

Operation and Control of a New Pressure Swing Batch Distillation System

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The pressure swing separation of a binary minimum azeotrope in a double column batch stripper is studied by rigorous simulation. For controlling the product compositions a simple scheme is presented. The parameters of PID controllers are determined. Influence of the most important parameter (division ratio of liquid leaving the top vessel) is investigated. The calculations were made for the mixture n-pentane-acetone by using a professional dynamic simulator.

1. Introduction

Binary pressure sensitive azeotropes can be separated by pressure swing distillation (PSD). Continuous PSD was first applied in the industry in 1928. Phimister and Seider (2000) studied first the batch (stripping) and semicontinuous application of PSD by simulation. Repke et al. (2007) were the first, who investigated experimentally the batch PSD (PSBD, pilot-plant experiments for the separation of a minimum azeotrope in a batch rectifier (BR) and stripper (BS)). Modla and Lang (2008) studied different batch configurations (BR, BS, combination of BR and BS and middle vessel column (MVC)) by feasibility studies and rigorous simulation for the separation binary (max. and min.) homoazeotropes. By modifying the MVC, which has not been proven suitable for the PSBD, they suggested two new double column batch configurations: rectifier (DCBR, Fig. 1a) and stripper (DCBS, Fig. 1b). They compared the different configurations for a given set of operational parameters without optimisation and control. For min. azeotropes the best results (minimal specific energy consumption for the same quality products) were obtained with the DCBS and for maximum azeotropes with the DCBR, respectively. The columns of these configurations can be operated practically in steady state.

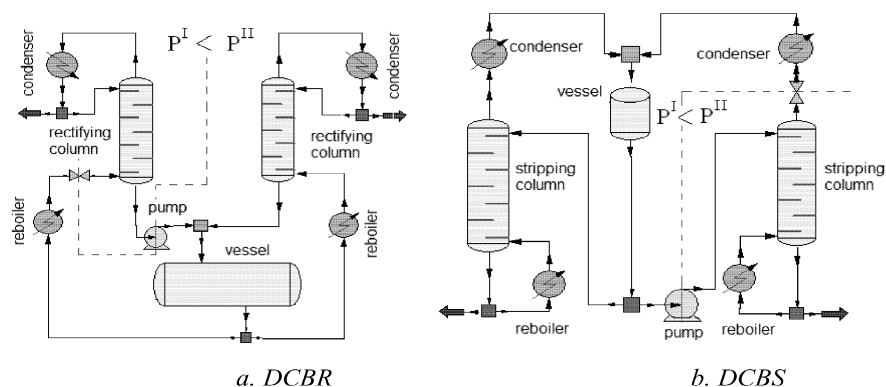


Figure 1. The scheme of a DCBR and DCBS

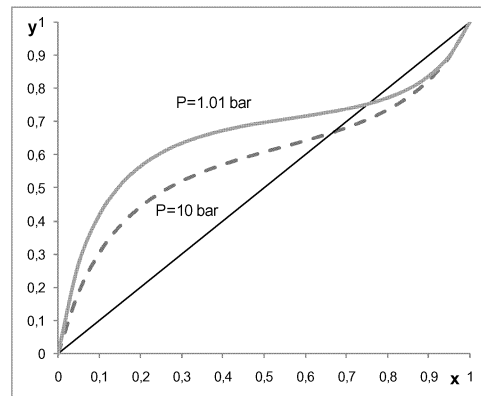
When operating these new configurations the liquid composition of the common vessel of the two columns must be kept between the two azeotropic compositions. The ratio of two product flow rates of a DCBS can be changed by varying the reboil ratios and/or the ratio of division of the liquid flow leaving the common vessel.

The goals of this paper are:

- to investigate the operation of the DCBS for the separation of a minimum azeotrope,
- to study a simple scheme for the control of product compositions,
- to investigate the effects and to determine the optimal value of the liquid division ratio.

The calculations were made for the mixture n-pentane-acetone by using a professional dynamic simulator (CCDCOLUMN).

The y - x equilibrium diagram and azeotropic data of the mixture studied are shown for the two different pressures in Fig. 2 and Table 1, respectively.



