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PROJECT TITLE:

Influence of feed rate and feed composition on a temperature controller in a binary distillation column

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By

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Abstract

Tuning of a temperature controller in a distillation column is very dependent on feed rate, feed composition and in some cases composition of the products of the column. Tight tuning done at high feed rate and minor content of light component in the feed may result in unstable control at lower feed rate and major content of light component in the feed. This project elaborates how oscillations caused by unstable control could be avoided in a binary distillation column. The proposed methods are smooth tuning of the controller and gain scheduling. The latter was found to be almost inevitable if distillation column has large variation in feed rate. While a smooth global tuning is found to be sufficient enough for variation in feed composition. Regarding the variation of composition in product streams, the gain scheduling was found to be unnecessary if set point in temperature controller is kept at some certain boundaries. MPC models at different operation conditions were also presented in this report.

Acknowledgment

A big gratitude is directed to Sigurd Skogestad for sharing his model of distillation column, but also for continuous help in preface and during the execution of the simulation. Another big acknowledgment is directed to Marius Stre Govatsmark for his encouraging and effective visits at NTNU during the semester and his e-mail support.

Contents

1 Introduction **1**

2 Theoretical background **3**
2.1 Degrees of freedom . . . 3
2.2 Location of the temperature controller . . . 3
2.3 Factors that influence the dynamic behavior . . . 5
2.3.1 Composition differences between two trays . . . 5
2.3.2 Tray holdup . . . 5
2.3.3 Delay in the reboiler . . . 6
2.4 Tuning of the temperature controller . . . 7
2.5 Correlation between operation point and tuning . . . 8
2.6 MPC . . . 8

3 Modeling and implementation **9**
3.1 Model modification . . . 9
3.2 Operation points . . . 10

4 Results **12**
4.1 Tuning results . . . 12
4.2 Simulation results . . . 14
4.3 MPC models . . . 18

5 Discussion **20**

6 Conclusion **23**

Appendices **25**

A Simulation **25**
A.1 Operation point 1 . . . 26
A.2 Operation point 2 . . . 27
A.3 Operation point 3 . . . 28
A.4 Operation point 4a . . . 29
A.5 Operation point 4b . . . 30
A.6 Operation point 4c . . . 31
A.7 Operation point 5 . . . 32

A.8 Operation point 6 33
A.9 Operation point 7 34
A.10 Operation point 8 35

B Models for MPC **36**
B.1 Operation point 1 36
B.2 Operation point 3 37
B.3 Operation point 4b 38
B.4 Operation point 4c 39
B.5 Operation point 5 40
B.6 Operation point 8 41

C Gain scheduling **42**

D Matlab code **44**

E Simulink model **54**

F Risk assessment **55**

1 Introduction

There are different methods that are used to separate two or more components from each other. The one that is used widely in industry is called distillation which is based on differences in volatilities of the components. A standard distillation column has three functional units; cooler, condenser and reboiler. These units give degrees of freedom that is used to define control structure of the column.

The feed that is entering the column can generally be classified by several variables. In this report these variables are reduces to: the rate, the composition and the fraction of liquid. These values contributes to the definition of the operation condition that column is currently operated on. Different control structures are used to stabilize and operate the column. In a real process plant the operation condition may have large variance putting the controllers to test continuously. A big deviation from normal operation condition, the one controller was originally designed for, may result in totally different behavior at other operation conditions. An example of this is often seen when controllers that are tuned "tight" on high feed rate as used at much lower feed rates. The observations in such cases are too aggressive control and even undesired oscillations, making the term "stabilizing control" inappropriate.

Such aggressive behavior is sometimes observed in Kårstø, one of the Statoils gas processing plants. The motivation of this project is to find a reason for such behavior and propose a solution to the problem.

Sigurd Skogestad has introduced several tuning method for smooth tuning of the controllers making the controllers less sensitive to changes in operation condition (Skogestad & Grimholt 2011). Another alternative for avoiding the unwanted aggressive behavior of the controller is to consider adaptive control in the name of gain scheduling (Jang, Annaswamy & Lavretsky 2008). However, gain scheduling requires good understanding of how operation condition, including other variables like pressure, temperature, composition etc., influence the operation of a particular column. The use of gain scheduling of full extent could therefore be unnecessary complicated.

All of the difficulties, with adaptive control, mentioned so far are still worth struggle for, if "tight" control of the column is required. However, a good alternative could in some cases be some kind of compromise between smooth tuning and adaptive control. For instance, the use of smooth control and simultaneous simple model of gain scheduling, by including only feed variation.

For illustration of importance of such solution, stabilizing temperature controller in a single binary column will be considered. First, optimal location of the controller will be discussed. Then this paper will investigate how factors like feed rate, feed composition and also composition of products influence the dynamics in a distillation column and accordingly change the PID tuning parameters like gain (K_c) and time constant (τ_I). Gain scheduling for some operation conditions will then be proposed. In the end models for supervisory layer will be discussed.

The model that is used for this purpose is called "column A", developed by Sigurd Skogestad. The nonlinear dynamic model is based on following assumptions; binary mixture, constant pressure, constant relative volatility, equilibrium on all stages, total condenser, constant molar flows, no vapor holdup, linearized liquid dynamics, but with included effect of vapor flow ("K2"-effect).

2 Theoretical background

2.1 Degrees of freedom

A systematic illustration of the column that is used in this report is presented in figure 2.1. With given feed, cooler to control the pressure and two level controllers there are two remaining degrees of freedom (DOF), that is reboiler duty and reflux ratio. In this report reflux ratio is set to be constant, while temperature controller (TC) is located at the lower part of the column as shown in figure 2.1.

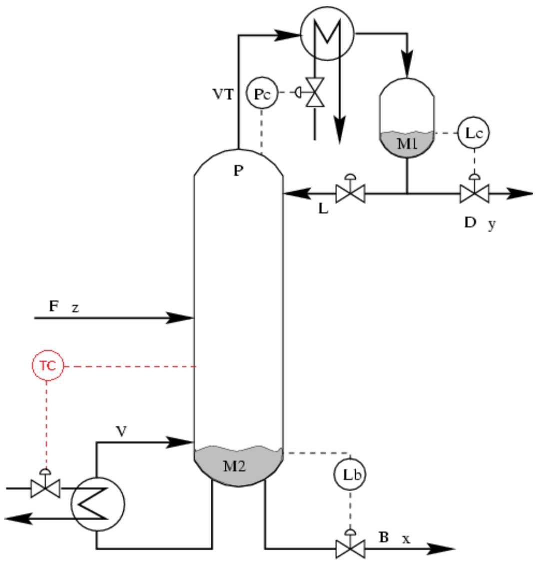


Figure 1: Systematic illustration of a simple distillation column, supplemented with a TC and constants reflux. (Jacobsen & Skogestad 1993)

2.2 Location of the temperature controller

The purpose of a temperature controller is to hold the temperature constant in one specific tray in a column. This can be done with either reflux or reboiler as manipulated variables, controlling the upper and lower part of the column respectively. A good choice of the tray provides more stable operation with disturbance in the feed. Furthermore, resulting in more or less stable production of the bottom and the top products of the column. Sigurd Skogestad has proposed interesting principles of how TC should be selected (Skogestad 2007) (Hori & Skogestad 2007). One of these papers (Hori &

Skogestad 2007) presents examples of structures that are "reasonable" for different type of columns. The paper considers three types of disturbances: feed rate ($\Delta F = \pm 20\%$), feed composition ($\Delta z_F = \pm 10\%$) and fraction of liquid in the feed ($\Delta q_F = \pm 10\%$). However, both Skogestad and Luyben (Luyben 2005) point out that the main disturbances that should be taken into condsideration are disturbances in the feed composition.

In addition, to find the best location of stabilizing temperature controller for all of three types of disturbance, Skogestad and Hori (Hori & Skogestad 2007) propose some rules for finding best location of temperature controller:

1. Steepest slope in temperature profile
2. Small optimal variation with respect to disturbances
3. Large sensitivity to input change

Both the first and the last rule give the same result and usually favor locations away from the column ends, where the temperature slope is largest, while the second rules favors location close to the column ends. One of way for calculating this is empirical method called "max gain rule". However, the method is not exact. Another method that is doubtless exact is called "exact local method" and was presented by Halvorsen (Halvorsen, Skogestad, Morud & Alstad 2003). The objective of the method is to find the worst-case steady state composition deviation. This is given in equation 1, where M_d and M_n includes disturbance and implementation errors.

$$\Delta X = \frac{\bar{\sigma}([M_d \;\; M_n])^2}{2} \tag{1}$$

In this report only single temperature controller together with reboiler is considered

Taken into consideration column A with temperature on the lower part of the column as the first controlled variable and top composition as the second controlled variable, the best location is on 70% from the bottom of the column (Hori & Skogestad 2007). The result conforms with the first rule for selecting location of TC (Luyben 2005). Which is only considering slope in the temperaturprofile. The location correspond to 15^{th} stage in column A and is chosen to be the best even though the remaining degree (reflux) is keeped unused.

2.3 Factors that influence the dynamic behavior

Several changes can be observed in the column when it is operated with different feed rates and feed compositions than it was originally designed for. The most important changes are that also influence the tuning of the controller:

- Stationary effect caused by changes in composition differences between two trays
- Dynamic effect on tray holdup (M_i).
- Time delay (Θ) in reboiler.

These will be more closely investigated in this report.

2.3.1 Composition differences between two trays

At each tray of the column there is certain composition of two product. The difference in composition between two trays is one of the factors that determines the behavior of the column. The effect can be written as shown in equation 2:

$$d\Delta x_i/dt = x_{i+1} - x_i \tag{2}$$

2.3.2 Tray holdup

Tray holdup (M) is amount of fluid at each tray in the column which can be divided into liquid on the sieve tray and downcomer (Wittgens & Skogestad 2000). The holdup is a function of feed rate and feed composition, which again are function of reflux (L). The correlation between M and L can be described by simplified Francis weir formula (Skogestad & Morari 1988):

$$M_{0i} = k_1 L_i^{2/3} \tag{3}$$

Making the assumption that holdup is the same at each tray and $k_1 = k_2 = k_i$, the ratio of two independent operation points yields:

$$\frac{M_0}{M_1} = \frac{k_1}{k_2}(\frac{L_0}{L_1})^{2/3} = (\frac{L_0}{L_1})^{2/3} \tag{4}$$

Rearranging equation 4 and making another approximation that reflux (L) and reboiler duty (V) are equally depended on feed rate (F), yields in equation 5:

$$M_1 = M_0(\frac{V_1}{V_0})^{2/3} \tag{5}$$

In the article "Evaluation of Dynamic Models of Distillation Columns with Emphasis on the Initial Response" (Wittgens & Skogestad 2000) Bernd Wittgens and Sigurd Skogestad looked at tray holdup in more detail. The idea is to split the holdup in several sections where liquid can be found, i.e. sieve tray, downcomer, inlet and outlet weir. As in the previous deduction, the basis of varying height over weir was taken in Francis weir formula, but in this case the formula was modified. The result is presented in equation 6.

$$h_{ow} = 44300 \cdot (\frac{L_{out}}{\rho_l \cdot 0.5 \cdot d_{weir}})^{0.704} \tag{6}$$

However, taking a ratio between two independent operation points of the columns yields the same equation 5 with exponential factor 0.704 instead of 2/3.

2.3.3 Delay in the reboiler

The saturated steam that is entering the reboiler goes through several time consuming stages. First of all, steam condenses in the shell side of the reboiler, then the heat is transfered through the wall, and eventually the fluid in the tube side of the reboiler is warmed up. Numerous bubbles that are formed generas the driving force in the column. The described dynamics are often neglected. The reason for that is that the time scale is often much smaller compared to other dynamics in the column. Instead delay (Θ), given by equation 7, with varying value at different operation point, is often included.

$$\Theta = \frac{M_0}{L} \approx L^{-1} \tag{7}$$

Approximation is made based on the fact that holdup is constant in condenser (M_D)

and reboiler (M_B). The approximation that L and V are equally depended on F from 2.3.2 can also be made here, resulting in equation 8:

$$\Theta_1 = \Theta_0 \cdot \frac{V_0}{V_1} \tag{8}$$

Where V_0 and Θ_0 are initial reboiler duty and corresponding delay, respectively.

2.4 Tuning of the temperature controller

The SIMC tuning rules (Skogestad & Grimholt 2011) are some of the useful rules to implement on PID controllers. In this report, the rules were therefore implemented to tune the temperature controller. Generally, tuning rules can be described as presented in equation 9-10, with controller gain (K_c) and time constant (τ_i) as tuning parameters:

$$K_c = \frac{1}{k} \cdot \frac{\tau_1}{\tau_c + \Theta} = \frac{1}{k'} \cdot \frac{1}{\tau_c + \Theta} \tag{9}$$

$$\tau_I = min\,\{4(\tau_c + \Theta), \tau_1\} \tag{10}$$

Making two assumptions that $\tau_c = c \cdot \Theta$ with $c \geq 1$ and $\tau_c + \Theta << \tau_1$, equation 9 and 10 gives equation 11 and 12 respectively.

$$K_c = \frac{1}{k' \cdot \Theta} \cdot \frac{1}{(c+1)} \tag{11}$$

$$\tau_I = 4\Theta \cdot (c+1) \tag{12}$$

Equation 11 shows that when c is a constant the factor $k' \cdot \Theta$ is one and only that decides K_c. While in equation 12 it is pure delay that decides the time constant for the controller.

2.5 Correlation between operation point and tuning

In equations 11 and 12 it is shown that tuning parameters K_c and τ_I are dependent only on k' and Θ. These values are characteristic for the operation condition that the column is "running on". The latter is elaborated in equation 8. While the first one can be characterized as presented in equation 13:

$$k' = \frac{d\Delta x_i/dt}{\Delta L} = \frac{x_{i+1} - x_i}{M_i} \tag{13}$$

By implementation of Francis weir formula presented in equation 5 and taking into consideration only the largest effect caused by holdup changes, the equation results in:

$$k' \approx M_1^{-1} = M_0^{-1}(\frac{V_1}{V_0})^{-2/3} \tag{14}$$

The non constant values in tuning gain, between current state (i) and reference state (0) can then be presented as follows:

$$k'_i \cdot \Theta_i = M_0^{-1}(\frac{V_i}{V_0})^{-2/3} \cdot \Theta_0 \cdot \frac{V_0}{V_i} = k'' \cdot V_i^{-5/3} \tag{15}$$

With constant term: $k'' = \frac{V_0^{5/3} \cdot \Theta_0}{M_0}$

2.6 MPC

Model Predictive Control (MPC) is a common control unit in supervisory layer. The generated steps in manipulated variables (MV) are often set points in layer below, that could be a PID controller. The timescale of the MPC is consequently bigger then in for layer below it. The advantages of MPC is that is uses a (multivariable) process model to prediction of future behavior, some types of mathematical programming (quadratic programming) for optimization of predicted future performance and can handle several types of constrains. MPC software that Statoil has continuously been using for the last years is called Septic. The software is equipped with complicated solvers and tunings parameters. These would not be discussed in this report.

3 Modeling and implementation

The model that was used is called "column A" and was developed by Sigurd Skogestad. The nonlinear dynamic model is based on following assumptions; binary mixture, constant pressure, constant relative volatility, equilibrium on all stages, total condenser, constant molar flows, no vapor holdup, linearized liquid dynamics, but with included effect of vapor flow ("K2"-effect). The model has 41 stage with feed located on 21^{th} stage, with numeration from the bottom of the column.

Some modification were done to the model. These are presented in the following subsection.

