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Distillation control structure selection for energy efficient operations

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Abstract

Previous methods on control structure selections in distillation columns are mainly concerned with processes operability such as control loop interactions, robustness to model uncertainty, and disturbance sensitivity. Few studies have been reported concerning the energy efficiency of various distillation column control structures. This paper details the choice of an energy efficient control configuration by incorporating thermodynamics second law in the selection criteria. In addition to using relative gain array for assessing control loop interactions, relative exergy array is used in assessing the energy efficiency of various control structures. The preferred control structure should have both good operability and good energy efficiency as distillation columns are the major energy consumer in the chemical industry. Detailed analysis of the performance of the control structures in the dynamic mode is presented. The proposed method is demonstrated on two binary distillation columns: methanol-water separation and benzene-toluene separation. Dynamic simulation results indicate that the proposed distillation control structure selection method is effective.

Keywords: exergy; relative exergy gain; control structures; distillation column control; energy efficiency.

1. Introduction

Distillation process is the most widely used separation process in chemical and petrochemical industries. Distillation accounts for about 95% for liquid separation and consumes 25-40% of the energy usage in the chemical and petrochemical industries and about 3% of the world energy usage [1, 2]. The quest for energy efficiency of chemical processes is still on because of the environmental, ecological and economic implications of energy usage. Distillation process is one of the prime targets in this quest because a little increase in the efficiency of a distillation column translates to a large reduction in its operation cost. For instance, it has been observed that a 10% energy saving in distillation column is equivalent to about 100,000 barrels of petroleum per day [3, 4].

There have been several studies aimed at improving the energy efficiency of distillation processes which has led to evolving distillation schemes different from the conventional ones. These include but

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not limited to vapour recompression, heat integrated distillation columns, and dividing wall columns [5-7]. The operation of distillation systems irrespective of whether conventional or heat integrated columns can be better improved with appropriate control schemes. Distillation has in fact been identified as the unit operation that could be significantly improved with good control with an estimate of about 15% reduction of energy if proper column control were in use [8]. One of the three issues identified by Skogestad and Morari [9] in the control of distillation columns is the control structure selection.

Relative gain array (RGA) is commonly used for the selection of the best control structure [10-13]. The steady state RGA however contains no information on the dynamic and disturbance on which distillation is hinged. This has led to the modifications of the RGA technique by different researchers to evolve techniques such as dynamic relative gain array (DRGA), effective relative gain array (ERGA), and relative normalized gain array (RNGA) [12-14].

However, despite all the modifications, control loop interaction analysis is no longer sufficient for the selection of the best control structure in the context of sustainable chemical industry. This is because of the cost and environmental implications of energy and the need to incorporate minimum energy usage in the selection of the control structure [15]. In this wise a number of tools based on application of thermodynamics in the process control regime have been developed. These include relative exergy array (REA), exergy ecoefficiency factor (EEF), and relative exergy destroyed array (REDA) [16-19]. These methods are all based on steady state and have not been validated in the dynamic state. This paper aims at using thermodynamics analysis in addressing the important issue of control structures selection for distillation columns. This is with a view of identifying the best energy efficient control strategies from a number of alternatives. This will then lead to an optimum control structure that will serve the dual purpose of achieving good product quality and minimum energy usage. The paper also presents a full detailed thermodynamic analysis of the control structures in the steady state with the aim of gaining insights into the exergy efficiency and exergy loss of each control structure. The viability of the selected control scheme in the steady state is further validated by the dynamic simulation in responses to various process disturbances and operating condition changes. This is to show the performance of the control structure in terms of composition control and energy efficiency.

The paper is organised as follows. Section 2 gives a brief overview of binary distillation systems and modelling. Section 3 presents distillation columns control structure selection based on relative exergy gain analysis. Simulations of two binary distillation columns, methanol-water separation and benzene-toluene separation, are used to demonstrate the proposed method. Results and discussions are given in Section 4. Finally, Section 5 draws some concluding remarks.

2. Binary distillation systems and modeling

2.1 Binary distillation columns

A simple binary distillation system is shown in Fig. 1. Methanol-water separation and benzene-toluene separation were considered in this study. The methanol-water system contains 50% methanol to be continuously rectified at 1 atm and at a rate of 4320kg h^{-1} to provide a distillate containing 99% methanol and a residue containing 1% methanol (by weight). The number of stages in the column is 16, and the feed is on the 4th tray with bottom up numbering.