Comp.	pentane(A)	acetone(B)
P [bar]	1.01	10
x_{az}	0.754	0.668
T_{az} [°C]	32.5	116.9
$T_{BP,A}$ [°C]	36.0	124.7
$T_{BP,B}$ [°C]	56.2	142.9

Figure 2. y - x diagrams of n-pentane-acetone Table 1. Azeotropic data (UNIQUAC parameters: 571.98 and 95.033 cal/mol)

2. Simulation method

The following simplifying assumptions were applied

- theoretical stages,
- negligible vapour hold-up,
- constant volumetric liquid plate hold-up.

The model equations to be solved are well known:

- a. Non-linear differential equations (material balances, heat balances),
- b. Algebraic equations (vapour-liquid equilibrium (VLE) relationships, summation equations, hold-up equivalence, physical property models).

For solving the above model equations we used the CCDCOLUMN flow-sheet simulator (ChemCad 6.0). For the simulation of columns simultaneous correction method was applied.

The following modules were used:

- DYNCOLUMN (column sections),
- DYNAMIC VESSEL (top vessel and product tanks),

-HEAT EXCHANGER, PUMP, VALVE,
 -MIXER, DIVIDER,
 -CONTROLLER, CONTROL VALVE.

The ChemCad model of the double column batch stripper with control of product compositions is shown in Fig. 3.

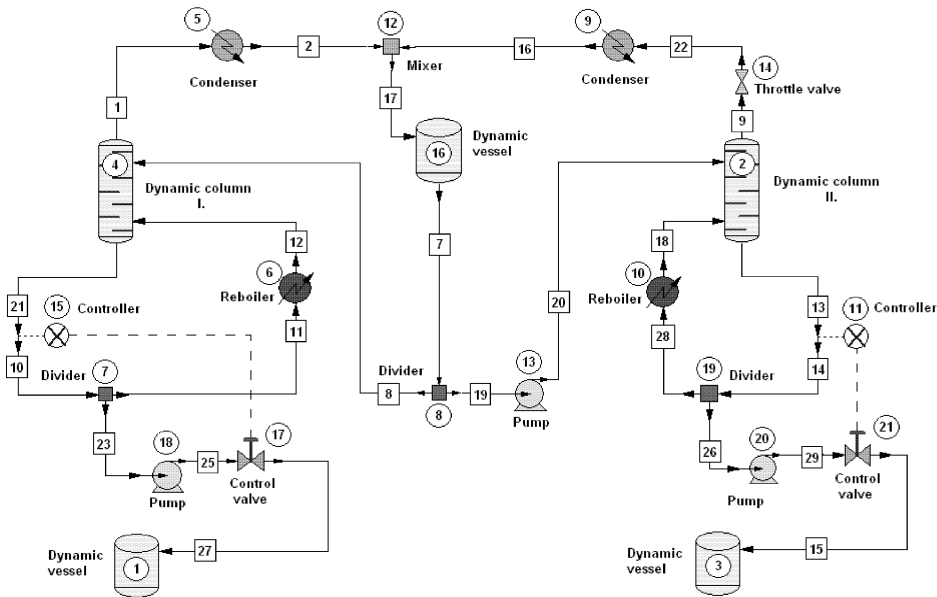


Figure 3. ChemCad model of the double column batch stripper.

3. Results

The number of theoretical stages for each column sections is 40. (The total condenser and total reboiler do not provide a theoretical stage.) The liquid hold-up is $2 \text{ dm}^3/\text{plate}$. The pressure of the columns: $P^{\text{LP}}=1.013 \text{ bar}$ and $P^{\text{HP}}=10 \text{ bar}$. At the start of the distillation plates of the columns are wet (they are filled with charge at its boiling point at the given pressure). The total flow rate of liquid leaving the common vessel: $L_{0,\text{total}} = L_0^{\text{LP}} + L_0^{\text{HP}} = 6 \text{ m}^3/\text{h}$. The quantity of charge containing 30 mol% pentane is 4.022 kmol (0.471 m^3). The prescribed purity is 98 mol% for both product. The reboil ratios R_s^{LP} and R_s^{HP} are changed by PID controllers manipulating the product flow rates W^{LP} and W^{HP} , respectively. The whole process is finished when the amount of liquid in the vessel decreases to 12.5 % of the charge.