3.1 Model modification

The objective of the presented modifications was to match the model to one specific distillation column at Kårstø, namely "butanesplitter". The column is the last one in series, and can be assumed to have a binary mixture of n-butane and isobutane and the remaining content of methane, ethane, propane and naphtha assumed be isolated in previous columns.

The original version of the model had four unused degrees of freedom. Both distillate (D) and bottom (B) flows were used to control levels in cooler (the top stage of the column) and reboiler (the bottom stage of the column) respectively. Reboiler duty (V) was used to control temperature on 15^th stage of the column. While the last degree of freedom was left unused. This is was done to point the focus on the TC.

However, column A has not neither temperature nor pressure in the calculations. Therefore some approximation were made. First of all it was assumed the column is operated at pressure of 1 bar. Then it was assumed that temperature at each stage (T_i) is only depended on composition on the stage (x_i), as shown in the equation 16.

$$T_i = T_{b,1} \cdot x_i + T_{b,2} \cdot (1 - x_i) = \Delta T \cdot x_i + T_2 \tag{16}$$

Where $T_{b,1}$ and $T_{b,2}$ are boiling temperatures to isobutane and n-butane respectively and ΔT is difference between boiling temperatures of the two components.

3.2 Operation points

The objectives for the project was to observe how tuning is changing with different feed rate, feed composition and desired composition of the products streams. Whereupon observe how temperature controller behaves at these conditions. Based on these condition there were defined 10 operation points. These are presented in table 1. Temperature profiles for each operation point are given in figure 2.

Table 1: Definition of different operation points. Specified values are marked bold.

Operation point	F	L	V	zF	xD	xB	T_{15} $[^oC]$
1	**1**	2.70629	3.20629	**0.5**	**0.01**	**0.01**	-3.94
2	**0.75**	2.02972	2.40472	**0.5**	**0.01**	**0.01**	-3.94
$3^{1)}$	**0.5**	1.35315	1.60315	**0.5**	**0.01**	**0.01**	-3.94
$4a^{2)}$	**0.5**	**2.02972**	2.27874	**0.5**	0.00194	0.00583	**-3.94**
$4b^{2)}$	**0.5**	**2.02972**	2.27756	**0.5**	0.00149	**0.01**	-5.03
$4c^{2)}$	**0.5**	**2.02972**	2.28194	**0.5**	**0.01**	0.00122	-1.64
5	**1**	2.66538	2.96130	**0.3**	**0.01**	**0.01**	-2.61
6	**1**	2.70420	3.10217	**0.4**	**0.01**	**0.01**	-3.23
7	**1**	2.67508	3.27712	**0.6**	**0.01**	**0.01**	-4.75
8	**1**	2.60026	3.30434	**0.7**	**0.01**	**0.01**	-5.64

1) Reflux constraint ignored 2) Reflux constraint active

For all of the operation points the nominal holdup at stages 2-40 was defined by the equation 5. While the delay in the reboiler was set by equation 8. The first three operation points were made with only difference in feed rate. Temperature profiles is for that reason the same for these operation points. In spite of ideology that both reboiler duty and reflux rate is linear dependent on feed rate, this statement will not hold in practice. The reason for that is that such low reflux rate would probably result in weeping in the column. For the other cases with feed low feed rate, the reflux rate was set to 75% of the steade state value with full feed rate.

Operation points 4a-4c have the same feed rate, feed composition and accordingly the same reflux rate (75%). The distinction is made on emphasis of the chose of the remaining controlled variable. In the first case (4a) setpoint of TC was kept the same as in the previous operation points. In the second case (4b) bottom composition was held on the desired value and this way resulting in high profit on the bottom product. While in the last case (4c) it was the destillate. For all of the three cases the steady state conditions was achived with reboiler duty as manipulated variable.

The remaining operation points (5-8) were created with consideration of the feed composition. The compositions of product streams are held at the desired values in all of the cases. The steady state conditions were achieved with both reboiler duty and reflux rate as manipulated variable for the product streams.

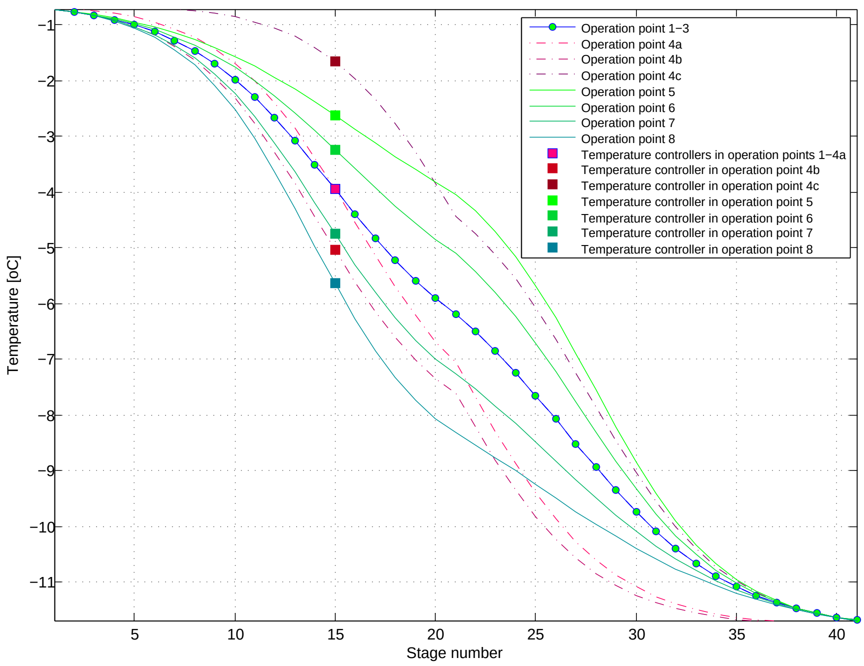


Figure 2: Temperature profiles at different operation points, with bottom to top enumeration.

M-file code "Starter_All.m" can be used to run the simulation at different operation points, both for tuning an open loop, but also to compare the tuning results with disturbances in feed rate and feed composition. The description is added in the begging of the file and is also giving in appendix D. For creation of MPC models the file "starter_All_MPCmodels.m" can be used. Both files runs simulink file "colas_nonlin_operation_All.mdl" which is also given in in appendix E.

4 Results

4.1 Tuning results

Temperature controller was tuned at different operation points. In all cases the tuning was done with equations 9 and 10, assuming pure integrating response and tuning parameter (τ_c) equal to total delay (Θ). The results of tuning are presented in table 2

Table 2: Tuning results with $\tau_c = \Theta$.

	Process		Control			
Operation point	Θ	k'	τ_c	K_c	τ_I	K_c/τ_I
1	2.00	0.9786	2.00	0.2555	16.000	0.015966
2	2.67	1.1781	2.67	0.1592	21.333	0.007460
3	4.00	1.5225	4.00	0.0821	32.000	0.002566
4a	2.81	1.6603	2.81	0.1070	22.513	0.004753
4b	2.82	1.8184	2.82	0.0977	22.524	0.004336
4c	2.81	0.4991	2.81	0.3565	22.481	0.015858
5	2.17	0.5926	2.17	0.3896	17.324	0.022490
6	2.07	0.7806	2.07	0.3099	16.537	0.018738
7	1.96	1.1743	1.96	0.2176	15.654	0.013901
8	1.94	1.3535	1.94	0.1904	15.525	0.012261

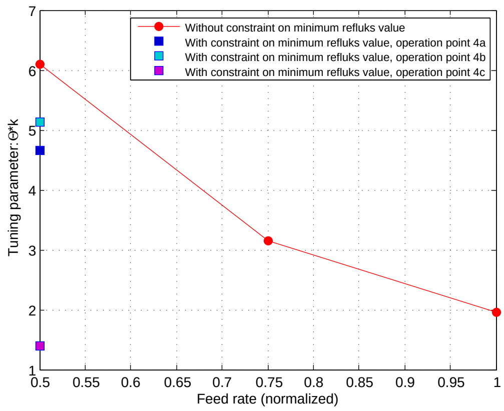
Comparison of calculated tuning parameter and the one observed in during the tuning (process value) is presented in table 3

Table 3: The calculated tuning parameter $k' \cdot \Theta$ in comparison with process value.

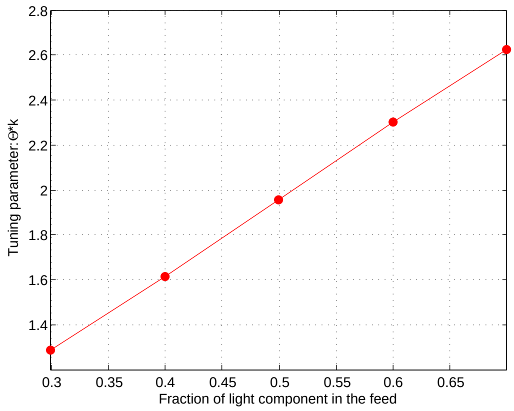
	Process values		Calculated values	
Operation point	$k' \cdot \Theta$	$(k' \cdot \Theta)^{1)}$	$k' \cdot \Theta$	$(k' \cdot \Theta)^{1)}$
1	1.957	1	4	1
2	3.147	1.607	6.461	1.615
3	6.090	3.112	12.699	3.175
4a	4.665	2.384	7.067	1.767
4b	5.128	2.620	7.073	1.768
4c	1.402	0.717	7.051	1.763
5	1.286	0.657	4.567	1.142
6	1.616	0.826	4.226	1.057
7	2.302	1.176	3.857	0.964
8	2.626	1.342	3.804	0.951

1) Scaled by first operation point

For consideration of gain scheduling it was found strong correlation between $k' \cdot \Theta$ and feed rate as well as $k' \cdot \Theta$ and feed composition. This is presented in figure 3.



(a) Results of tuning at different feed rates



(b) Results of tuning at different feed composition

Figure 3: Correlation between $k' \cdot \Theta$ and feed rate a), feed composition b).

As shown in figure 3 and more elaborated in appendix C the tuning parameters are hard dependent on how the distillation column is operated. For instance, impurities in the distillate or bottom product would reduce tuning significant. However, if set point in temperature controller is held constant (operation point 4a), a linear correlation on tunings parameters and feed rate is observed. The regression line is evaluated to be: $y = -5.42x + 7.32$ with gain scheduling factor of -5.42 as sketched in figure 4.

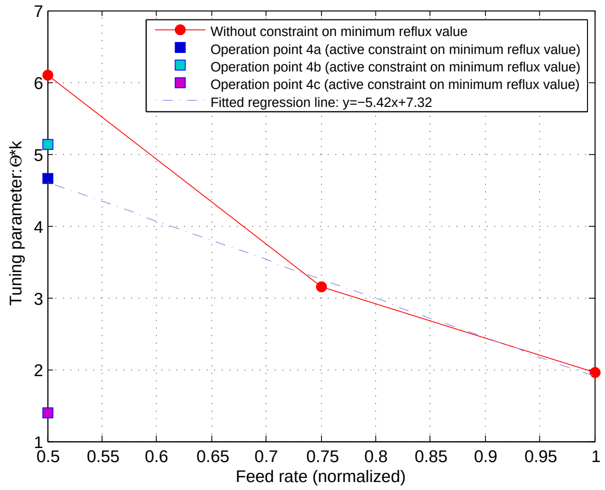
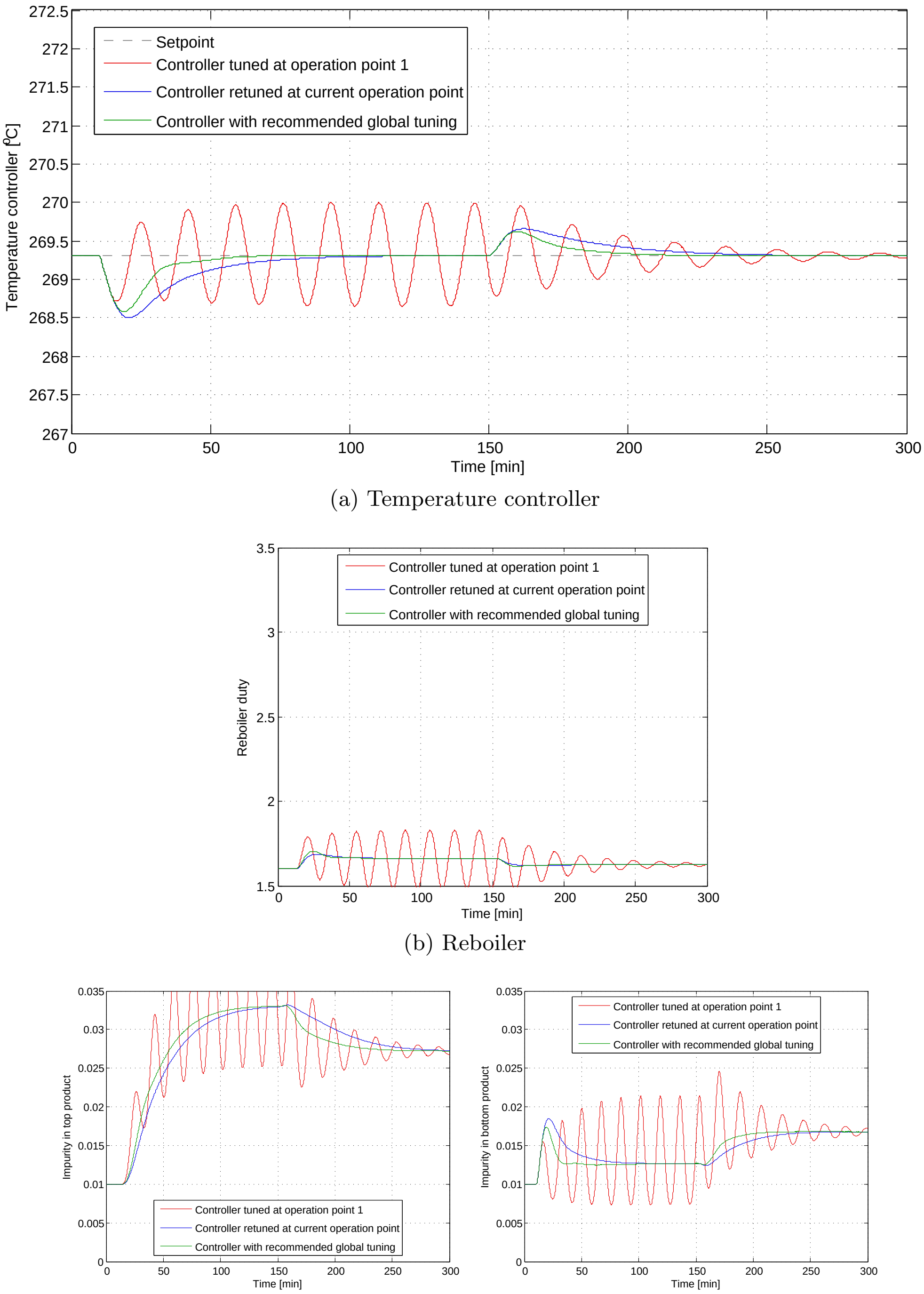


Figure 4: Correlation between $k' \cdot \Theta$ and feed rate. Regression line is fitted to operation points with the same setpoint in TC.

4.2 Simulation results

The obtained tunings, that are showed in table 2 were implemented at different operation points. The idea was to set tuning done at first operation point to the test at other operation points and compare with new tuning at current operation point. A global tuning was also suggested with characteristic smooth behavior.

The testing procedure was set to be disturbance at feed rate after 10 with magnitude of 0.1 and feed composition after 150 minutes with magnitude if 0.05. All of the simulations are presented sequentially in appendix A. In this section operation point 3, 4b, 8 are presented in figures 5- 7 respectively. All of the figure show temperature controller, reboiler duty and impurity in distillate and bottom product with approximately the same window as operator would see it.



