $T$ is the temperature, $P$ is the pressure, and $x$ is the liquid-vapour phase fraction.

The total condenser was used and the pressure $P_{\text{con}} = 1379\,\text{kPa}$ was in it. The pressure in the reboiler was $P_{\text{reb}} = 1413\,\text{kPa}$.

The system of the considered de-propanizer unit depicted in Fig. 1 was modelled in UniSim Design to obtain a precise model describing of the complex behaviour, see Fig. 2. The scheme shows the distillation column (Fig. 2 (I)) and the main streams. The vapour phase entered the condenser (Fig. 2, (II)), and the liquid phase flew into the reboiler (Fig. 2, (III)). The feed entered into the 5th tray (Fig. 2, (IV)). The outlet streams of the distillation column were distillate (Fig. 2, (V)), and bottom product (Fig. 2, (VII)). The control setup of the distillation column is discussed in detail in Section IV.

TABLE I. Composition of the feed.

| component | mole fraction [–] |
|---|---|
| methane | 0.1955 |
| ethane | 0.1462 |
| propane | 0.1058 |
| i-butane | 0.1076 |
| n-butane | 0.0983 |
| i-pentane | 0.0891 |
| n-pentane | 0.0768 |
| n-hexane | 0.0903 |
| n-heptane | 0.0517 |
| n-octane | 0.0387 |

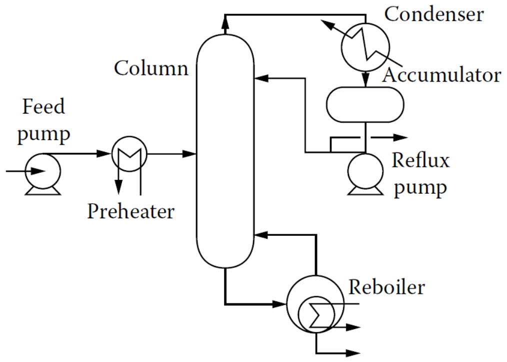


Fig. 1. Scheme of the considered distillation column equipment setup [3, pp. 1821].

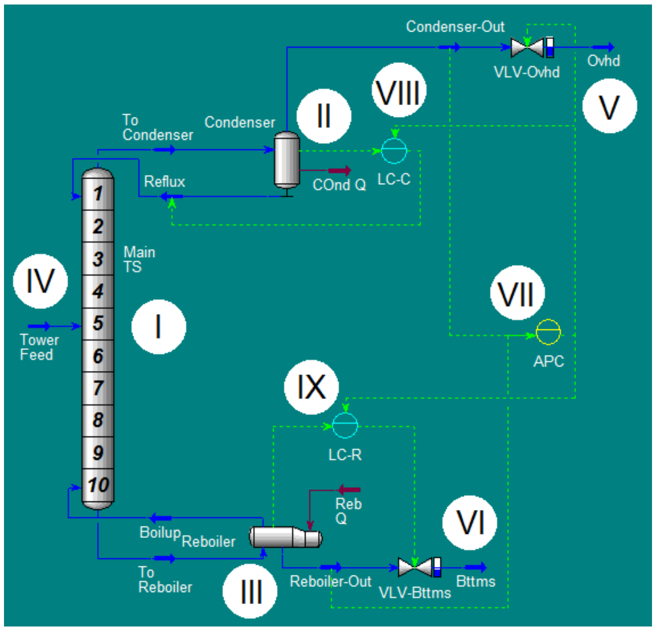


Fig. 2. Scheme of APC of the distillation column in UniSim Design: (I) column, (II) condenser, (III) reboiler, (IV) feed, (V) distillate, (VI) bottom product, (VII) APC controller, (VIII) LC-C controller, and (IX) LC-R controller.