Figure 1

For the benzene–toluene mixture, a continuous fractionating column is used to separate $30,000\text{ kg h}^{-1}$ of a mixture with 44% benzene and 56% toluene at 95°C into an overhead product containing 95% benzene and a bottom product containing 5% benzene at a pressure of 1 atmosphere and actual reflux ratio of 3.5. The relative volatility of the mixture is given as 2.5. The number of stages is 11 with the feed on the 5th stage numbering from bottom up.

2.2 Modeling of distillation columns

The mathematical model of a distillation column is an aggregation of individual theoretical stages. Fig. 2 shows a generic stage.

Figure 2

The dynamic state equations for a general stage are given below.

Component continuity equation:

$$\frac{d(M_j x_{j,i})}{dt} = L_{j+1} x_{j+1,i} + V_{j-1} y_{j-1,i} + F_j^l x_{j,i} + F_j^v y_{j,i} - L_j x_{j,i} - V_j y_{j,i} \quad (1)$$

where M_j is the liquid hold up in the j th stage, $x_{j,i}$ is the composition of i th component in the liquid at the j th stage, $y_{j,i}$ is the composition of i th component in the vapour at the j th stage, L_j is the liquid flow rate leaving from the j th stage, V_j is the vapour flow rate leaving from the j th stage, F_j^l is the liquid feed rate, F_j^v is the vapour feed rate, and t is time.

Total continuity equation:

$$\frac{d(M_j)}{dt} = L_{j+1} + V_{j-1} + F_j^l + F_j^v - L_j - V_j \quad (2)$$

Energy equation:

$$\frac{d(M_j h_j)}{dt} = L_{j+1} h_{j+1} + V_{j-1} H_{j-1} + F_j^l h_j + F_j^v H_j - L_j h_j - V_j H_j \quad (3)$$

where h_j is the liquid enthalpy from the j th stage and H_j is the vapour enthalpy from the j th stage.

Equilibrium equations:

$$y_{j,i} = K_{ji} x_{j,i} \quad (4)$$

$$x_{j,i} = \frac{l_{ji}}{L_j} \quad (5)$$

$$y_{j,i} = \frac{v_{ji}}{V_j} \quad (6)$$

where K_{ji} is the equilibrium constant for the i th component on the j th stage, l_{ji} is the liquid flow rate of the i th component from the j th stage, and v_{ji} is the vapour flow rate of the i th component from the j th stage.

Liquid summation equation:

$$L_j = \sum_{i=1}^C L_{ji} \quad (7)$$

where C is the total number of components and $C=2$ for the case of binary distillation.

Vapour summation equation:

$$V_j = \sum_{i=1}^C V_{ji} \quad (8)$$

Usually dynamic simulations of distillation systems are made with some assumptions to simplify the model [11].

If the specific liquid enthalpy is assumed to be a function of the specific heat capacity C_p , then

$$h_{L,j} = C_p(\Delta T) \quad (9)$$

The specific vapour is assumed to be

$$h_{V,j} = h_{L,j} + H_{vap} \quad (10)$$

where H_{vap} is the heat of vaporisation.

The liquid stream was modelled using linearised tray hydraulics incorporating activity coefficient equations.

3. Distillation control structure selection

Generally, variables that are needed to be controlled for a binary distillation column are composition of the distillate x_D , composition of the bottom product x_B , liquid level in the reflux drum, liquid level in the base drum and pressure in the column. Usually, the manipulated variables are reflux flow, L , reboiler vapour flow, V , distillate flow, D , bottom product flow, B , and condenser duty. Column pressure is usually controlled by the condenser duty and various distillation column control configurations refer to the pairing of other controlled and manipulated variables. Some typical distillation column control schemes include LV , DV , and LB control configurations.

3.1 RGA analysis

A multi-input multi-output (MIMO) system usually has interactions among the control loops. For better control of a process, control loop interactions should be minimised as a high degree of loop interaction makes the control difficult. Relative gain array (RGA) proposed by Bristol [20] is a tool that can be used to quantify control loop interactions. Relative gain is the ratio of the steady state gain when the loops are open to the steady state gain with all other loops closed.

The relative gain between the i th controlled variable and the j th manipulated variable is represented mathematically as

$$\alpha_{ij} = \frac{\left(\frac{\Delta y_i}{\Delta u_j}\right)_{all\ loops\ open}}{\left(\frac{\Delta y_i}{\Delta u_j}\right)_{all\ loops\ closed\ except\ the\ u_j\ loop}} \quad (11)$$

$$= \frac{open\ loop\ gain}{closed\ loop\ gain}$$

RGA is then obtained when the relative gains for all the pairing combinations in a multi-loop control system are calculated and put in an array.

$$RGA = \begin{bmatrix} \alpha_{11} & \cdots & \alpha_{1n} \\ \vdots & \ddots & \vdots \\ \alpha_{n1} & \cdots & \alpha_{nn} \end{bmatrix} \quad (12)$$

A relative gain of 1 on the diagonal of RGA indicates that there are no control loop interactions. The strategy then is to match the controlled and manipulated variables when α_{ii} is nearest to 1 and to avoid the pairings with close to zero or negative relative gains.