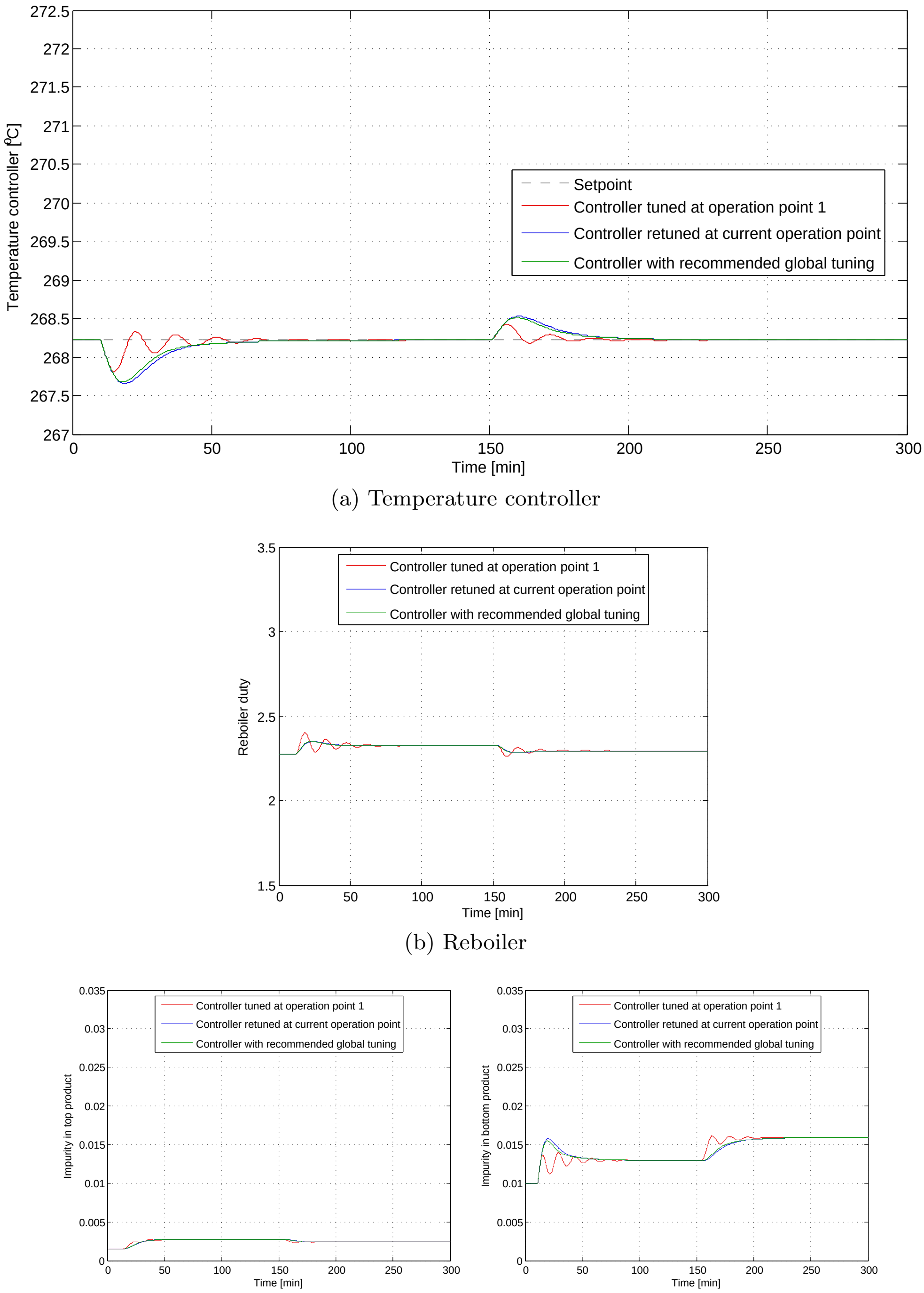
(a) Temperature controller

(b) Reboiler

(c) Impurity in top product

(d) Impurity in bottom product

Figure 5: Behavior of temperature controller a) at operation point 3 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes.



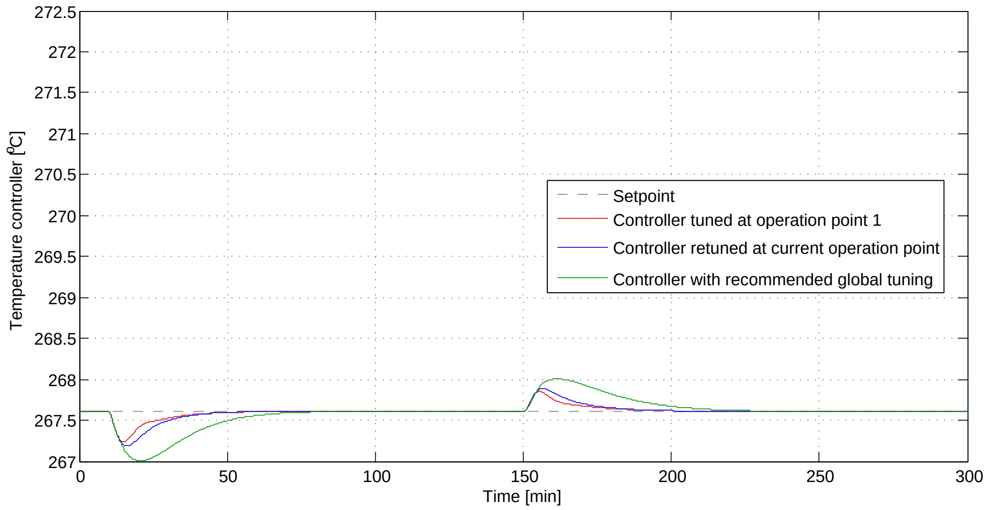
(a) Temperature controller

(b) Reboiler

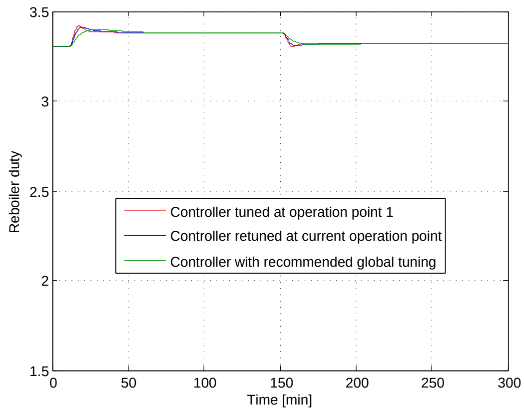
(c) Impurity in top product

(d) Impurity in bottom product

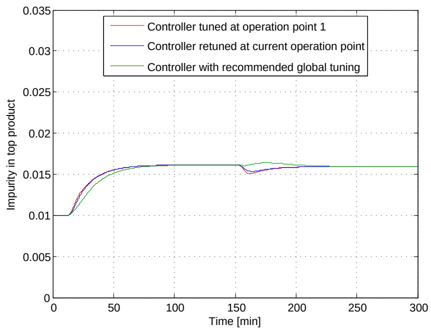
Figure 6: Behavior of temperature controller a) at operation point 4b with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes.



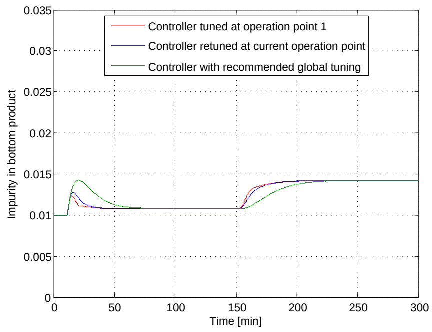
(a) Temperature controller



(b) Reboiler



(c) Impurity in top product



(d) Impurity in bottom product

Figure 7: Behavior of temperature controller a) at operation point 8 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes.

4.3 MPC models

MPC models for some chosen operation points were made. These models could be used in a MPC software, like for instance Septic. The models have not been through any kind of modification for a MPC controller, but were rather made for illustration of how a change in setpoint of temperature controller (MV in MPC) influences impurities in top and bottom stream products (CV in MPC). The range of y-axis in the model is adapted to the area of the model at each case and therefore not fixed. This is done for two reasons. First of all to clearly illustrate the difference between different tuning, but also to make it more realistic to the one Septic would generate. The range of an impact of a single step (0.5^0C) is displayed on bottom right subplot at each figure.

Models for chosen operation points 1, 3, 4b, 4c, 5 and 8 are presented sequentially in appendix B. In this section two models for operation point 4b are presented in figures 8 and 9.

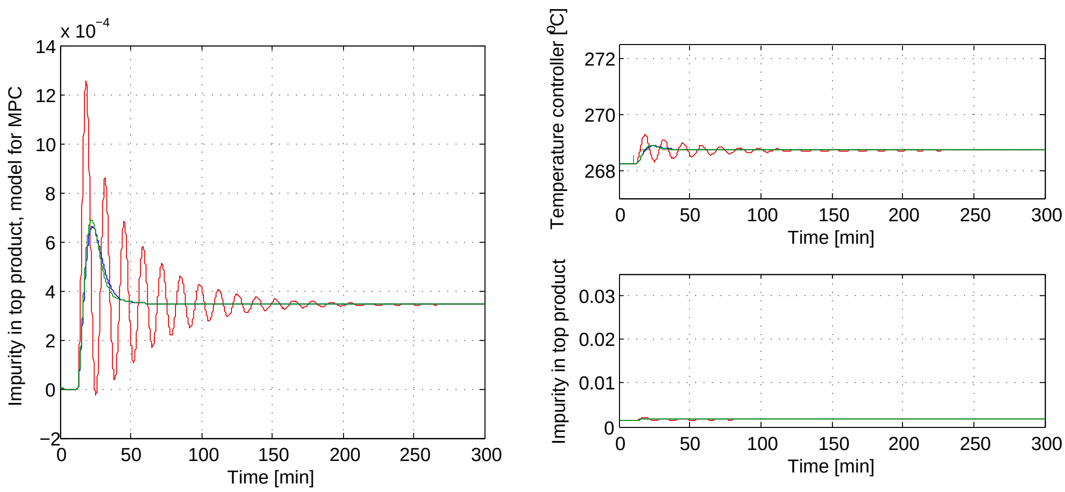


Figure 8: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

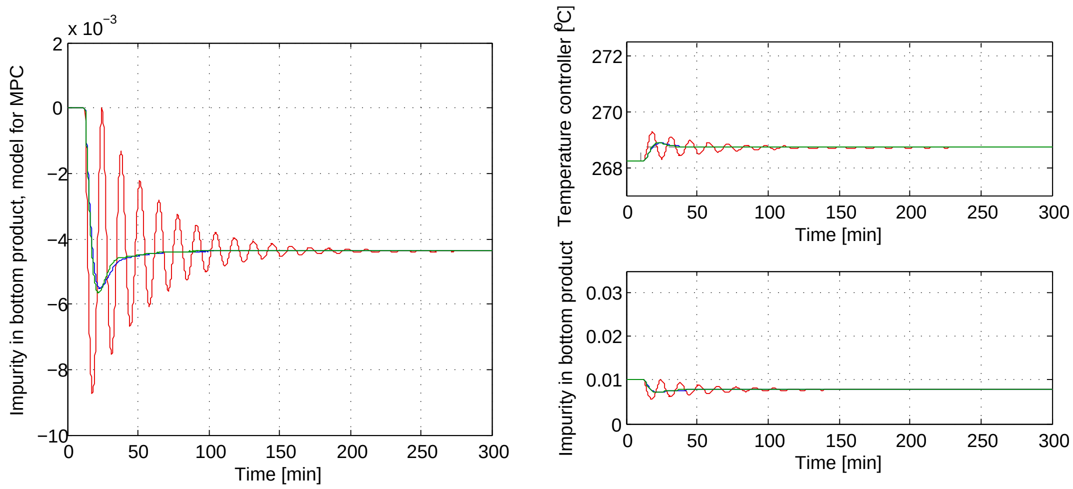


Figure 9: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

5 Discussion

In equation 15 it was found that controller gain is non linear correlated to reboiler duty. In this report the reference values of reboiler duty, delay in the reboiler and liquid tray holdup were used from the first operation point and applied in calculations of $k' \cdot \Theta$ with equation 15. The values were then compared with values from process (table 3).

The results show that all of the approximation done in section 2.3 can be excepted at certain operation conditions. Calculated and process values at for example operation points 1-3, with only differences in feed rate, follows the same trend. However, the trend is not followed at operation points 5-8. The reason for that could be approximation done in equations 13- 14, since feed composition influences composition at each stage, what is clearly observed in temperature profiles in figure 2.

Regarding operation points 4a-4c it is observed that process values changes, but the calculated values remains almost the same, due to the fact that they are dependent only at reboiler duty, which has almost no variation (table 1). The variation of the process values can be explained by investigate these operation points one by one.

When feed rate drops significant, constraint on minimum reflux would be reached (75%) and further operation needs to be specified by which product stream should be given up to control. In that case operator of the column would have three alternatives:

1. Give up composition control in both product streams and rather control the temperature in the column (operation point 4a). In this case both product stream would be overpurified, but temperature would be kept at the same setpoint.

2. Give up composition control in distillate product and rather prioritize control of the bottom stream (operation point 4b). This would result in increase of purification in the distillate stream. The choice could be preferred if the price of the light component is smaller then the heavy component.

3. Give up composition control in bottom product and rather prioritize control of distillate (operation point 4c). This would result in increase of purification in the bottom stream. The choice could be preferred if the price of the heavy component is smaller then the light component.

In the case with binary mixture with n-butane and isobutane. The latter one is expected to be more expensive and should therefore be produced at largest possible scale. If the feed rate is set, the best solution is to keep impurity at highest profitable value and overpurify the other stream. Since isobutane is the light component in the mixture, operation point 4c is the most profitable from these three operations.

From the control point of view operation point 4c is the one with smallest process gain (table 2), but a small increase of reboiler duty has a tremendous effect on top product at this operation point. The explanation for that is lying in the fact that temperature controller at this point is far from nominal value. A small disturbance in the feed would therefore pollute the product stream far over the desired level, making the product hard to sell. However, operation points 4a and 4b does not have this behavior. One suggestion for the best operation point could be a compromise between operation point 4b and 4c. This could be done by adding constrains in the temperature controller. On the other hand, it is important to state the fact that reflux was not used as a manipulated variable in simulations with disturbances and problem with the tremendous effect on the top product would be by far from so tremendous if reflux would be used to control composition in distillate by for example MPC, which can handles reflux constraint in a proper way.

Nevertheless if the prices of products are expected to be varying and changes of active constrains are demanded. The set point of TC i forced to be changing. An option in such case could be an individual gain schedule. However, this can be avoided if setpoint of TC is designed to have only small variations, i.e. add constraints as explained above. Even though this will not result in maximum utilization of the valuable product, the control structure would be much simpler.

Regarding simulation results, it was obvious that tuning done at first operation point gave oscillatory behavior at lower feed rates. A global tuning was therefore suggested. The tuning had smooth behavior in all of the operation points, but could be considered too slow at high feed rate and high content of light component in the feed. Consequently, a linear gain scheduling with consideration of feed rate was suggested. The factor was found to be 5.42 and was based on the fact that temperature controller would be kept at the same value at different feed rates. Even though in theory it was found non linear relation between reboiler duty and tuning parameters, the empirically obtained linear simplification is suitable when constrain on minimum reflux is taken into consideration.

Gain scheduling with consideration of feed composition was not evaluated. On of the reasons for that is that tuning from the first operation point was not found to have oscillatory behavior at these operation points. Recommendation for further work could be to see if this is the case with other models then column A.