TABLE II. Properties of the streams.

| stream (tray) | $\dot{m}$ [kg/h] | $T$ [°C] | $P$ [kPa] | $x$ [%] |
|---|---|---|---|---|
| feed (5) | 2428 | 32.2 | 1535 | 30 |
| distillate (con.) | 647 | 13.7 | 1468 | 100 |
| bottom (reb.) | 1781 | 134.7 | 1472 | 3 |
| reflux (1) | 489 | 14.5 | 1538 | 100 |
| vapour (1) | 1135 | 35.4 | 1534 | 0 |
| liquid (10) | 4180 | 119.7 | 1543 | 100 |

## III. APC Design

This section introduces the considered configuration of APC design for the distillation column. For APC design two controlled variables (CVs) were considered, i.e., $y_1$ the mole fraction of the i-butane in the distillate, and $y_2$ the mole fraction of propane in the bottom product. Three manipulated variables (MVs) were: $u_1$ the level in the condenser, $u_2$ the level in reboiler, and $u_3$ the valve opening the output flow of the distillate.

### A. Formulation of APC

APC is designed based on a feasible solution of the convex optimization problem in the form of quadratic programming (QP), see e.g. [17], chap. 4.4. In each control step, the following QP is solved:

$$\min \frac{1}{2}\|z\|_P + w^\top z + r, \tag{2a}$$
$$\text{s.t.:}\;\; H_{\text{iq}} z \preceq h_{\text{iq}}, \tag{2b}$$
$$H_{\text{eq}} z = h_{\text{eq}}, \tag{2c}$$

where (2) represents the quadratic cost function to be minimized subject to optimizer $z$. (2b), (2b) respectively are the inequality and equality constraints. Constant matrices $P \succ 0$, $H_{\text{iq}}$, $H_{\text{eq}}$ and vectors $q$, $h_{\text{iq}}$, $h_{\text{eq}}$ have appropriate dimensions. $r$ is a constant.

QP of APC is formulated in a compact form of (2) based on the following control problem. The cost function of APC design is given by

$$\min \sum_{p=0}^{N-1} \left( \|y(k+p|k) - y_{\text{sp}}(k+p|k))\|_Q + \|u(k+p|k) - u_0(k+p|k))\|_R \right), \tag{3}$$

where $Q \in \mathbb{R}^{n_y \times n_y} \succeq 0$, $R \in \mathbb{R}^{n_u \times n_u} \succ 0$ are the weighting matrices of CVs and MVs, respectively. $y_{\text{sp}} \in \mathbb{R}^{n_y}$, $u_0 \in \mathbb{R}^{n_u}$ respectively are the set-point values of CVs and the associated steady-state values of MVs. The linear and constant terms of (2a) were neglected, i.e., $w = r = 0$. The constraints of APC design are considered in the form:

$$y_{\min} \preceq y(k+p|k) \preceq y_{\max}, \tag{4a}$$
$$u_{\min} \preceq u(k+p|k) \preceq u_{\max}, \tag{4b}$$
$$y(k+p+1|k) = f_{i,j}(y(k+p|k), u(k+p|k)), \tag{4c}$$
$$y(k|k) = y(k), \tag{4d}$$

for $\forall p \geq 0$, where $y_{\min}, y_{\max} \in \mathbb{R}^{n_y}$, $u_{\min}, u_{\max} \in \mathbb{R}^{n_u}$ are the limits on the CVs and MVs, respectively. $y(k)$ is the system initial condition, i.e., the measurement of CVs in the $k$-th control step.

### B. Prediction Model of APC Design

Future behavior of the controlled system in MPC can be generally predicted using the set of $(i \times j)$ linear models $f_i, j$. The models $f_i, j$ can be represented in the form of *transfer functions* in the Laplace domain $\mathcal{L}$,

$$G_{i,j}(s) = \frac{Y_j(s)}{U_i(s)} e^{-Ds}, \tag{5}$$

where $G_{i,j}(s)$ is the transfer function from the $i$-th MV to the $j$-th CV, $Y_j(s)$, $U_i(s)$ are real-valued polynomials in $s$ and $D$ is the system time delay.