3.2 Thermodynamic analysis

Exergy is from a combination of the 1st and 2nd laws of thermodynamics. It is a key aspect of providing better understanding of the process and quantifying sources of inefficiency and distinguishing quality of energy used [21-23]. It is a tool for determining how energy efficient a process is [24, 25]. Exergy represents the part of energy, which can be converted into maximum useful work. It is used to establish criteria for the performance of engineering devices [26]. Unlike energy, exergy is not conserved and gets depleted due to irreversibilities in the processes [27]. The greater the extent of irreversibilities is, the greater the entropy production is.

The basis of the exergy concept was laid almost a century ago but was introduced as a tool for process analysis in the 1950s by Keenan and Rant [28]. Szargut et al. introduced the concept of chemical exergy and its associated reference states [29]. It is common to use ambient pressure and temperature as $P_0 = 101.325\text{kPa}$ and $T_0 = 298.15\text{K}$.

The total exergy of a stream is calculated as

$$Ex_{total} = Ex_{phy} + Ex_{chem} + Ex_{mixing} \quad (13)$$

$$Ex_{phy} = H - H_0 - T_0(S - S_0) \quad (14)$$

$$Ex_{phy} = \Delta H - T_0 \Delta S \quad (15)$$

$$\Delta Ex_{chem} = \sum n_i b_{chi} + RT_0 \sum n_i \ln a_i \quad (16)$$

For an ideal solution, the activity $a_i = \text{mole fraction}$. In the above equations, b_{chi} is the chemical exergy for component i , a_i is the activity coefficient of component i , H is the total enthalpy, S is the total entropy, T_0 is the reference temperature, H_0 and S_0 are enthalpy and entropy respectively measured at reference conditions.

For a heat source such as the reboiler, if Q_z is a heat source at an absolute temperature, T_z , and if T_0 is the ambient temperature, then the work equivalent of heat is given by

$$W_{\max} = \frac{(T_z - T_0)Q_z}{T_z} \quad (17)$$

This is the absolute theoretical maximum work recoverable. Eq. (17) is used in calculating the exergy of the reboiler.

Exergy efficiency of a system is calculated as

$$\varphi = \frac{\sum Ex_{out}}{\sum Ex_{in}} \quad (18)$$

While the exergy loss of a system is given as $I = \sum Ex_{in} - \sum Ex_{out}$.

For a binary distillation system the total exergy in and total exergy out are given as

$$Total\ Ex_{in} = Ex_{feed} + Ex_{Reboiler} + Ex_{Reflux} + Ex_{Boil\ up} \quad (19)$$

$$Total\ Ex_{out} = Ex_{Distillate} + Ex_{Bottoms} \quad (20)$$

In the above equations, Ex_{feed} , $Ex_{Reboilers}$, Ex_{Reflux} , Ex_{Boilup} , $Ex_{Distillate}$, and Ex_{Bottom} are, respectively, the exergy in the feed stream, reboiler, reflux stream, boil up stream, distillate product stream, and bottom product stream.

3.3 Relative exergy array

Relative exergy gain is defined as “the ratio of the gain change in the steady state exergy of the controlled stream with respect to that of the manipulated stream when all loops are open to the gain change in the steady state exergy of the controlled stream with respect to that of the manipulated stream when all other loops are closed and in perfect control” [19]. This is given in Eq. (21).

$$\beta_{ij} = \frac{\left(\frac{\Delta Ex(y_i)}{\Delta Ex(u_j)} \right)_{all\ loops\ open}}{\left(\frac{\Delta Ex(y_i)}{\Delta Ex(u_j)} \right)_{all\ loops\ closed\ except\ the\ u_j\ loop}} \quad (21)$$

Putting all the relative exergy gains in an array gives the relative exergy array:

$$REA = \begin{bmatrix} \beta_{11} & \cdots & \beta_{1n} \\ \vdots & \ddots & \vdots \\ \beta_{n1} & \cdots & \beta_{nn} \end{bmatrix} \quad (22)$$

REA is based on the RGA concept by replacing relative gain with relative exergy gain. The exergy gain ratio is usually calculated after a step input change in the manipulated variable. It gives the amount of exergy change in the controlled variable resulting from the exergy change in the manipulated variable and hence provides information on the thermodynamic efficiency of the pairing. This permits a good insight to the energy efficiency of a process right from the design stage and allows for the choice of optimum combination of loops.