Later in the report there were found MPC models for some interesting operation point. Some strong oscillations were noticed with controller tuning from the first operation point. These were observed, as expected, in operation points 3 and 4b. Even Septic, which is a good MPC software would have problems controlling a process with these (red) models. Regarding the global tuning, in most operations it had a similar model as retuned controller and Septic would probably not have problems.

The alarming fact with developed MPC models appeared when gain of one model developed at one operation point was mutually compared with another model developed at different operation point. Put another way, if models that were developed at one operation points would be used at other feed rates and feed compositions, the MPC would probably have total different behavior. The vast variation of the gain is observed between operation points 3 and 4c regarding the models for distillate and operation points 4b and 5 regarding the models for bottom product. The variation can be summarized with following enumeration:

- If the CV is initially overpurified, the models would have small gain.
- If the CV is kept at the constrain, but the other product stream is overpurified, the models would have abundant large gain.
- An increase of feed rate with same composition in product streams and set point for TC does not result in significant change of the model.
- An increase of light component content in the feed stream results in gain reduction of the model. Equivalently, low content of light component results in gain increase of the model.

The models were made for illustration of how a change in setpoint of temperature controller in influences impurities distillate and bottom stream products. The models have not been through any kind of modification for a MPC controller, and should not be implemented directly in a MPC software. Recommendation for further work could be to evaluate the behavior of MPC at different operation points and different MPC models.

6 Conclusion

In this project behavior of a single temperature controller in a distillation was investigated. Column A was used as a model for distillation column with a simple relationship between composition and temperature at each stage. Several operation points were defined, with variation in feed rate, feed composition and composition in product streams. Tuning of the temperature controller was done at initial operation conditions and evaluated at other operation points. The result was that initial tunings had oscillatory characteristics when the feed rate was reduced significantly. It was therefore suggested linear gain scheduling with magnitude of 5.42. It was also suggested that global tuning for the controller could be used at various feed compositions. Regarding variation of composition in product streams, the tuning at current feed rate was found to be suitable if set point in temperature controller is kept at some certain boundaries. Considering supervisory control layer, it as found that the developed models at certain operation conditions should be evaluated before implementation.

Trondheim 16.desember 2011

Iakov Dolgov

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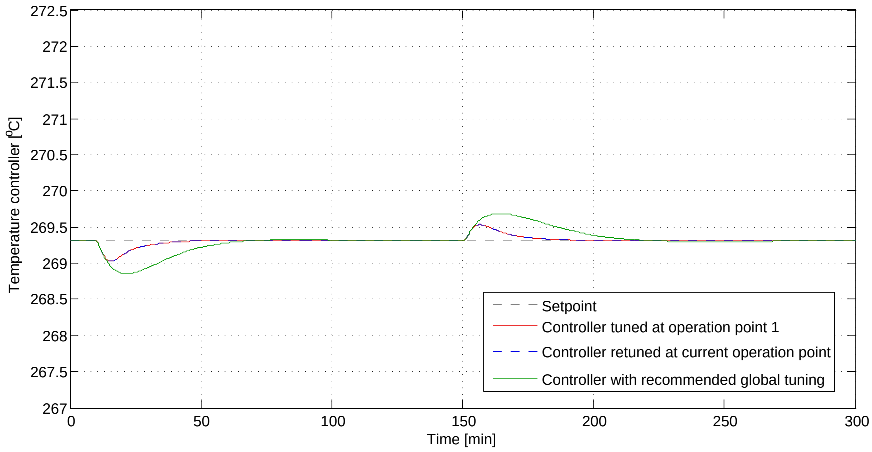
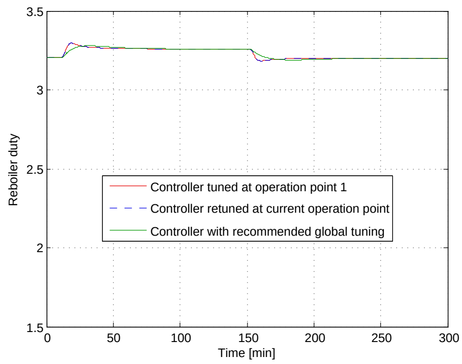
Appendices

A Simulation

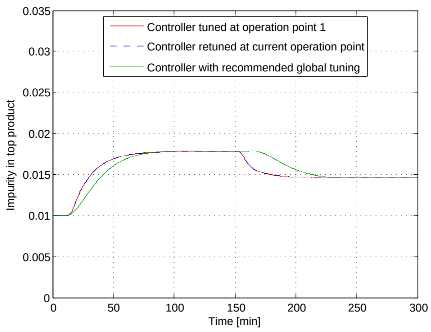
A.1 Operation point 1



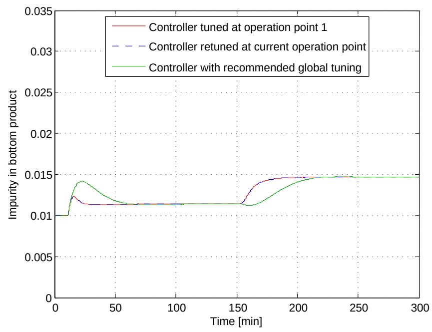
(a) Temperature controller



(b) Reboiler



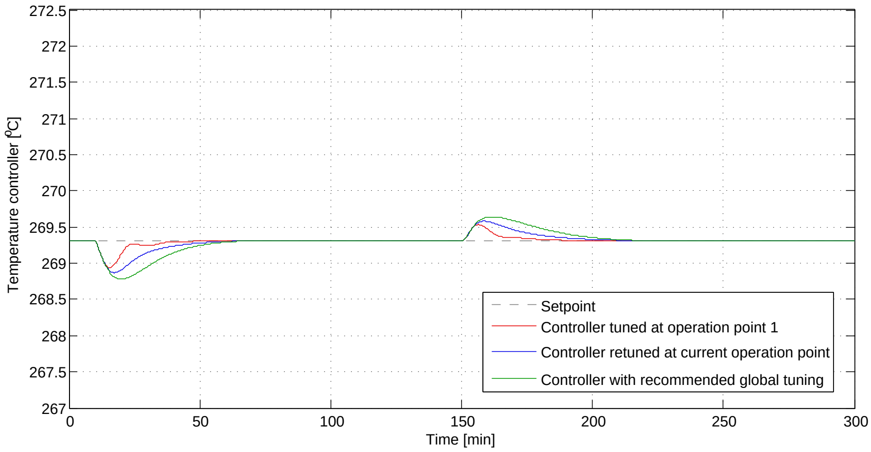
(c) Impurity in top product



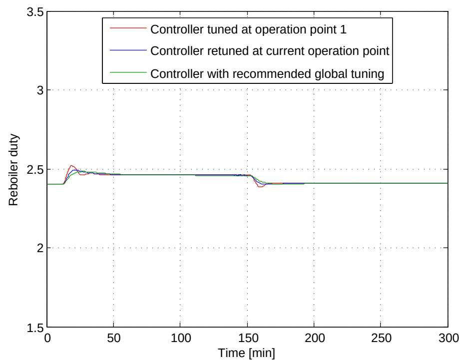
(d) Impurity in bottom product

Figure 10: Behavior of temperature controller a) at operation point 1 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

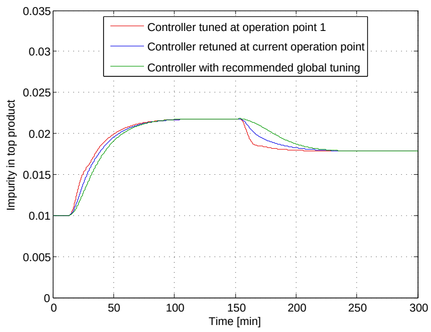
A.2 Operation point 2



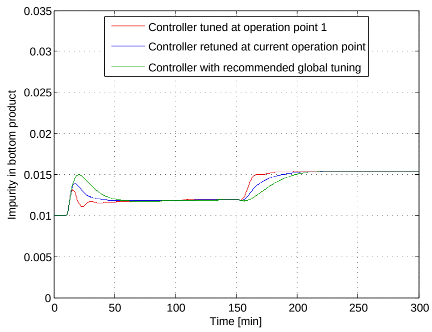
(a) Temperature controller



(b) Reboiler



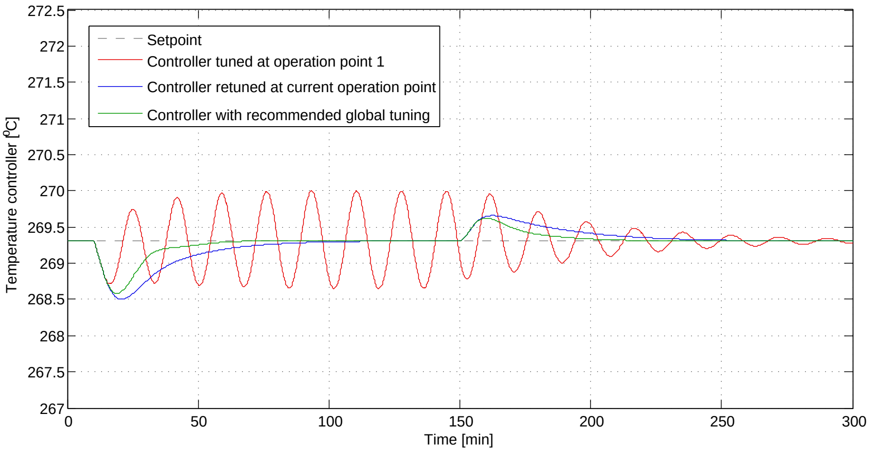
(c) Impurity in top product



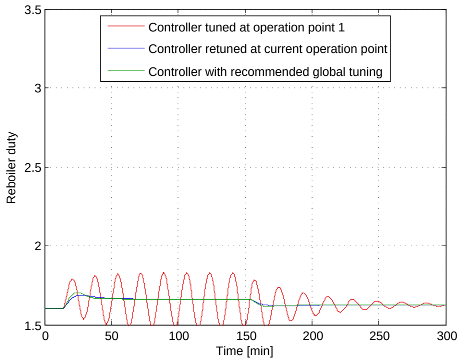
(d) Impurity in bottom product

Figure 11: Behavior of temperature controller a) at operation point 2 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

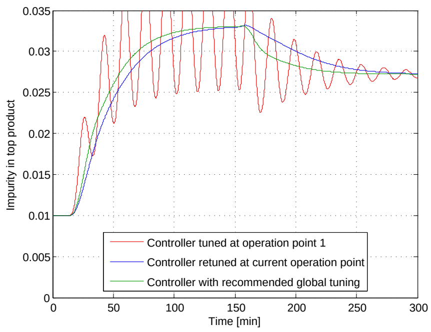
A.3 Operation point 3



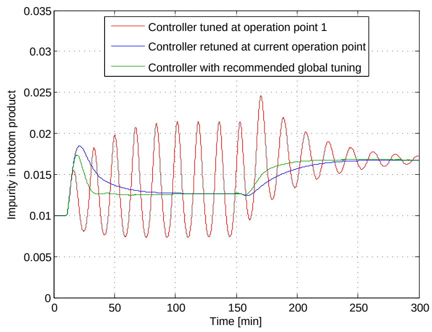
(a) Temperature controller



(b) Reboiler



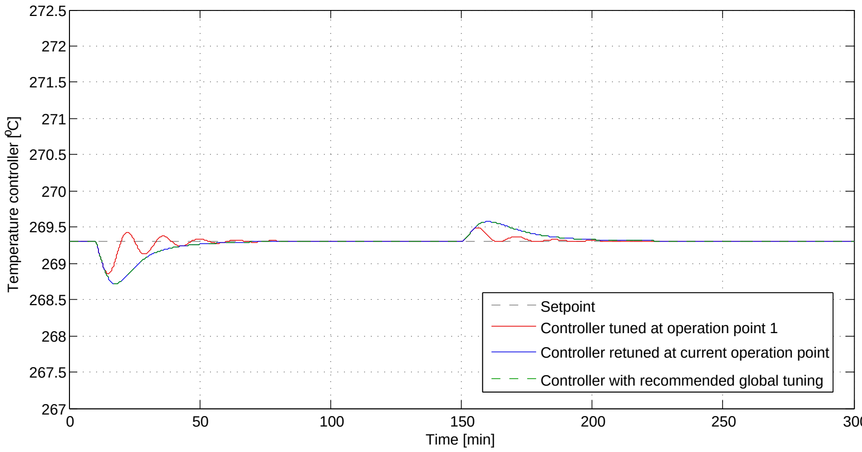
(c) Impurity in top product



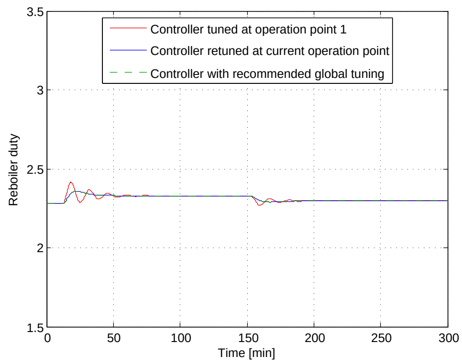
(d) Impurity in bottom product

Figure 12: Behavior of temperature controller a) at operation point 3 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

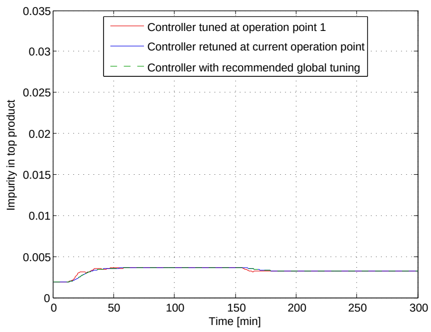
A.4 Operation point 4a

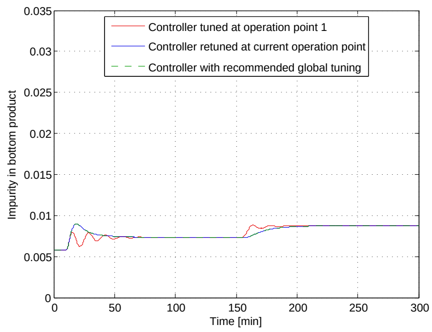


(a) Temperature controller



(b) Reboiler

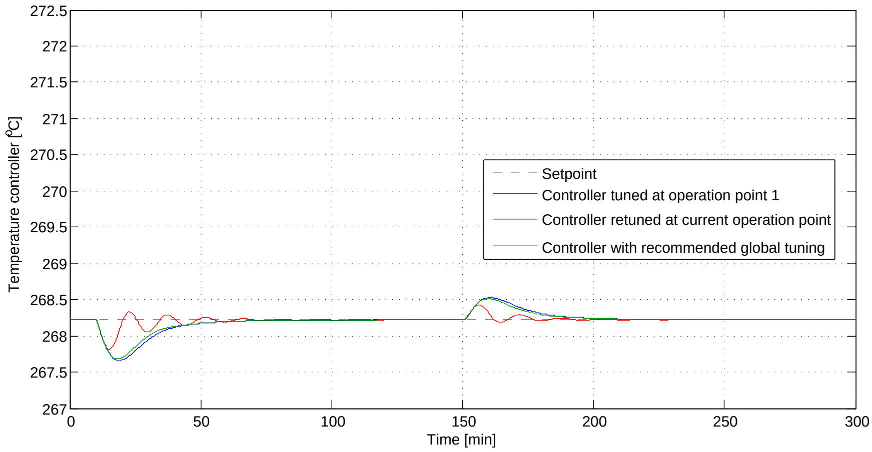




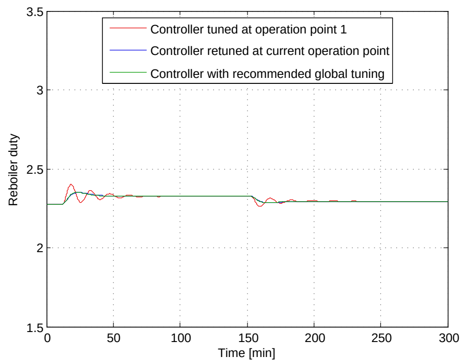
(c) Impurity in top product (d) Impurity in bottom product