Considering two CVs and three MVs, we obtained the set of six single-input and single-output (SISO) decoupled models in the form of (5). These models served as the prediction models to design APC by Profit Design Studio environment by Honeywell [18]. They were obtained using an auto-tuning identification tool of Profit Design Studio, and they are given by:

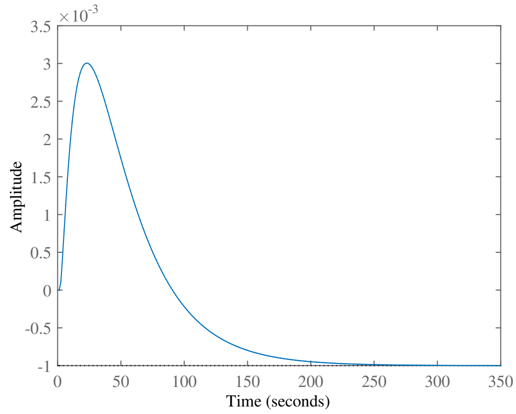


Fig. 3. Step response of $G_{1,1}$.

$$G_{1,1}(s) = \frac{-0.001(-186s^2 - 214s + 1)}{1102s^3 + 618s^2 + 52.7s + 1} e^{-2s}, \tag{6a}$$
$$G_{1,2}(s) = \frac{1.43 \times 10^{-5}(46105s^2 - 324s + 1)}{20536s^3 + 2835s^2 + 184s + 1} e^{-1s}, \tag{6b}$$
$$G_{1,3}(s) = \frac{-9.86 \times 10^{-6}(-2630s^2 - 614s + 1)}{567s^3 + 186s^2 + 24.7s + 1} e^{-1s}, \tag{6c}$$
$$G_{2,1}(s) = \frac{-0.0031(152s^2 + 33.5s + 1)}{5483s^3 + 1242s^2 + 61.9s + 1} e^{-2s}, \tag{6d}$$
$$G_{2,2}(s) = \frac{-1.58 \times 10^{-6}(22597s^2 + 650s + 1)}{5593s^3 + 1664s^2 + 61.6s + 1} e^{-1s}, \tag{6e}$$
$$G_{2,3}(s) = \frac{-6.19 \times 10^{-5}(936s^2 + 70.3s + 1)}{1647s^3 + 983s^2 + 55.2s + 1} e^{-1s}. \tag{6f}$$

The set of prediction models in (6) were used just for the APC design purposes. For the simulation of the closed-loop control performance, the complex model of distillation column was designed in UniSim Design, see Fig. 2. The step-responses of the models in (6) are depicted in Figs. 3–8. As can be seen, the ideally decoupled system was stable, but some of the SISO systems showed periodic or non-minimum phase behaviour.

## IV. Results and Discussion

APC was designed in Profit Design Studio and implemented in UniSim Design environment using a block `Profit Controller`. The following setup for APC design was used: the boundaries on CVs and MVs were set:

$$0\,\% \leq y_i(k) \leq 100\,\%, \tag{7a}$$
$$0\,\% \leq u_j(k) \leq 100\,\%, \tag{7b}$$

for $i = 1, 2$, $j = 1, 2, 3$. Set-points and corresponding steady-state MVs values were:

$$y_{\text{sp}} = [0.0195, 0.0641]^\top, \tag{8a}$$
$$u_0 = [76.8, 47.0, 30.3]^\top. \tag{8b}$$

The square diagonal matrices in the cost function (3) were intensively tuned to ensure the required control performance,

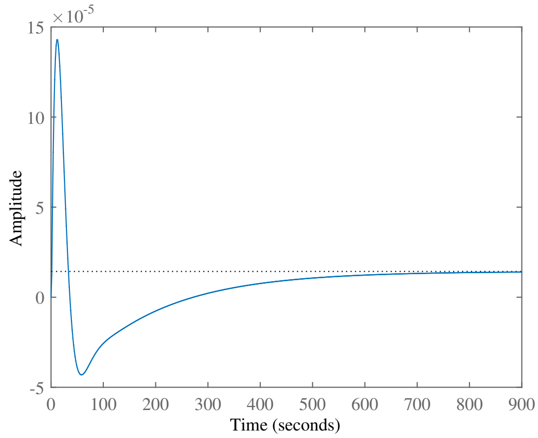


Fig. 4. Step response of $G_{1,2}$.