First, the parameters of the two PID controllers (A_p , T_I and T_D) providing stable, good quality control of the product compositions in the whole region of liquid division ratio ($\phi=L_0^{\text{LP}}/L_{0,\text{total}}$) studied are determined. Then, the influence of this operational parameter on the performance of the PSBS is studied and its optimum value yielding the minimal overall specific energy consumption $((SQ^{\text{LP}} + SQ^{\text{HP}})/(SW^{\text{LP}} + SW^{\text{HP}}))$ is determined.

3.1 Tuning of PID controllers

Our aim is to determine a set of parameters of the PID controllers which provide good quality control of product compositions by taking into consideration the usual criterions (maximal overshoot, control time, number of oscillations).

The quality of control is determined by the evolution of not only the controlled variables (composition of the two bottoms products) but also that of the position of the two control valves (varying the flow rate of the two bottoms product). The following criteria of quality of control are given concerning the two control valves:

- maximal overshoot: 25 %,
- maximum number of oscillations during the settling time T_S (within an error band of $\pm 5\%$): 3.

In Fig. 4 the evolution of the position of the control valve and bottoms composition of Column I for an inappropriate set of controller parameters is shown. (Both position of the control valve and value of controlled variable show oscillations without damping.)

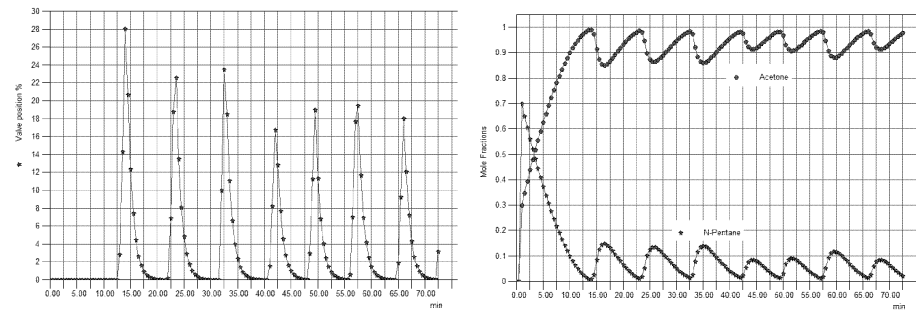


Figure 4. Evolution of the position of the control valve and bottoms composition for an inappropriate setting of PID parameters (Column I, $PB=10\%$, $T_I=1\text{ min}$, $T_D=0$)

For an appropriate tuning the evolution of the position of the control valves, bottoms compositions and reboil ratios is shown in Fig. 5. (Table 2 contains the parameters of PID controllers and control quality data.)