Figure 13: Behavior of temperature controller a) at operation point 4a with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

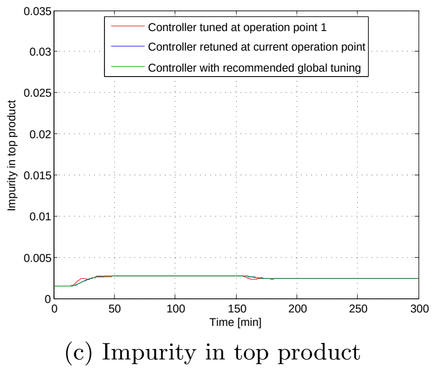
A.5 Operation point 4b



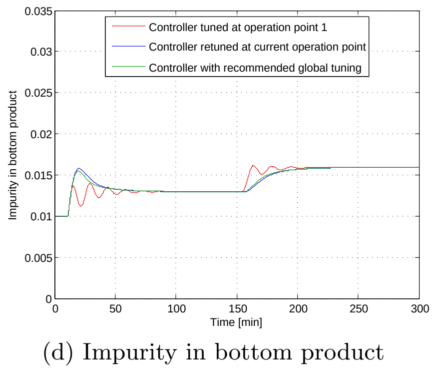
(a) Temperature controller



(b) Reboiler



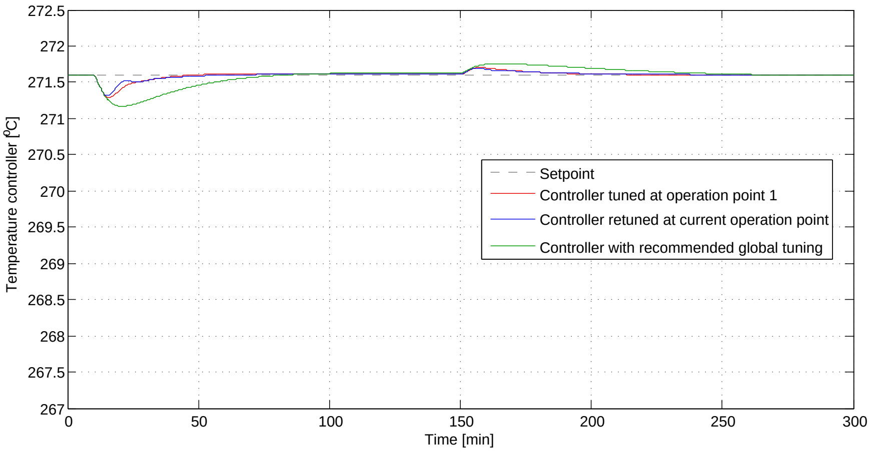
(c) Impurity in top product



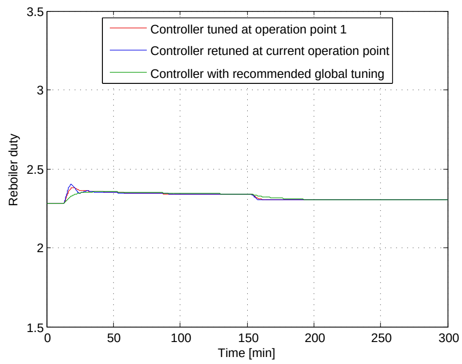
(d) Impurity in bottom product

Figure 14: Behavior of temperature controller a) at operation point 4b with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

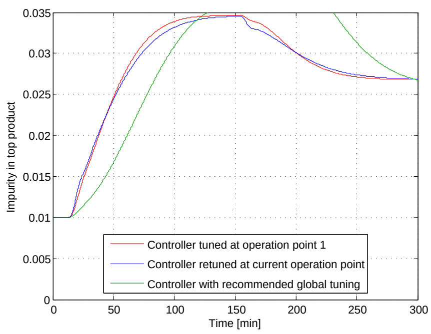
A.6 Operation point 4c



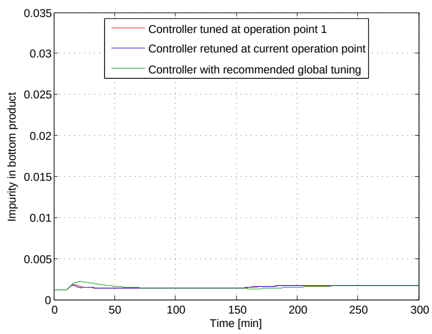
(a) Temperature controller



(b) Reboiler



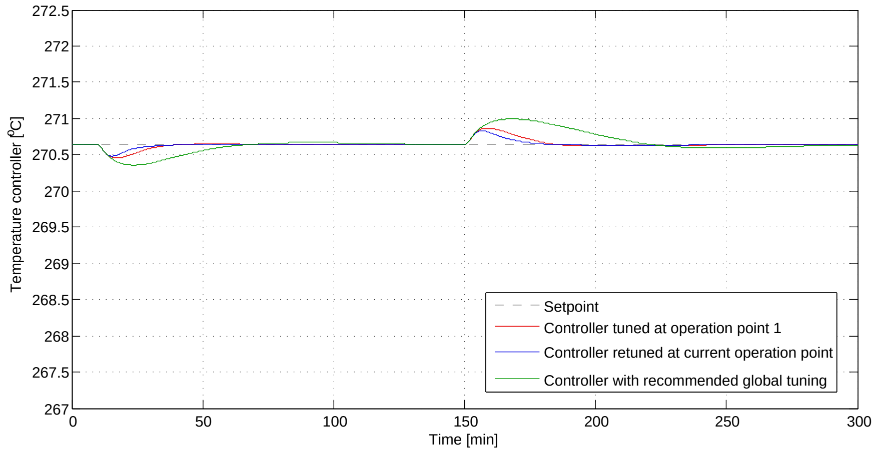
(c) Impurity in top product



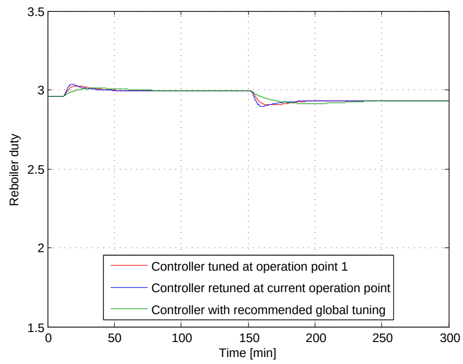
(d) Impurity in bottom product

Figure 15: Behavior of temperature controller a) at operation point 4c with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

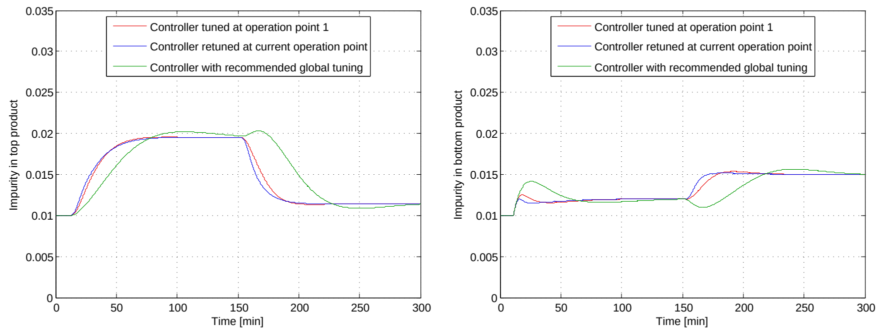
A.7 Operation point 5



(a) Temperature controller



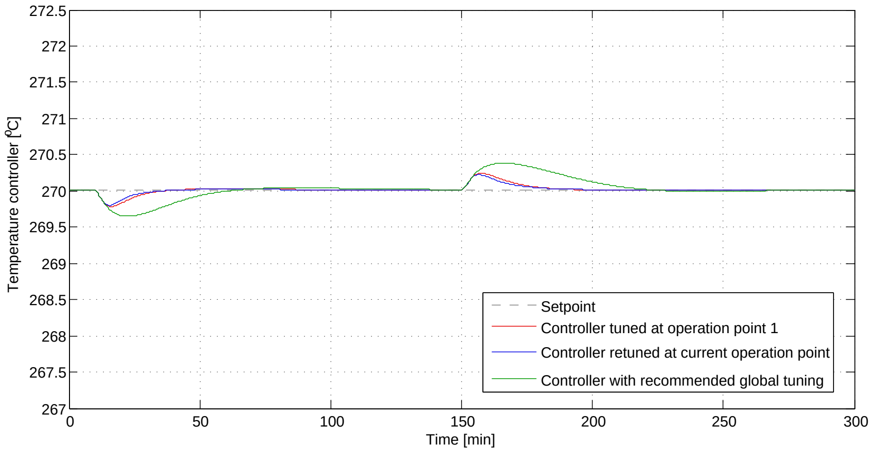
(b) Reboiler



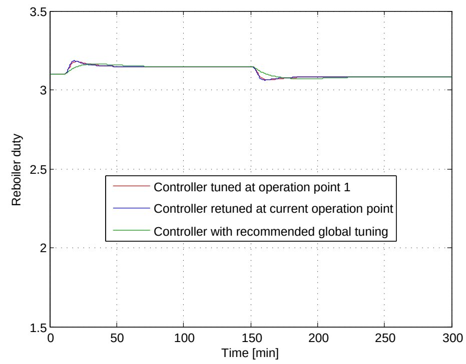
(c) Impurity in top product (d) Impurity in bottom product

Figure 16: Behavior of temperature controller a) at operation point 5 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

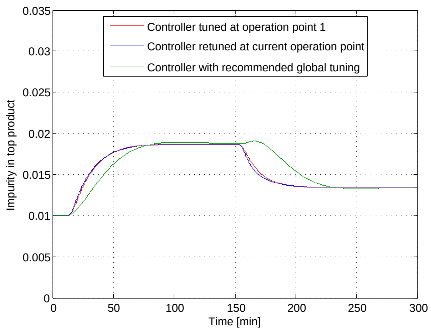
A.8 Operation point 6



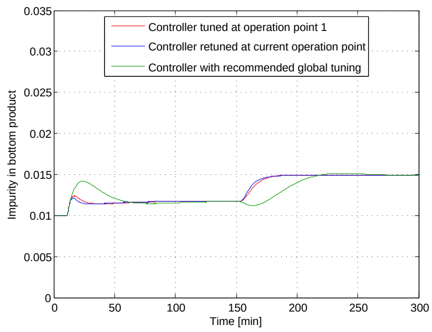
(a) Temperature controller



(b) Reboiler



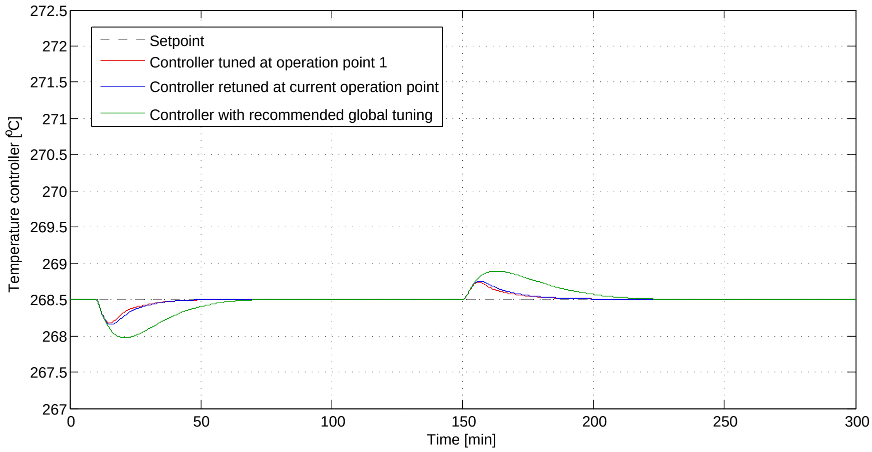
(c) Impurity in top product



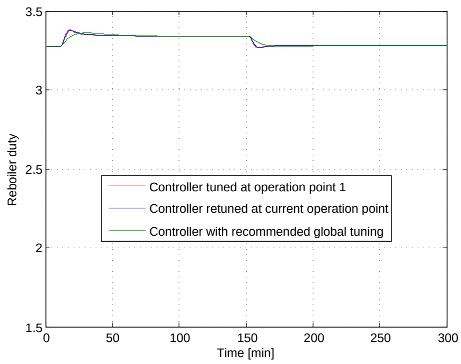
(d) Impurity in bottom product

Figure 17: Behavior of temperature controller a) at operation point 6 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

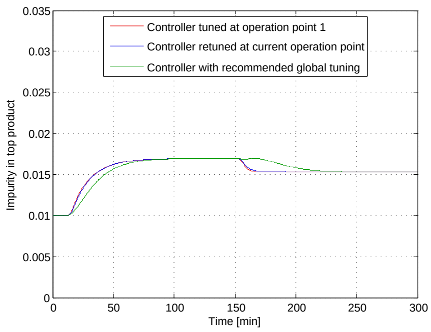
A.9 Operation point 7



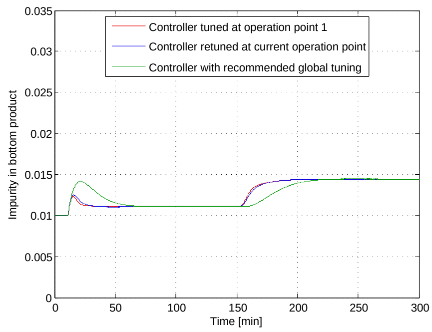
(a) Temperature controller



(b) Reboiler



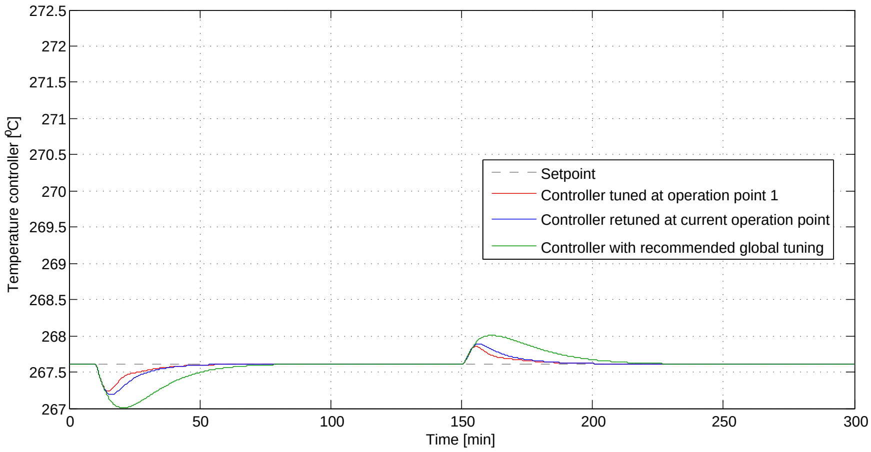
(c) Impurity in top product



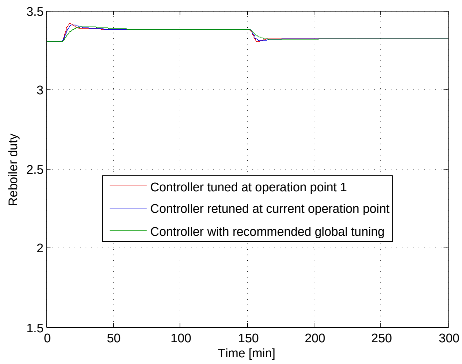
(d) Impurity in bottom product

Figure 18: Behavior of temperature controller a) at operation point 7 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

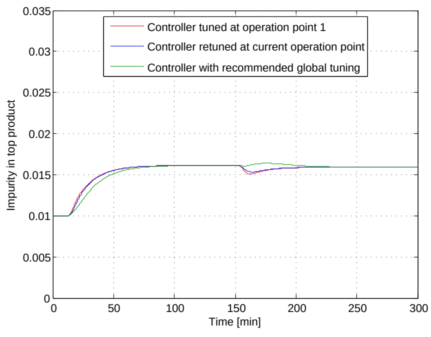
A.10 Operation point 8



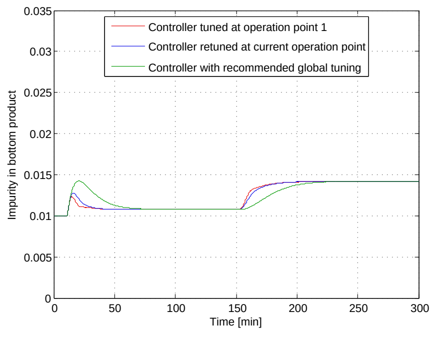
(a) Temperature controller



(b) Reboiler



(c) Impurity in top product



(d) Impurity in bottom product

Figure 19: Behavior of temperature controller a) at operation point 8 with disturbance in feed rate (dF=0.1) after 10 minutes and feed composition (dzF=-0.05) after 150 minutes. Reboiler duty b) was used as input for the controller resulting in impurity changes in respectively top c) and bottom product streams d).