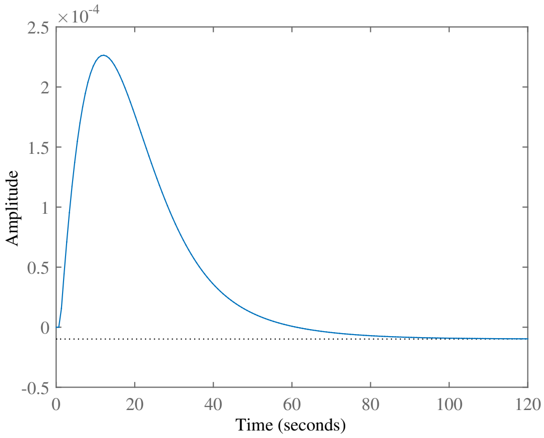


Fig. 5. Step response of $G_{1,3}$.

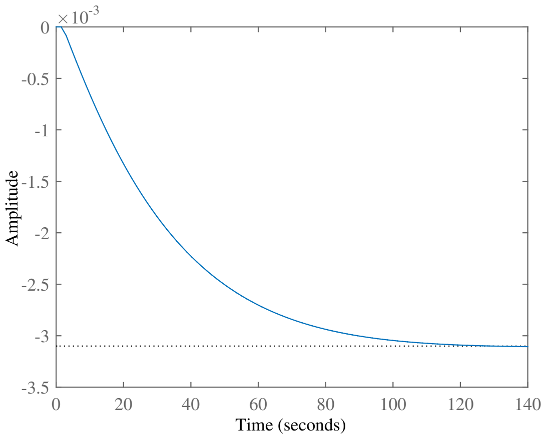


Fig. 6. Step response of $G_{2,1}$.

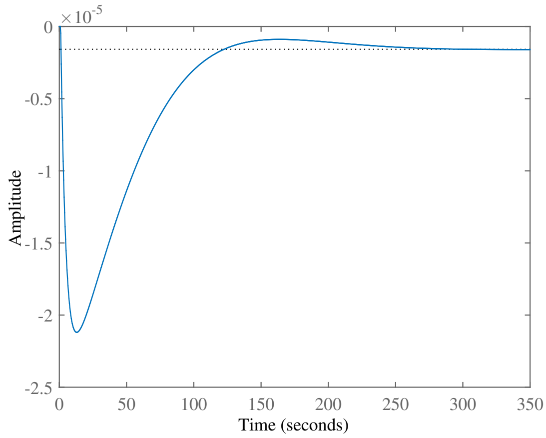


Fig. 7. Step response of $G_{2,2}$.

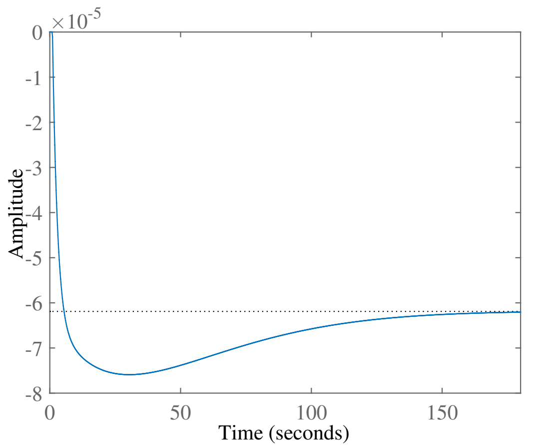


Fig. 8. Step response of $G_{2,3}$.

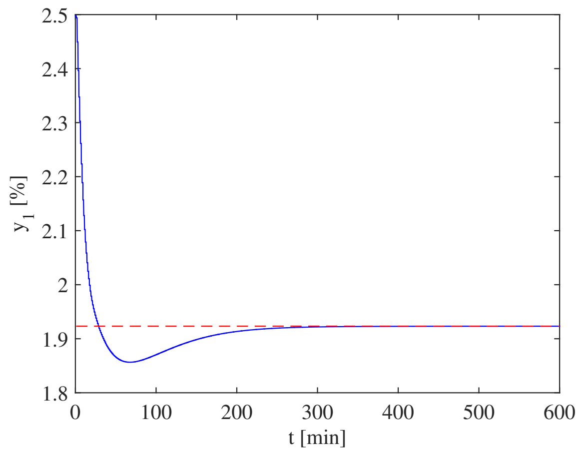


Fig. 9. Control trajectory of $y_1$ ensured by APC.