REA indicates the exergy efficiency effects of pairing each of the manipulated variables to each of the controlled variables. It is defined analogous to the relative gain array. If the value of a relative exergy gain on the diagonal of REA is equal to 1, then it indicates the thermodynamic efficiency of the control loop under consideration is not affected by the other control loops [16, 18, 19]. This control loop pairing will be good in terms of thermodynamic efficiency. The value of a relative exergy gain greater than 1 implies that the exergy change from the open loop is much more pronounced. In this case, interaction from the variables in the process will decrease the process exergy change. The value of a relative exergy gain less than 1 indicates the exergy change due to open loop is less and hence an increase in exergy changes when the loops are closed. If the sign is negative, closing the control loop will improve the thermodynamic efficiency of the process but if on the other hand the sign is positive, this shows that the thermodynamic efficiency of the process will be decreased by the control loop. In control structure selection, a control loop pairing with relative exergy gain close to one is preferred.

4. Results and discussions

The methanol-water system was simulated from the fundamental first principle model in MATLAB while the benzene-toluene system was simulated using HYSYS. Three control configurations, LV, DV and LB, are considered for each system. Transfer function models are identified from the open loop step response data and are given in Tab. 1.

Table 1

The RGA and REA results from the steady state analysis are shown in Tabs. 2 and 3 for methanol-water and benzene-toluene systems respectively. Tabs. 4 and 5 show the open loop simulation results

under the three control configurations for the methanol-water system and benzene-toluene system respectively. For the methanol-water system, the RGA values obtained for all the considered control configurations are quite good. In terms of good control, any of the structures will be usable judging from RGA. If RGA value is greater than 0.5 but less than 1, this will be the preferred loop as it will minimise interaction [30]. Hence LB and DV will be good choices. For RGA greater than 1 as found in LV, higher controller gain will be required. This was confirmed in the closed loop dynamic simulation. The controller gain for LV is much higher than for the other structures. This however is not striking LV structure out as regards to good control. In terms of REA however, when the relative exergy gain is equal to 1, it is the preferred choice as the exergy efficiency is not affected by the control loop interactions [19]. For the three control structures considered, the relative exergy gain for the LB control structure is much closer to 1 than the other two control structures. The LB control structure will be the preferred choice with respect to thermodynamic efficiency. The steady state analysis of the control structures shows LB as the preferred control structure in terms of controllability and thermodynamic efficiency.

Table 2

In Tab. 3, the control structures for the benzene-toluene system show marked variations in terms of RGA and REA. The diagonal RGA values for LV control structure are less than zero and those for the LB control structure are much higher than 1. Negative diagonal elements in RGA indicate that closing the loop will change the sign of the effective gain. These structures may not be considered. RGA value for the DV control structure is greater than 0.5 but less than 1. The DV control structure therefore will be the preferred control structure. Considering the REA values, the LB and DV control structures could be chosen. However, though the LB control structure looks good for energy efficiency, its RGA value knocks it off if both controllability and energy efficiency are considered. The tool could aid in decision making and gives opportunity for consideration of design options. The steady state analysis of the benzene-toluene system shows DV as the control structure of choice.

Table 3

Table 4

The open loop simulation results for the two systems show some inconsistencies. For example, the exergy efficiency of LB structure for methanol-water system is not always the highest as predicted from the REA and RGA analysis. The same goes for DV structures in the benzene-toluene system. The overall decisions regarding a controller design should not be based on the steady state analysis alone [10]. There is a strong need for a detailed dynamic simulation analysis.

Table 5

In order to validate the steady state analysis results in the dynamic state, the closed loop response of each of the control structures to disturbances in the feed flow rate and changes in the setpoints of the distillate and bottom compositions were studied. PI controller was used on each of the control configuration. The controllers were tuned using Ziegler-Nichols tuning combined with the BLT tuning method [31]. The controllers were first tuned using the Ziegler-Nichols tuning as if they are for single input and single output systems without control loop interactions and then detuned using the BLT tuning method to account for control loop interactions. Fig. 3 shows the responses of the various control configurations to changes in distillate and bottom product setpoints for methanol-water system. The corresponding exergy analysis and reboiler energy usage are shown in Tabs. 6 and 7.