B Models for MPC

B.1 Operation point 1

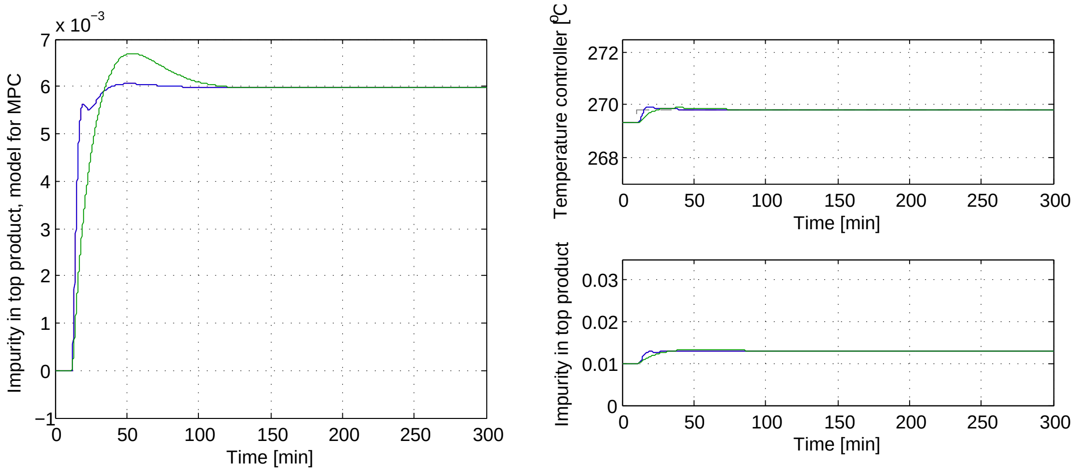


Figure 20: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

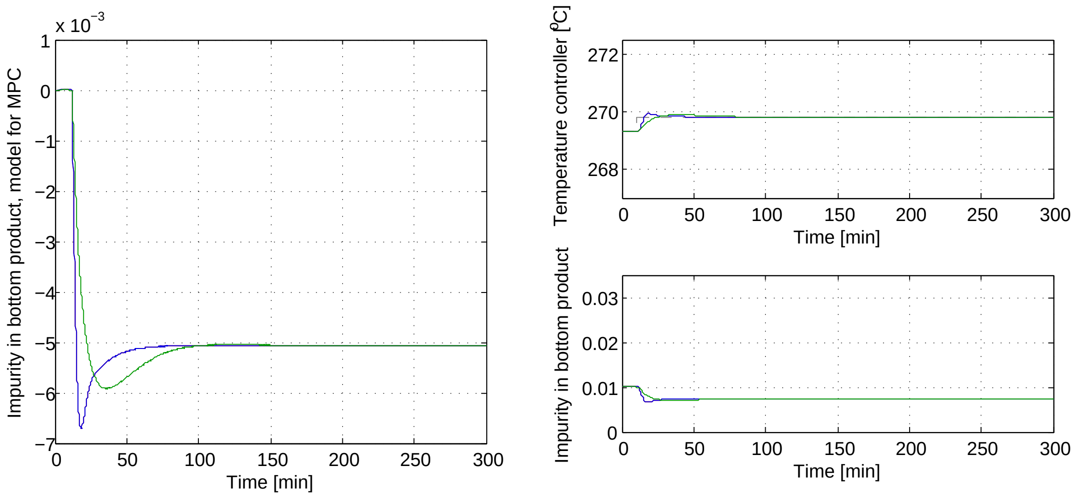


Figure 21: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

B.2 Operation point 3

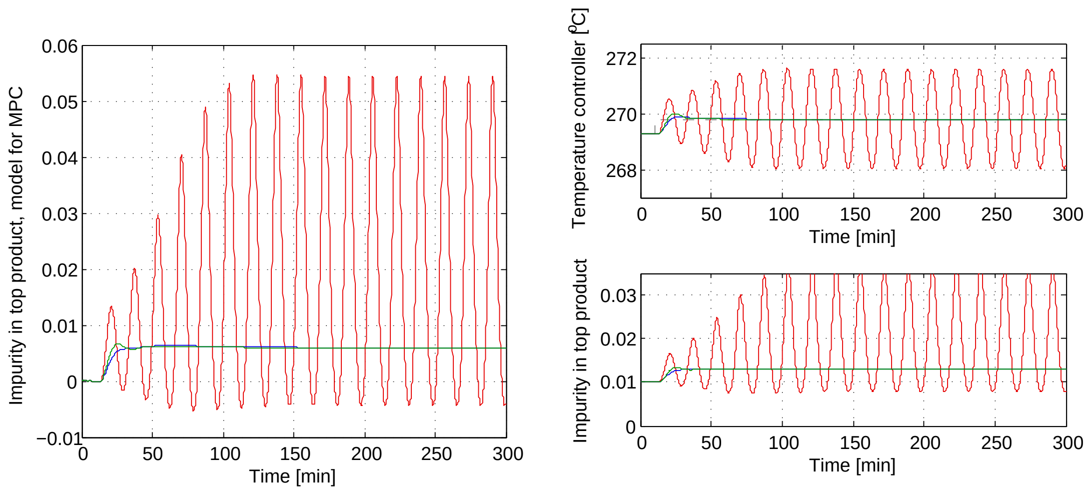


Figure 22: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

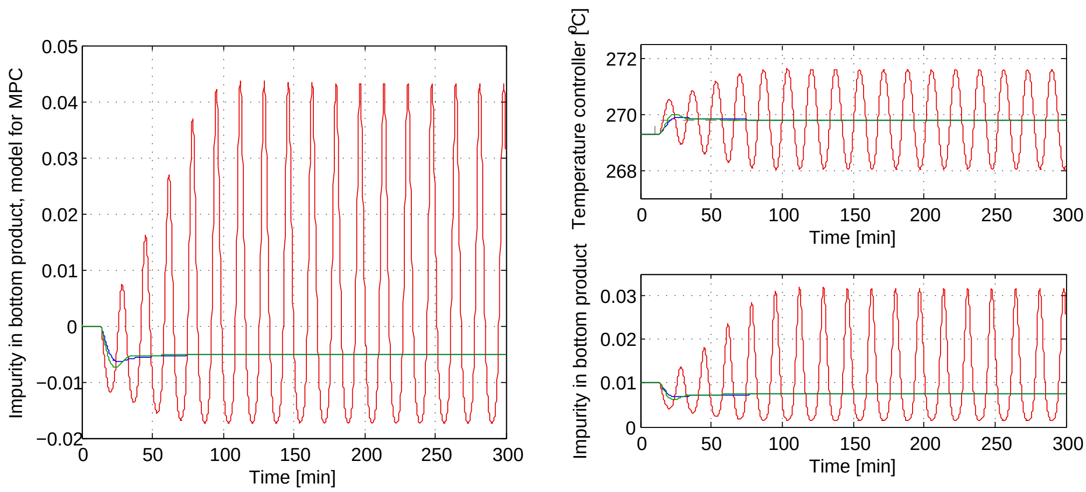


Figure 23: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

B.3 Operation point 4b

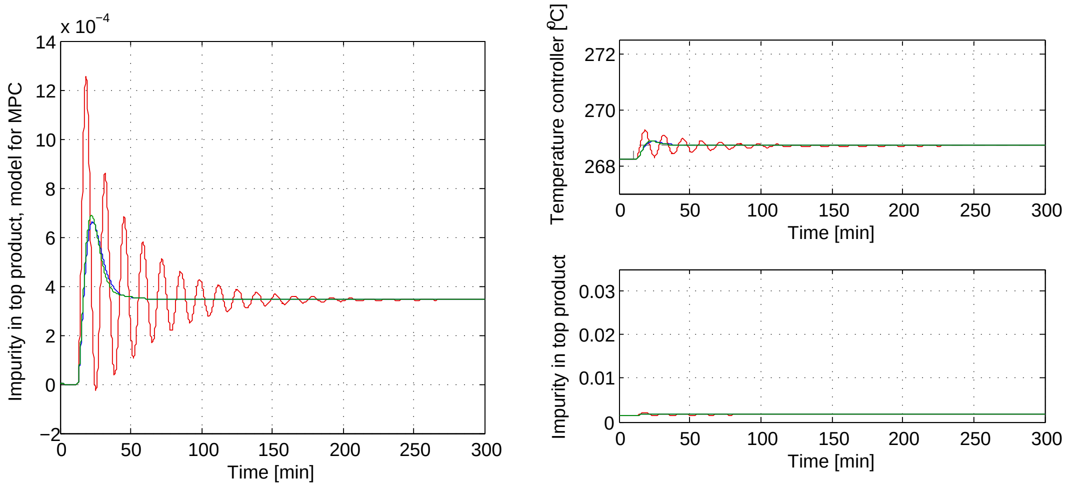


Figure 24: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

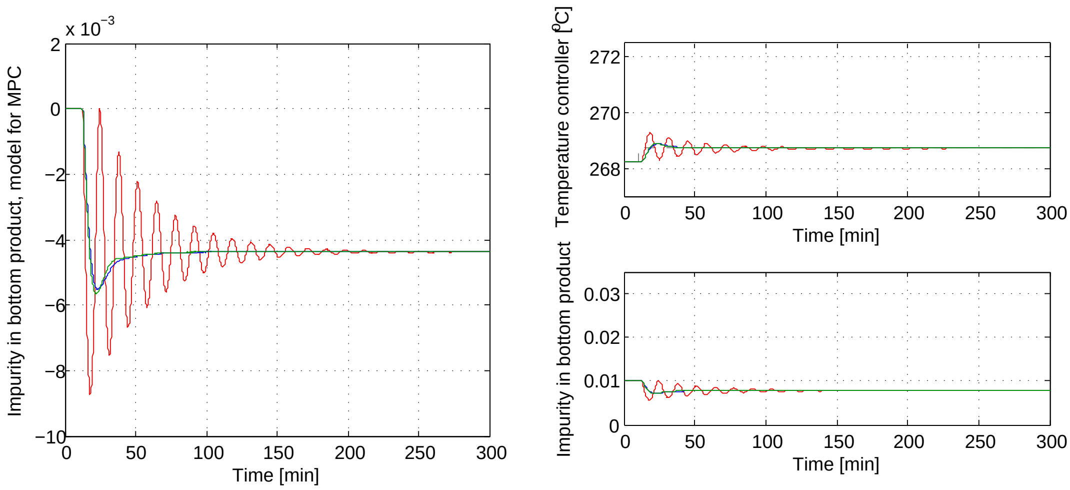


Figure 25: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

B.4 Operation point 4c

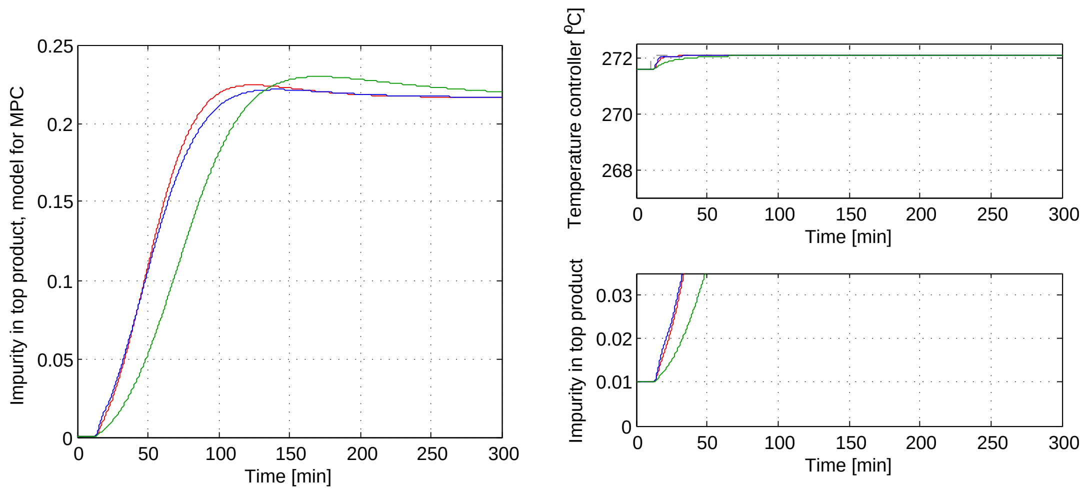


Figure 26: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

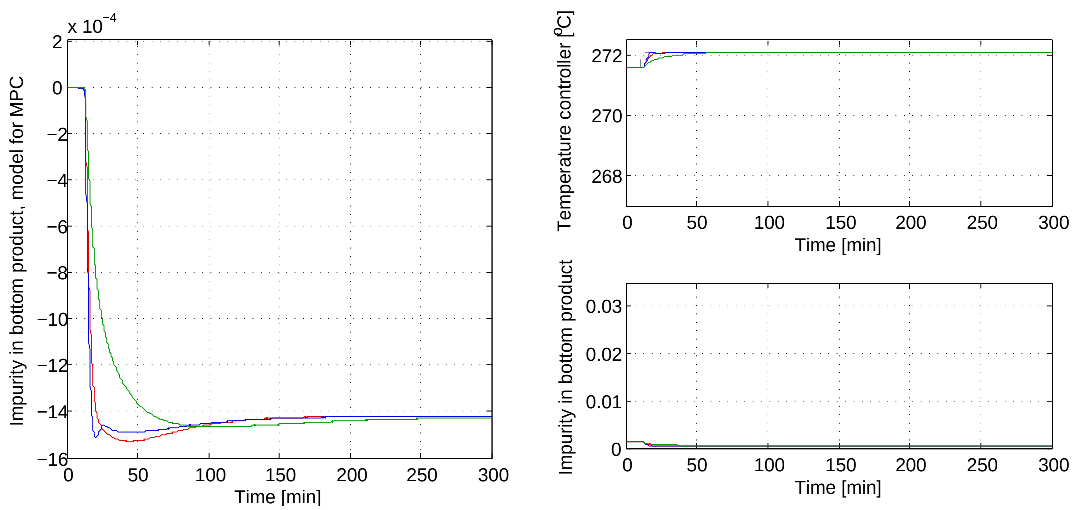


Figure 27: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

B.5 Operation point 5

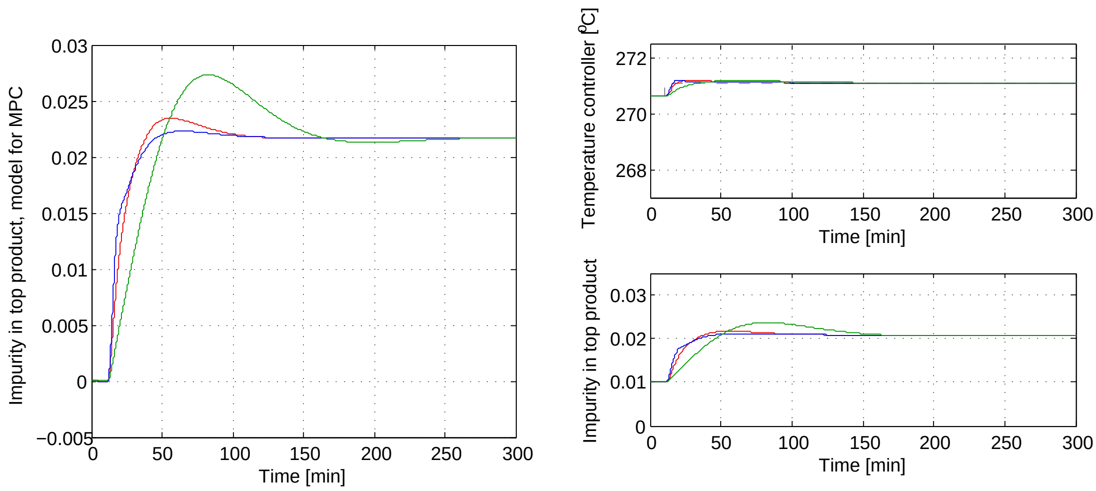


Figure 28: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

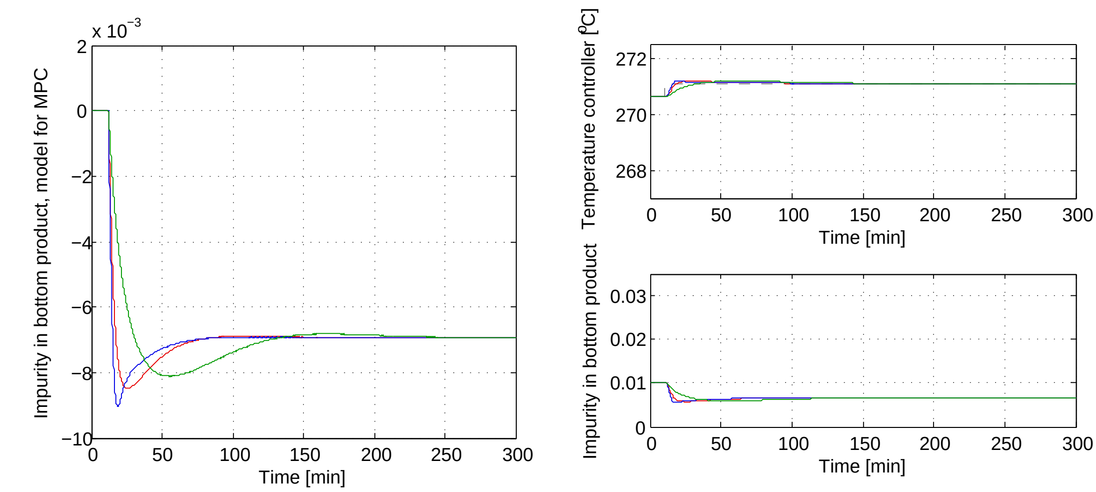


Figure 29: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

B.6 Operation point 8