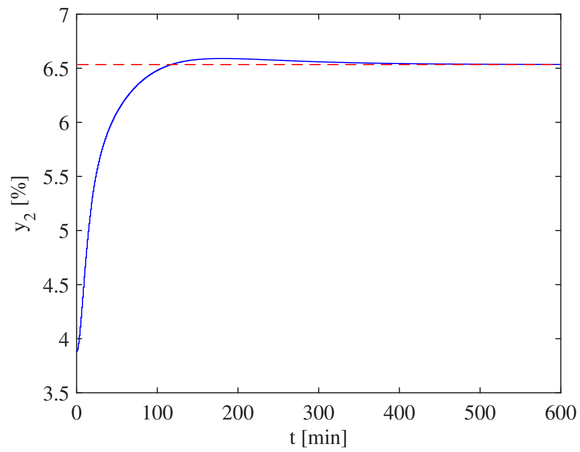


Fig. 10. Control trajectory of $y_2$ ensured by APC.

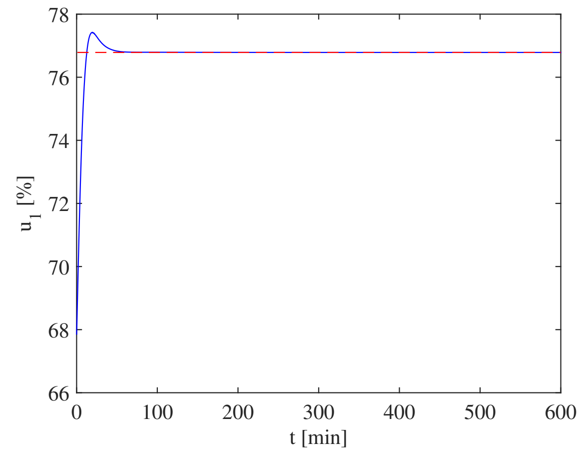


Fig. 11. Output of the APC controller $u_1$

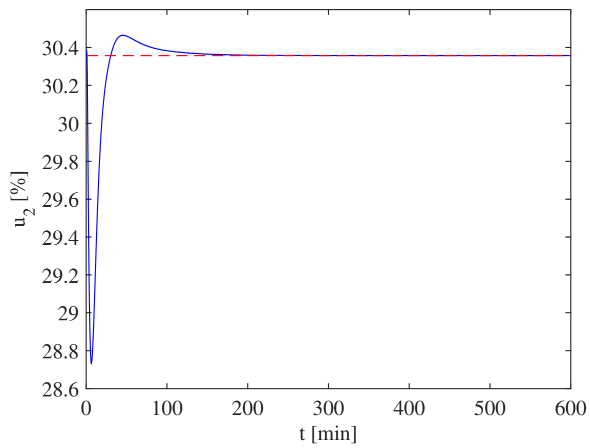


Fig. 12. Output of the APC controller $u_2$

and their values were:

$$Q = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}, \quad R = \begin{bmatrix} 1 & 0 & 0 \\ 0 & 100 & 0 \\ 0 & 0 & 10 \end{bmatrix}. \tag{9}$$

The closed-loop behaviour of the distillation column was simulated using `Peng-Robinson Fluid Package`. The simulation scheme build-up in UniSim Design is shown in Fig. 2. The scheme illustrates the contolled process described in Section IV. The APC control scheme included an supervising APC controller (Fig. 2, (VII)) that supervised two PID controllers, LC-C for level control in the condenser (Fig. 2, (VIII)), and LC-R for level control in the reboiler (Fig. 2, (IX)).