Figure 3

Close loop dynamic simulations were used to confirm the preferred choice from RGA and REA analysis. For the methanol-water system closed loop simulation results in Tab. 6, the exergy efficiency for the LB control structure is higher than those for the other two control structures except for increase in feed rate. And as expected, the exergy loss is lower under the LB control structure than under the other two control structures. This shows that the LB control structure is thermodynamically more efficient than the LV and DV control for the methanol-water system. The responses of all the control structures to setpoint changes further confirm the controllability of the structures and show that any of the structures could be used to bring about desired separation specification. To achieve it with the minimum usage of energy however, the LB control configuration will be the optimum choice.

Table 6

A dual composition control is used here because it yields less variation in downstream units and a more uniform quality of the final products. The large disparity in the exergy efficiency of the control structures in the open loop steady state to that in the closed loop as revealed in Tabs. 4 and 6 for methanol-water system is a result of the different composition specifications for the two cases due to setpoint changes in the closed loop response. This shows that high purity distillation is at a cost of energy. A cut in purity specification for example from 0.99 and 0.01 to 0.94 and 0.17 for top and bottom compositions respectively could result in as much as 30% more of exergy efficiency and a reduction in exergy loss.

Table 7

Also, the results in Tab. 6 for differing setpoint changes show the LB control configuration as more energy efficient for distillate setpoint changes and bottom product setpoint change. In addition, the LB control configuration favours an increase in feed disturbance in terms of exergy efficiency as compared to other control configurations. These observations reveal the need to incorporate thermodynamic analysis to aid the decision of energy efficient control configuration selection for distillation column operations. This will be a valuable tool in choosing control configuration for design and operation of distillation systems. Overall, the LB control configuration has a lower exergy loss and improved exergy efficiency than other control configurations. This information is quite revealing and shows the potential for bringing about energy efficient control operation of distillation processes. The reboiler exergy for each of the configuration at different variations considered also reveals the LB configuration as the structure with the least consumption of exergy.

Figure 4

For the benzene-toluene system, response to setpoint change in distillate and bottom compositions are shown in Fig. 4. The exergy efficiencies of the responses setpoint changes and changes in feed rate are shown in Tab. 7. For all the 4 cases of deviations from the nominal steady state considered, exergy efficiencies for the DV control structure are greater than those for the LV control structure and greater than those for two cases in the LB control structure. This trend follows that predicted from steady state REA analysis. Reboiler exergy differs from the reboiler energy because exergy analysis is a tool for assessing quality of energy and quantifying sources of inefficiency and recoverable energy in a system. Exergy analysis also takes into account entropy generation in a system and hence indicates “useful energy” of a system. The change in reboiler exergy per time at the closed loop simulation is shown in Figs. 5 to 8 for each of the control structures. It can be seen that the DV control structure overall has less reboiler exergy than the other control structures. A full detailed analysis of the performance of the control structure should be supplemented with a detailed dynamic analysis as presented. This will give a measure of confidence on a preferred control structure and as well quantifies its exergy consumption.

Figure 5

Figure 6

Figure 7

Figure 8

5. Conclusions

In this paper both RGA and REA are used in the selection of appropriate distillation control structures. The preferred distillation control structure should have good operability based on RGA analysis and high exergy efficiency based on REA analysis. The effectiveness of the method was demonstrated by dynamic simulation. It should be stressed here that the decisions regarding the controller should be based on the dynamic simulations in addition to the steady state REA and RGA analysis. The simulation results were found to confirm the control structure selection results. RGA as a tool for selecting control structure should be supplemented with REA to determine an energy efficient control structure. The tools when combined can help in choosing from various design alternatives. It will therefore aid in choosing an optimum structure right from the design stage. It could equally be effective tool in selecting optimum operations of distillation system. The tool as presented here is limited to the distillation unit. An overall energy analysis of the whole plant might be made to determine the effectiveness of the method on the plant as a whole.