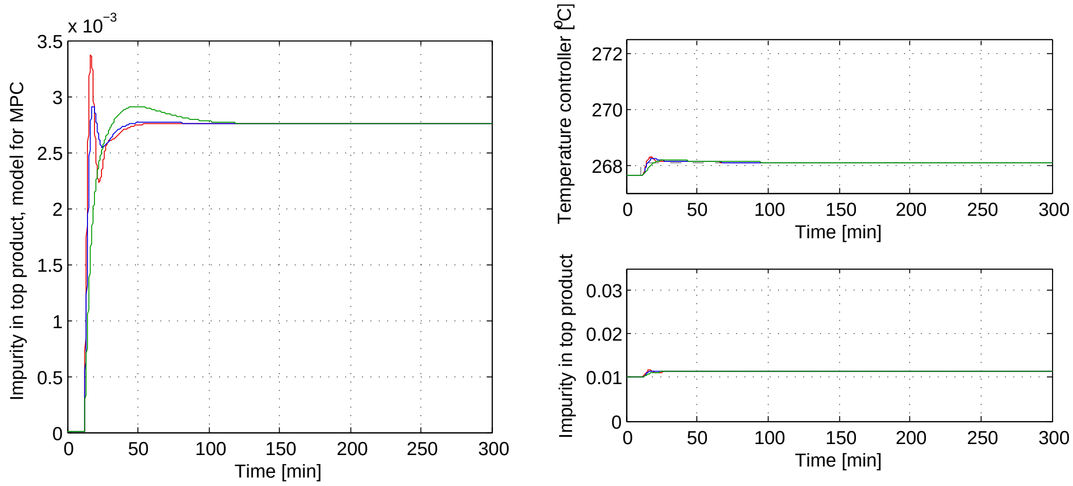


Figure 30: MPC model (left) for impurity in distillate (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

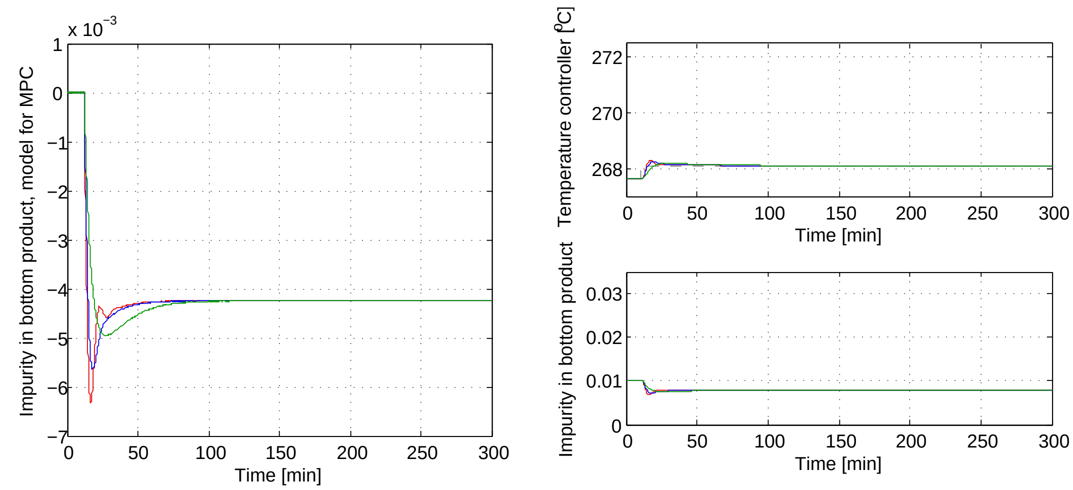
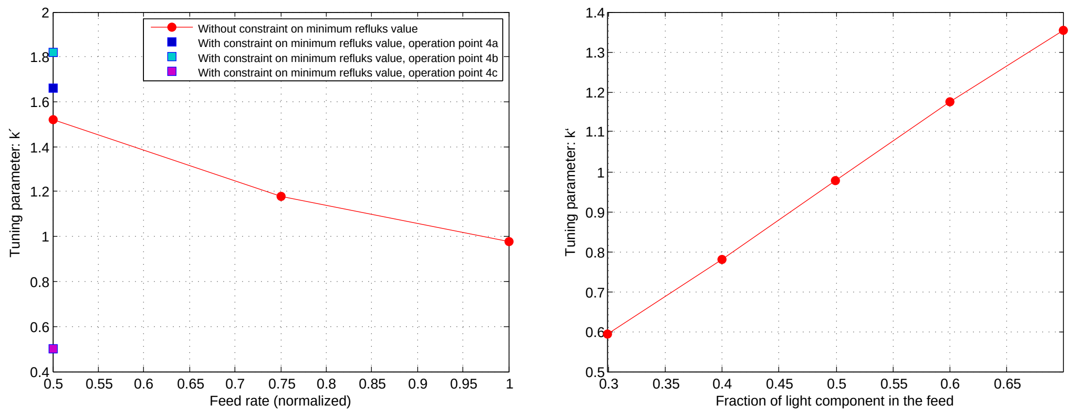


Figure 31: MPC model (left) for impurity in bottom product (bottom right) with generated step in setpoint of TC (top right). With TC tuned at operation point 1 (—), TC retuned at current operation point (—) and TC with recommended global tuning (—).

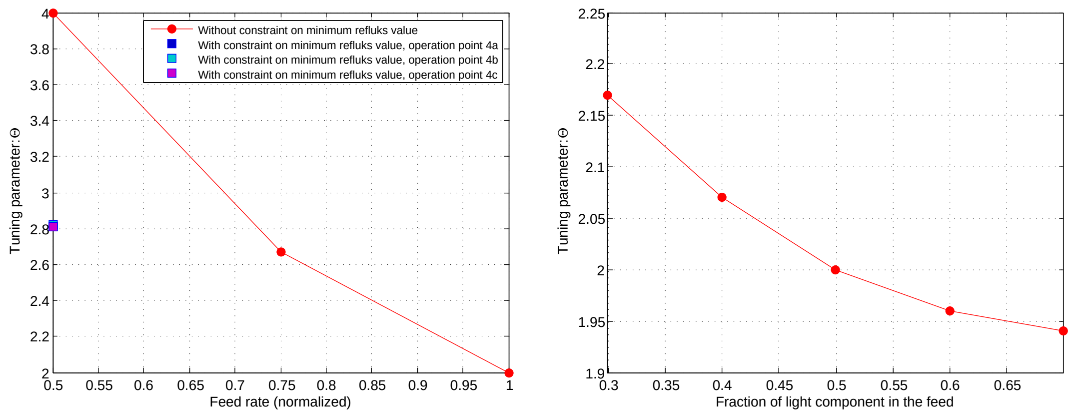
C Gain scheduling

Correlation between feed rate, feed composition and tuning parameters (k' and Θ).



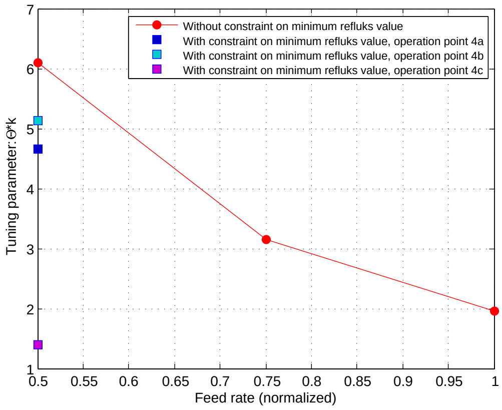
(a) Results of tuning at different feed rates (b) Results of tuning at different feed composition

Figure 32: Correlation between k' and feed rate a), feed composition b).

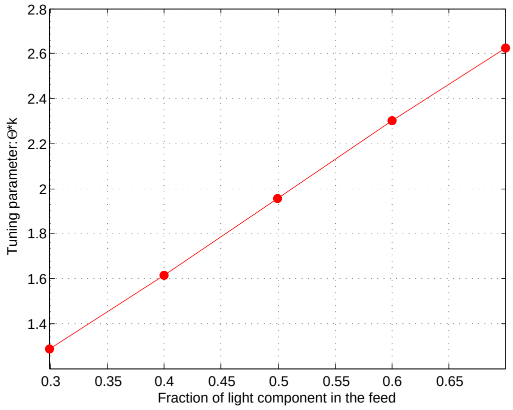


(a) Results of tuning at different feed rates (b) Results of tuning at different feed composition

Figure 33: Correlation between Θ and feed rate a), feed composition b).



(a) Results of tuning at different feed rates



(b) Results of tuning at different feed composition

Figure 34: Correlation between $k' \cdot \Theta$ and feed rate a), feed composition b).