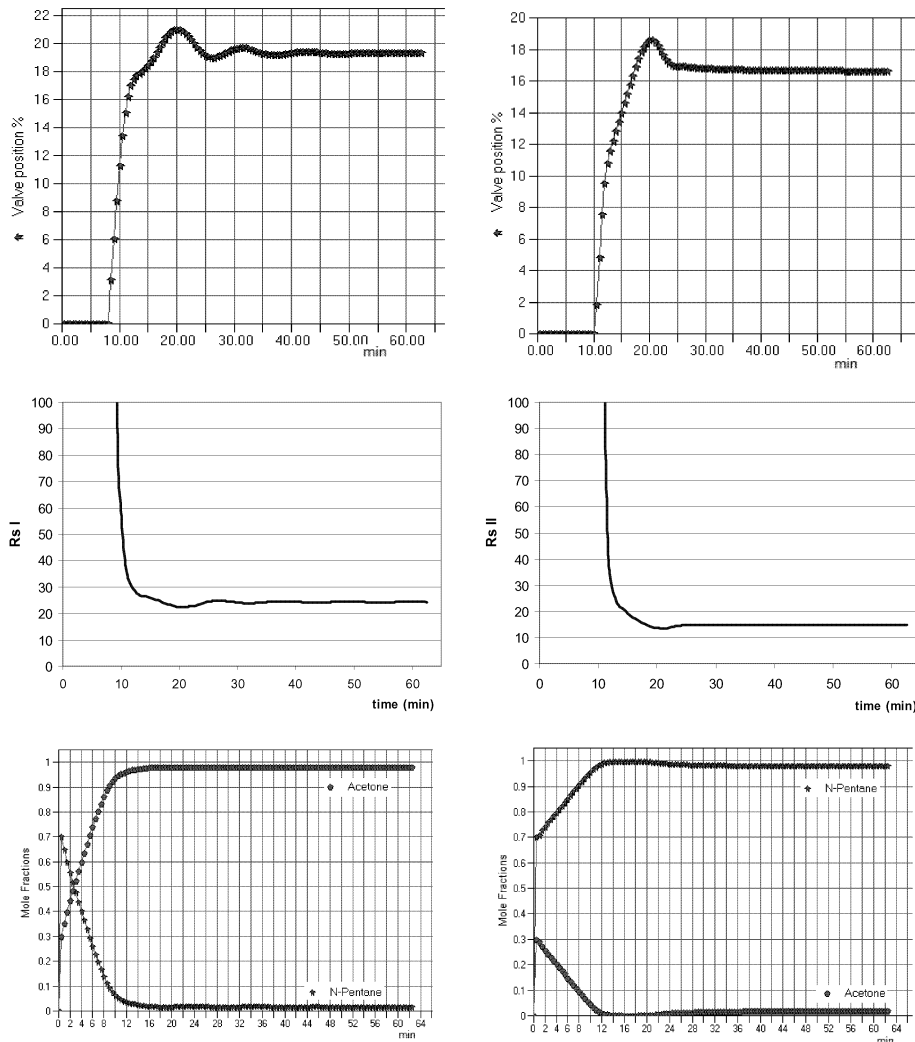
Table 2. PID parameters and control quality data for an appropriate tuning

a. PID parameters:

	PB, %	T_I , min	T_D , min	Set point
Column I	80	2.5	0	0.982
Column II	50	10	1	0.98

b. Control quality data:

	Column I:	Column II:
Maximal overshoot:	$(21.03-19.35)/19.35=0.08$	$(18.60-16.66)/16.66=0.11$
Settling time, min:	$28.5-8=20.5$	$25-10=15$
No. of oscillations within T_S :	2	1



a. Column I

b. Column II

Fig. 5. Evolution of control valve positions, reboil ratios and bottoms composition for an appropriate tuning

3.2 Influence of the liquid division ratio

The liquid division ratio is varied in the region 0.3-0.9. The specific energy consumption is minimal at $\phi=0.55$ (Fig. 6). Prescribed purity products are obtained with good recovery (Table 3). This table contains also the most important results for the process, such as the total and specific energy consumptions of the production.

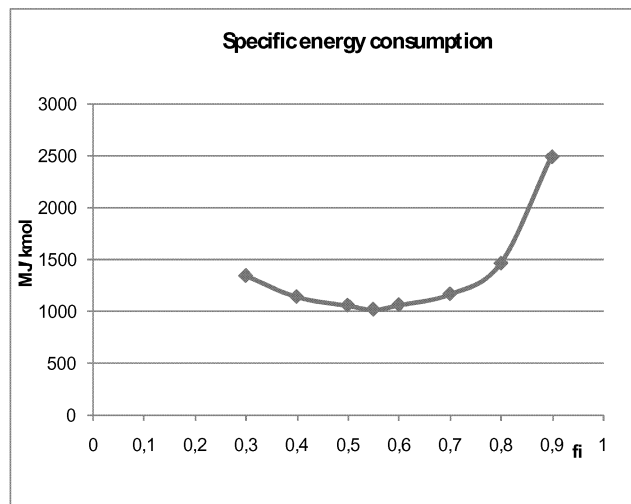


Fig. 6. Influence of the liquid division ratio on the specific energy consumption

Table 3. Most important results of the production for the optimal liquid division ratio

N-pentane recovery	%	86.5
Acetone recovery	%	79.5
N-pentane purity	mol %	98.65
Acetone purity	mol %	97.96
Total energy (SQ)	MJ	3517
Specific energy: SQ/(SW _A + SW _B)	MJ/mol	1019
Production time	min	62.5

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