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Symbols used

C : the total number of components

F_j : the feed rate to the j th stage

H_j : the vapour enthalpy from stage j

h_j : the liquid enthalpy from stage j

K_{ji} : the equilibrium constant for component i on the j th stage

L_j : the liquid rate from j th stage

l_{ji} : the liquid flow rate of i th component from j th stage

N : the total number of stages

V_j : the vapour rate from j th stage

v_{ji} : the vapour flow rate of i th component from j th stage

x_{ji} : the mole fraction of i th component in the liquid state on stage j

y_{ji} : the mole fraction of i th component in the vapour state on stage j

REFERENCES

- [1] F. E. Rush, *Chem. Eng. Prog.* **1980**, 76, 44-49.
- [2] G. Hewitt, J. Quarini, M. Morell, *The Chemical Engineer* **1999**, 690, 16-18.
- [3] G. Soave, J.A. Feliu, *Appl. Therm. Eng.* **2002**, 22, 889-896. DOI: 10.1016/S1359-4311(02)00006-6
- [4] R. E. Fitzmorris, R. S. H. MAH, *AIChE J.* **1980**, 26, 265-273. DOI: 10.1002/aic.690260209
- [5] A. A. Kiss, S. J. Flores Landaeta, C. A. Infante Ferreira, *Energy* **2012**, 47, 531-542. DOI:10.1016/j.energy.2012.09.038
- [6] Y. Kim, *Korean J. Chem. Eng.* **2012**, 29, 1680-1687. DOI: 10.1007/s11814-012-0106-0
- [7] H. Chun, Y. Kim, *Korean J. Chem. Eng.* **2013**, 30(7), 1473-1479. DOI: 10.1007/s11814-013-0067-y
- [8] S. R. Dartt, *Chem. Eng. Prog.* **1980**, 81, 11-14.
- [9] S. Skogestad, M. Morari, *AIChE J.* **1980**, 33(10), 1620-1635. DOI: 10.1002/aic.690331006
- [10] S. Skogestad, P. Lundström, E. W. Jacobsen, *AIChE J.* **1990**, 36(5), 753-764. DOI: 10.1002/aic.690360512
- [11] S. Skogestad, *Int. J. Model. Ident. Control* **1997**, 18(3), 177-217. DOI: 10.4173/mic.1997.3.1
- [12] Q. Xiong, W. J. Cai, M. J. He, *J. Process Control* **2005**, 15(7), 741-747. DOI: 10.1016/j.jprocont.2005.03.008
- [13] M. J. He, W. J. Cai, W. Ni, L. H. Xie, *J. Process Control* **2005**, 19(6), 1036-1042. DOI: 10.1016/j.jprocont.2009.01.004
- [14] T. Mc Avoy, Y. Arkun, R. Chen, D. Robinson, P. D. Schnelle, *Control Eng. Practice*, **2003**, 11(3), 907-914. DOI: 10.1016/S0967-0661(02)00207-1
- [15] M. T. Munir, W. Yu, B. R. Young, *UKACC Int. Conf. on Control*, Cardiff, UK, September **2012**.
- [16] M. T. Munir, W. Yu, B. R. Young, *Chem. Eng. Res. Des.* **2012**, 90(1), 110-118. DOI: 10.1016/j.cherd.2011.06.015.
- [17] M. T. Munir, W. Yu, B. R. Young, in *Proc. of the 8th Int. Symp. On Advanced Control of Chemical Processes* (Eds: V. Kariwala, L. Samavedham, R. D. Braatz) IFAC, Singapore **2012**, 8(1), 292-297. DOI: 10.3182/20120710-4-SG-2026.00032
- [18] M. T. Munir, W. Yu, B. R. Young, *The Canadian Journal of Chemical Engineering*, **2013**, 91(10), 1686-1694. DOI: 10.1002/cjce.21797
- [19] J. M. Montelongo-Luna, W. Y. Svrcek, B. R. Young, *The Canadian Journal of Chemical Engineering*, **2011**, 89(3), 545-549. DOI: 10.1002/cjce.20422
- [20] E. Bristol, *IEEE Trans. Autom. Control*, **1966**, 11(1), 133-134. DOI: 10.1109/TAC.1966.1098266.

- [21] H. Jin, M. Ishida, M.. Kobayashi, M. Nunokawa, *J. Energy Res. Technol.*, **1997**, 119(4), 250-256. DOI: 10.1115/1.2794998
- [22] M. A. Rosen. I. Dincer, *Int. J. Energy Res.* **1997**, 21, 643-654. DOI: 10.1002/(SICI)1099-114X(19970610)21:7<643::AID-ER284>3.0.CO;2-I
- [23] A. Doldersum, *Energy Convers. Manage.* **1998**, 39, 1781-1789. DOI: 10.1016/S0196-8904(98)00066-1
- [24] V. R. Dhole, B. Linnhoff, *Comput. Chem. Eng.* **1993**, 17, 549-560. DOI:10.1016/0098-1354(93)80043-M
- [25] Y. Demirel, *Sep. Sci. Technol.* **2004**, 39, 3897-3942. DOI: 10.1081/SS-200041152.
- [26] H. Asada, E. C.Boelman, *Build. Serv. Eng.Res. Technol.* **2004**, 25, 197-209. DOI: 10.1191/0143624404bt104oa
- [27] S. Sengupta, A. Datta, S. Duttagupta, *Int. J. Energy Res.* **2007**, 31(1), 14-28. DOI: 10.1002/er.1224.
- [28] A. P. Hinderink, F. P. J. M. Kerkhof, A. B. K. Lie, J. De Swaan Arons, H. J. Van Der Kooi, *Chem. Eng. Sci.*, 51(20), 4693-4700. DOI: 10.1016/0009-2509(96)00220-5
- [29] J. Szargut, D. R. Morris, F. R. Steward, *Exergy analysis of thermal, chemical and metallurgical processes* , Hemisphere Publishing Corporation, New York **1988**.
- [30] B. A. Ogunnaiké, W. H. RAY, *Process dynamics, modeling and control*, Oxford University press inc., New York **1994**.
- [31] W. L. Luyben, *Practical distillation control*, Van Nostrand Reinhold, New York **1992**.