D Matlab code

```
%———————————————————————————————————————————————————————————————
%———————————————————————————————————————————————————————————————
%_________________Starter of the Distillation column______________________
%<<<<                                                              >>>>%
%  This file can run both with open loop, to tune the controller based on %
%  which operation point is chosen, but also with closed loop to simulate %
%  how controller handles disturbances in feed rate and feed composition. %
%  To choose between open and closed loop open simulink file             %
%  "colas_nonlin_operation_All" and change the position of the           %
%  "Manual Switch" by dobbelklicking on it. Then run this file           %
%  "Starter_All".
%<<<<                                                              >>>>%


clc
clear all
clf
N=41;   %Number of stages
i=1:N; n(i)=i; % Matrix with stages in increasing order
Tb1=261.45; %Boiling point of iso-butane
Tb2=272.65; %Boiling point of n-butane
deltaTb= Tb1-Tb2; %Boiling point difference

%Initial delay in boiler
Timedelay=2;


%TC
TC=15; %Location of TC, numbering from the bottum of the column
T_under=TC-1; %Stages under TC
T_control=1;
T_over=N-TC; %Stages over TC

%Initial values
tot_time=5;
Steptime=0; Stepsize=0; %Reboiler
Steptime_setpunkt=0; Stepsize_setpunkt=0; %Setpoint in TC
dF_time=0; dF_size=0; dzF_time=0; dzF_size=0; % Disturbances

global OP F0 V0 L0

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
%                        Choose operation point                         %
OP=10;  %<<———————                                                       %
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%
if OP==1
    %Operation point 1
```

```
    load cola_init_1
    load Parameters
    P=P(1); I=I(1); T_s=T_s1(1);
elseif OP==2
    %Operation point 2
    load cola_init_2
    load Parameters
    P=P(2); I=I(2); T_s=T_s1(2);
elseif OP==3
    %Operation point 3
    load cola_init_3
    load Parameters
    P=P(3); I=I(3); T_s=T_s1(3);
elseif OP==4
    %Operation point 4a
    load cola_init_4a
    load Parameters
    P=P(4); I=I(4); T_s=T_s1(4);
elseif OP==5
    %Operation point 4b
    load cola_init_4b
    load Parameters
    P=P(5); I=I(5); T_s=T_s1(5);
elseif OP==6
    %Operation point 4c
    load cola_init_4c
    load Parameters
    P=P(6); I=I(6); T_s=T_s1(6);
elseif OP==7
    %Operation point 5
    load cola_init_5
    load Parameters
    P=P(7); I=I(7); T_s=T_s1(7);
elseif OP==8
    %Operation point 6
    load cola_init_6
    load Parameters
    P=P(8); I=I(8); T_s=T_s1(8);
elseif OP==9
    %Operation point 7
    load cola_init_7
    load Parameters
    P=P(9); I=I(9); T_s=T_s1(9);
elseif OP==10
    %Operation point 8
    load cola_init_8
    load Parameters
    P=P(10); I=I(10); T_s=T_s1(10);
else
    disp('Error in determing the operation point')
end
```

```
F0=Uinit(5);
V0=Uinit(2);
L0=Uinit(1);
% Runs simulation to fin if the loop is closed or opened
sim('colas_nonlin_operation_All')
Timedelay= Timedelay*3.20629/Setpoint_V(1);
%%


if any(RUN)>=1
%------------------ Runing with closed loop ------------------------------
    Steptime_setpunkt=10; Stepsize_setpunkt=0.0;
    dF_time=10; dF_size=0.1; dzF_time=150; dzF_size=-0.05;
    tot_time=300;
    sim('colas_nonlin_operation_All')
    Y=T_on15th; % To be analysed
    y1_1=y1;
    y2_1=y2;
    V1=V;
    t1=t;

    %Plot of Temperatureprofiles
    figure (1)
    subplot(1,2,1)
    T_grader=T-273.25;
    figure(1)
    plot(n,T_grader(1,:),n,T_grader(end,:),TC,T_grader(end,TC), 'o')
    legend('Temperatureprofile before disturbance',...
        'Temperatureprofile after disturbance',...
        'Location of temperature controller')
    title('Temperatureprofiles','BackgroundColor',[.99 .99 .99])
    xmin=n(1);
    xmax=n(end);
    ymin=min(T_grader(1,end),T_grader(end,end));
    ymax=max(T_grader(1,1),T_grader(end,1));
    axis([xmin xmax ymin ymax])
    grid on
    xlabel('Stage number')
    ylabel('Temperature [oC]')

    %Plot of steprespons
    subplot(1,2,2)
    plot(t, Setpoint_T, t, Y)
    legend('Step in setpoint','Step respons')
    title('Behaviour of the temperature contoller with change in setpoint',...
        'BackgroundColor',[.99 .99 .99])
    grid on
    xlabel('Time [min]')
    ylabel('Temperature controler [oC]')
```

```
    %Simulation with tuning done in Operation point 1
    clear P I
    load Parameters
    P=P(1); I=I(1);
    sim('colas_nonlin_operation_All')
    Y2=T_on15th;
    y1_2=y1;
    y2_2=y2;
    V2=V;
    t2=t;

    %Simulation with Global tuning (tuning from operation point 4a)
    clear P I
    load Parameters
    P=P(4); I=I(4);
    sim('colas_nonlin_operation_All')
    Y3=T_on15th;
    y1_3=y1;
    y2_3=y2;
    V3=V;
    t3=t;


%------------------------------------- PLOTS -------------------------------------------------------
%-------------------------- Temperature controller ----------------------------------------------
    if OP==1
        figure (2)
        plot(t3, Setpoint_T, '--','color', [0.5 0.5 0.5])
        hold on
        plot(t2, Y2, 'color', [0.9 0 0])
        hold on
        plot(t1, Y, '--', 'color', [0 0 0.9])
        hold on
        plot(t3, Y3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 267 272.5])
        xlabel('Time [min]')
        ylabel('Temperature controller [^oC]')
        legend('Setpoint','Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    %------------------- Impurity in top product ---------------------------------------
        figure (3)
        plot(t2, y1_2, 'color', [0.9 0 0])
        hold on
        plot(t1, y1_1, '--', 'color', [0 0 0.9])
        hold on
        plot(t3, y1_3, 'color', [0 0.6 0])
```

```
        grid on
        axis([0 tot_time 0 0.035])
        xlabel('Time [min]')
        ylabel('Impurity in top product')
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    %---------- Impurity in bottom product ---------------------------
        figure (4)
        plot(t2, y2_2, 'color', [0.9 0 0])
        hold on
        plot(t1, y2_1, '--', 'color', [0 0 0.9])
        hold on
        plot(t3, y2_3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 0 0.035])
        xlabel('Time [min]')
        ylabel('Impurity in bottom product')
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

      %------------ Reboiler duty ------------------------------------
        figure (5)
        plot(t2, V2, 'color', [0.9 0 0])
        hold on
        plot(t1, V1, '--', 'color', [0 0 0.9])
        hold on
        plot(t3, V3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 1.5 3.5])
        xlabel('Time [min]')
        ylabel('Reboiler duty')
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    elseif OP==4
        figure (2)
        plot(t3, Setpoint_T, '--','color', [0.5 0.5 0.5])
        hold on
        plot(t2, Y2, 'color', [0.9 0 0])
        hold on
        plot(t1, Y, 'color', [0 0 0.9])
        hold on
        plot(t3, Y3, '--', 'color', [0 0.6 0])
        grid on
```

```
    axis([0 tot_time 267 272.5])
    xlabel('Time [min]')
    ylabel('Temperature controller [^oC]')
    legend('Setpoint','Controller tuned at operation point 1',...
        'Controller retuned at current operation point^{ }_{ }_{ }',...
        'Controller with recommended global tuning',...
        'Location', 'Best')

%---------- Impurity in top product -----------------------
    figure (3)
    plot(t2, y1_2, 'color', [0.9 0 0])
    hold on
    plot(t1, y1_1, 'color', [0 0 0.9])
    hold on
    plot(t3, y1_3, '--', 'color', [0 0.6 0])
    grid on
    axis([0 tot_time 0 0.035])
    xlabel('Time [min]')
    ylabel('Impurity in top product')
    legend('Controller tuned at operation point 1',...
        'Controller retuned at current operation point^{ }_{ }_{ }',...
        'Controller with recommended global tuning',...
        'Location', 'Best')

%-------- Impurity in bottom product -----------------------
    figure (4)
    plot(t2, y2_2, 'color', [0.9 0 0])
    hold on
    plot(t1, y2_1, 'color', [0 0 0.9])
    hold on
    plot(t3, y2_3, '--', 'color', [0 0.6 0])
    grid on
    axis([0 tot_time 0 0.035])
    xlabel('Time [min]')
    ylabel('Impurity in bottom product')
    legend('Controller tuned at operation point 1',...
        'Controller retuned at current operation point^{ }_{ }_{ }',...
        'Controller with recommended global tuning',...
        'Location', 'Best')

  %------------ Reboiler duty -----------------------------
    figure (5)
    plot(t2, V2, 'color', [0.9 0 0])
    hold on
    plot(t1, V1, 'color', [0 0 0.9])
    hold on
    plot(t3, V3, '--', 'color', [0 0.6 0])
    grid on
    axis([0 tot_time 1.5 3.5])
    xlabel('Time [min]')
    ylabel('Reboiler duty')
```

```
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    else
        figure (2)
        plot(t3, Setpoint_T, '--','color', [0.5 0.5 0.5])
        hold on
        plot(t2, Y2, 'color', [0.9 0 0])
        hold on
        plot(t1, Y, 'color', [0 0 0.9])
        hold on
        plot(t3, Y3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 267 272.5])
        xlabel('Time [min]')
        ylabel('Temperature controller [^oC]')
        legend('Setpoint','Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    %------------ Impurity in top product ---------------------------
        figure (3)
        plot(t2, y1_2, 'color', [0.9 0 0])
        hold on
        plot(t1, y1_1, 'color', [0 0 0.9])
        hold on
        plot(t3, y1_3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 0 0.035])
        xlabel('Time [min]')
        ylabel('Impurity in top product')
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

    %---------- Impurity in bottom product ---------------------------
        figure (4)
        plot(t2, y2_2, 'color', [0.9 0 0])
        hold on
        plot(t1, y2_1, 'color', [0 0 0.9])
        hold on
        plot(t3, y2_3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 0 0.035])
        xlabel('Time [min]')
        ylabel('Impurity in bottom product')
        legend('Controller tuned at operation point 1',...
```

```
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')

      %------------- Reboiler duty ------------------------------
        figure (5)
        plot(t2, V2, 'color', [0.9 0 0])
        hold on
        plot(t1, V1, 'color', [0 0 0.9])
        hold on
        plot(t3, V3, 'color', [0 0.6 0])
        grid on
        axis([0 tot_time 1.5 3.5])
        xlabel('Time [min]')
        ylabel('Reboiler duty')
        legend('Controller tuned at operation point 1',...
            'Controller retuned at current operation point^{ }_{ }_{ }',...
            'Controller with recommended global tuning',...
            'Location', 'Best')
    end
else
%----------------- Runing with open loop ------------------------------
%-----------------  For tuning of TC  ---------------------------------
    Steptime=15; Stepsize=0.02;
    tot_time=500;
    sim('colas_nonlin_operation_All')
    Y=T_on15th; % To be analysed


    %Plot of Temperatureprofiles
    figure (1)
    subplot(2,2,[1 3])
    T_grader=T-273.25;
    figure(1)
    plot(n,T_grader(1,:),n,T_grader(end,:),TC,T_grader(end,TC), 'o')
    legend('Temperatureprofile before disturbance',...
        'Temperatureprofile after disturbance',...
        'Location of temperature controller')
    title('Temperatureprofiles','BackgroundColor',[.99 .99 .99])
    xmin=n(1);
    xmax=n(end);
    ymin=min(T_grader(1,end),T_grader(end,end));
    ymax=max(T_grader(1,1),T_grader(end,1));
    axis([xmin xmax ymin ymax])
    grid on
    xlabel('Stage number')
    ylabel('Temperature [oC]')


    %SIMC tuning
    i=1:length(t); t_simc(i)=i;
```

```
t_start=round(interp1(t,t_simc,Steptime+Timedelay,'linear'));     %or '↩
    nearest'
k63=Y(t_start)+(.63*(Y(end)-Y(t_start)));

delay=Timedelay;
t_action= t(t_start+delay:end);
Y_action=Y(t_start+delay:end);

%Finding tau_c
tau_c=1*delay;


%Finding timekonstant
[diff63 I]=min(abs(Y_action-k63));
tau1_plot=t_action(I);
%tau12=t(I)
tau1=tau1_plot-delay-Steptime;


subplot (2,2,2)
plot(t, Y, tau1_plot, k63, 'o')
hold on
[AX,H1,H2]= plotyy(tau1_plot, k63, t, Setpoint_V);
set(get(AX(1),'Ylabel'),'String','Stepresponse in ouput');
set(get(AX(2),'Ylabel'),'String','Step in input');
grid on
legend('Step respons', '\tau_1')
title('Step respons in V-T controller','BackgroundColor',[.99 .99 .99])

%Finding K_c
Range_Y=1;
Range_u=1;
k=((Y(end)-Y(1))/Range_Y)/(Stepsize/Range_u);
K_c=tau1/(k*(tau_c+delay));

%Finding tau_I
tau_I=min(tau1, 4*(tau_c+delay));

subplot (2,2,4)
text(0.05,0.7, ['SIMC tuning results', char(10)...
    char(10)...
    'delay=',num2str(delay), char(10)...
    '\tau_c=',num2str(tau_c), char(10)...
    '\tau_1=',num2str(tau1), char(10)...
    'k=',num2str(k), char(10)...
    'K_c=' ,num2str(K_c), char(10)...
    '\tau_I=',num2str(tau_I), char(10)...
    'K_c/\tau_I=',num2str(K_c/tau_I)],...
    'BackgroundColor', 'white', 'EdgeColor',[0 0 0])
set(gca,'Visible','off')
```

```
    %Assume pure integrator
    k_merket= (Y_action(150)-Y_action(1))/(Stepsize*(t_action(150)-t_action(1)))↩
        ;
    K_c2= 1/(k_merket*(tau_c+delay));
    tau_I2= 4*(tau_c+delay);
    subplot (2,2,4)
    text(0.7,0.7, ['SIMC tuning results 2', char(10)...
        char(10)...
        'delay=',num2str(delay), char(10)...
        '\tau_c=',num2str(tau_c), char(10)...
        'k`=',num2str(k_merket), char(10)...
        'K_c=' ,num2str(K_c2), char(10)...
        '\tau_I=',num2str(tau_I2), char(10)...
        'K_c/\tau_I=',num2str(K_c2/tau_I2)],...
        'BackgroundColor', 'white', 'EdgeColor',[ 0 0 0])
    set(gca,'Visible','off')

end
```

E Simulink model

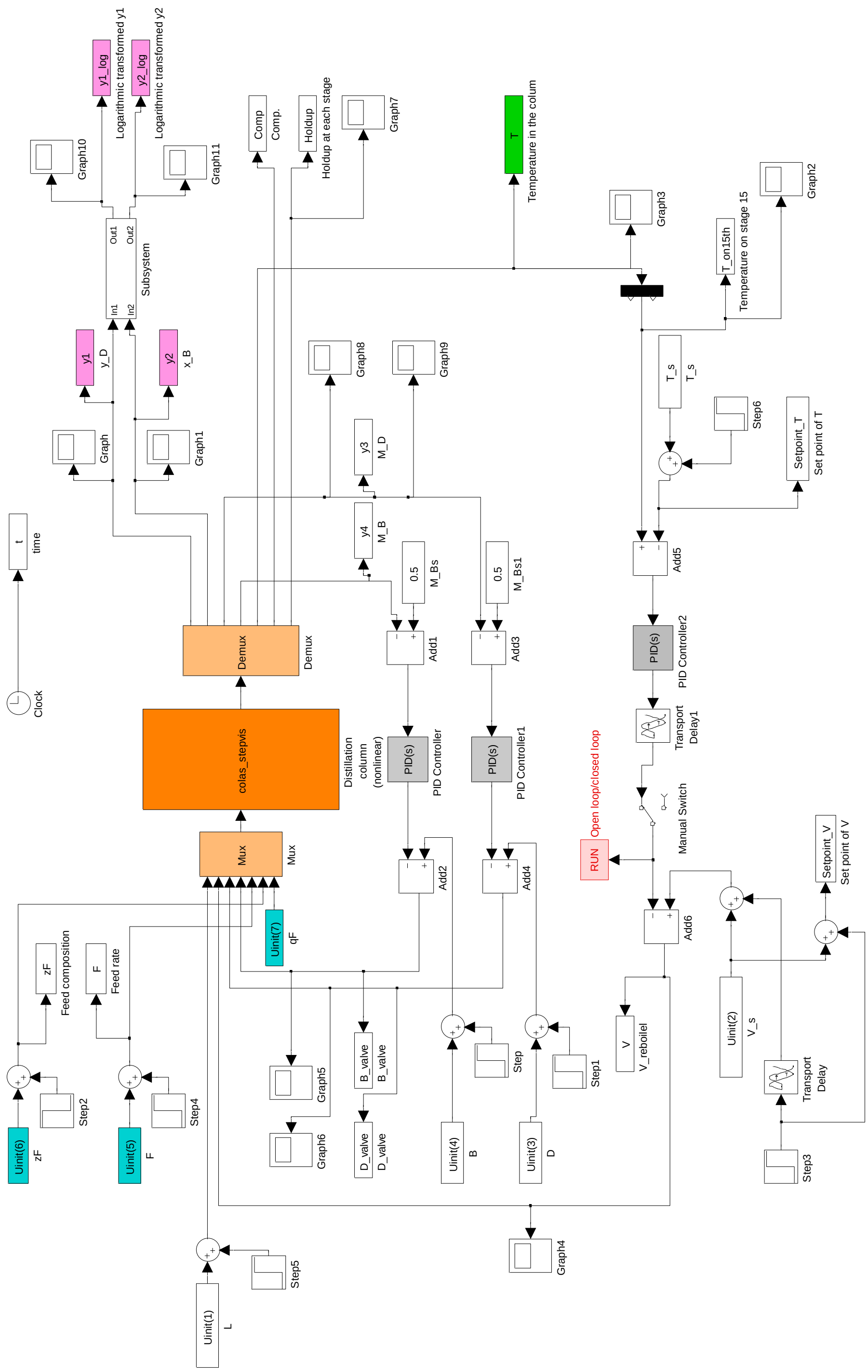


Figure 35: Simulink model used in simulation.

F Risk assessment

No risks were assumed to be relevant for this project.