The APC of $y_1$ and $y_2$ was implemented using a cascade control setup, i.e., the set-points of two *slave* PID controllers were computed by an *master* APC controller. The manipulated variable of APC controller $u_1$ represented the set-point for the PID controller LC-C that controlled the level in the condenser by reflux flow/rate. The manipulated variable of APC controller $u_2$ was the set-point for the PID controller LC-R that controlled the level in reboiler by handling the valve opening for bottom/product flow. The third MV from APC controller $u_3$ controlled the distillate output flow by valve opening, see Fig. 2.

The parameters of the PID controllers were intensively tuned subject to periodic and non-minimum-phase behaviour systems in (6) to ensure the required control performance. The considered PID controllers had the form:

$$G_{\mathrm{LC-C}}(s) = \frac{-2.00s + 0.20}{s}, \tag{10a}$$

$$G_{\mathrm{LC-R}}(s) = \frac{-3.30s + 0.33}{s}. \tag{10b}$$

The aim of APC was the set-point tracking and assuring maximal mole fractions of i-butane in the distillate and propane in the bottom product according to the conditions $y_1 \leq 0.2\,\%$ and $y_2 \leq 7.0\,\%$, respectively.

The simulation results of the closed-loop control performances for the APC controller are depicted in Figs. 9–12. Figs. 9, 10 show the control trajectories of $y_1$, $y_2$ ensured by the designed APC, respectively. Fig. 9 shows the control trajectory of $y_1$ of the designed APC. As can be seen, the undershoot occurred, but the steady-state error was removed. Analogous, Fig. 10 depicts the control performance of CV $y_2$. The control trajectory indicated just slight overshoot. On the other hand, $y_2$ performed slower dynamics compared to $y_1$. Figs. 11, 12 show the trajectories of the manipulated variables $u_1$, $u_2$ calculated by the APC controller. The trajectory of $u_3$ is omitted as it had the constant value, i.e. the flow rate of the distillate was constant.

The control performance of APC control was judged using various quality criteria, see Tab. III, where $t_{\mathrm{set}}$ represents the settling time for the considered 0.5 %-neighbourhood of the set-point value. ISE is given by

$$\mathrm{ISE}_i = \int_0^{600} (y_i - y_{\mathrm{sp},i})^2 \mathrm{d}t \approx \sum_{k=0}^{600} (y_i - y_{\mathrm{sp},i})^2, \tag{11}$$

TABLE III. QUALITY CRITERIA OF DE-PROPANIZER CONTROL.

| variable | $t_{\mathrm{set}}$ [min] | ISE [–] | $\sigma_{\max}$ [%] |
|---|---|---|---|
| $y_1$ | 27.3 | 0.01 | 11.6 |
| $y_2$ | 273.3 | 0.62 | 2.1 |
| $u_1$ | 28.2 | 244.60 | 7.1 |
| $u_2$ | 302.7 | 22.68 | 6.3 |

Analogous, ISEs of MVs were evaluated for $(u_i - u_{0,i})^2$. In Tab. III, $\sigma_{\max}$ stands for the maximal overshoot/undershoot

$$\sigma_{\max} = \frac{\max(|y_i|) - y_i(600)}{y_i(600) - y_i(0)} \times 100\,\%, \tag{12}$$

where $\sigma_{\max}$ was analogous evaluated also for MVs. We recall, that the quality criteria depended not only on the APC setup, but are influenced on the tuned PID controllers. The total value of the quadratic quality criterion (3) evaluated for the simulation of the closed-loop control was $J = 2\,513.1$.

## V. CONCLUSION

This paper presents the successful implementation of APC for of the distillation column. The APC controller was designed using Profit Design Studio software, and the closed-loop control performance was evaluated using UniSim Design environment. The complex model of the distillation column was handled as multiple-inputs and multiple-outputs system. APC was implemented to optimize the control performance of the de-propanizer unit. The application of APC controller ensured good the control performance criteria. The designed APC and the tuned PID controllers ensured the offset-free control performance and satisfied the requirements on the upper limit of CVs. The next research will be focused on the implementation of APC on the laboratory distillation column UOP3CC using UniSim Design via OPC server.

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