Table 1. Transfer function models for different control structures

Control structures	methanol-water	benzene-toluene
<i>LV</i>	$G(s) = \begin{pmatrix} \frac{0.259}{1+5.21s} - \frac{1.109}{1+5.14s} \\ \frac{0.114}{1+5.31s} - \frac{2.196}{1+4.42s} \end{pmatrix}$	$G(s) = \begin{pmatrix} \frac{1.608}{1+1.1s} - \frac{9.9}{1+0.34s} \\ \frac{2.372}{1+3.3s} - \frac{2.5}{1+0.9s} \end{pmatrix}$
<i>DV</i>	$G(s) = \begin{pmatrix} \frac{-2.42}{1+3.61s} + \frac{1.36}{1+3.96s} \\ \frac{-1.19}{1+4.04s} - \frac{1}{1+3.03s} \end{pmatrix}$	$G(s) = \begin{pmatrix} \frac{-4.35}{1+1.42s} + \frac{8.49}{1+2.13s} \\ \frac{-1.03}{1+1.7s} - \frac{7.6}{1+1.67s} \end{pmatrix}$
<i>LB</i>	$G(s) = \begin{pmatrix} \frac{1.45}{1+3.88s} + \frac{1.08}{1+6.01s} \\ \frac{-1.1}{1+4.65s} + \frac{2.19}{1+3.63s} \end{pmatrix}$	$G(s) = \begin{pmatrix} \frac{1.34}{1+1s} + \frac{2.16}{1+3.35s} \\ \frac{1.07}{1+4.23s} + \frac{2.83}{1+5.6s} \end{pmatrix}$

Table 2. Results for RGA and REA analysis (methanol-water)

Control structures	RGA	REA
<i>LV</i>	$\begin{bmatrix} 1.2858 & -0.2858 \\ -0.2858 & 1.2858 \end{bmatrix}$	$\begin{bmatrix} 1.3056 & -0.3056 \\ -0.3056 & 1.3056 \end{bmatrix}$
<i>DV</i>	$\begin{bmatrix} 0.5992 & 0.4008 \\ 0.4008 & 0.5992 \end{bmatrix}$	$\begin{bmatrix} 0.6088 & 0.3912 \\ 0.3912 & 0.6088 \end{bmatrix}$
<i>LB</i>	$\begin{bmatrix} 0.7277 & 0.2723 \\ 0.2723 & 0.7277 \end{bmatrix}$	$\begin{bmatrix} 0.7331 & 0.2669 \\ 0.2669 & 0.7331 \end{bmatrix}$

Table 3. Results for RGA and REA analysis (benzene-toluene)

Control structures	RGA	REA
<i>LV</i>	$\begin{bmatrix} -0.2065 & 1.2065 \\ 1.2065 & -0.2065 \end{bmatrix}$	$\begin{bmatrix} 0.0007 & 0.9993 \\ 0.9993 & 0.0007 \end{bmatrix}$
<i>DV</i>	$\begin{bmatrix} 0.7908 & 0.2092 \\ 0.2092 & 0.7908 \end{bmatrix}$	$\begin{bmatrix} 0.8290 & 0.1710 \\ 0.1710 & 0.8290 \end{bmatrix}$
<i>LB</i>	$\begin{bmatrix} 2.5606 & -1.5606 \\ -1.5606 & 2.5606 \end{bmatrix}$	$\begin{bmatrix} 1.0349 & -0.0349 \\ -0.0349 & 1.0349 \end{bmatrix}$

Table 4. Open loop results for the methanol-water system

Control configuration and operating conditions	Exergy efficiency (%)	Exergy loss (kJ/hr)	Reboiler exergy (kJ/hr)	Reboiler energy (kJ/hr)	X_D	X_B
<i>LV</i>						
Steady state	74.6	1.2775×10^7	1.2632×10^4	2.9201×10^4	0.9266	0.1673
Step change in reflux rate	72.2	1.4278×10^7	1.2632×10^4	2.9201×10^4	0.9436	0.1748
Step change in reboil energy	77	1.1636×10^7	1.3604×10^4	3.1391×10^4	0.9023	0.1192
<i>DV</i>						
Steady state	74.04	1.3110×10^7	1.2632×10^4	2.9201×10^4	0.9303	0.1690
Step change in distillate rate	79	9.87×10^6	1.2660×10^4	2.9201×10^4	0.8962	0.1523
Step change in reboil energy	68.8	1.68×10^7	1.3554×10^4	3.1391×10^4	0.960	0.1462
<i>LB</i>						
Steady state	74.6	1.2760×10^7	1.2567×10^4	2.9047×10^4	0.9282	0.1706
Step change in reflux rate	72.8	1.39×10^7	1.2859×10^4	2.9740×10^4	0.9377	0.1633
Step change in bottom rate	72.7	1.36×10^7	1.1717×10^4	2.7116×10^4	0.9479	0.2107

Table 5. Open loop results for the benzene-toluene system

Control configurations and operating conditions	Exergy efficiency (%)	Exergy loss (kJ/hr)	Reboiler exergy (kJ/hr)	Reboiler energy (kJ/hr)	X_D	X_B
<i>LV</i>						
Steady state	47.75	1.573×10^7	9.7250×10^6	1.282×10^7	0.9500	0.0500
Step change in reflux rate	48.12	1.549×10^7	9.7250×10^6	1.282×10^7	0.9894	0.0106
Step change in reboil energy	46.29	1.750×10^7	1.065×10^7	1.3878×10^7	0.8548	0.1452
<i>DV</i>						
Steady state	47.75	1.573×10^7	9.7250×10^6	1.282×10^7	0.9500	0.0500
Step change in distillate rate	47.79	1.577×10^7	9.7250×10^6	1.282×10^7	0.9501	0.0499
Step change in reboil energy	46.63	1.734×10^7	1.053×10^7	1.387×10^7	0.959	0.041
<i>LB</i>						
Steady state	47.75	1.573×10^7	9.7250×10^6	1.282×10^7	0.9500	0.0500
Step change in reflux rate	47.69	1.601×10^7	9.958×10^6	1.313×10^7	0.9821	0.0179
Step change in bottom rate	48.81	1.463×10^7	9.287×10^6	1.225×10^7	0.9821	0.0179

Table 6. Closed loop simulation results for the methanol-water column

Control configurations and operating conditions	Exergy eff (%)	Exergy loss (kJ/hr)	Reboiler exergy (kJ/hr)	Reboiler energy (kJ/hr)
<i>LV</i>				
Nominal steady state ($X_D=0.99$; $X_B=0.01$)	37.83	6.8400×10^7	2.6014×10^4	6.0358×10^4
-0.5% change in feed	37.97	6.5424×10^7	2.4634×10^4	5.7161×10^4
+0.5% change in feed	38.36	7.0091×10^7	2.7297×10^4	6.3336×10^4
X_D at 0.95 and X_B at 0.01	45.16	4.9028×10^7	2.4202×10^4	5.6010×10^4
X_D at 0.90 and X_B at 0.005	32.81	8.9843×10^7	3.2418×10^4	7.4789×10^4
X_D at 0.95 and X_B at 0.005	30.38	1.0085×10^8	3.3105×10^4	7.6619×10^4
<i>DV</i>				
Nominal steady state ($X_D=0.99$; $X_B=0.01$)	37.99	6.8751×10^7	2.5968×10^4	6.0252×10^4
-0.5% change in feed	37.41	6.7100×10^7	2.4669×10^4	5.7239×10^4
+0.5% change in feed	38.52	7.0426×10^7	2.7266×10^4	6.3264×10^4
X_D at 0.95 and X_B at 0.01	44.39	5.0399×10^7	2.4278×10^4	5.6201×10^4
X_D at 0.90 and X_B at 0.005	32.26	9.3510×10^7	3.2518×10^4	7.5026×10^4
X_D at 0.95 and X_B at 0.005	30.06	1.0362×10^8	3.3014×10^4	7.6387×10^4
<i>LB</i>				
Nominal steady state ($X_D=0.99$; $X_B=0.01$)	41	5.5796×10^7	2.5269×10^4	5.9362×10^4
-0.5% change in feed	42.42	4.968×10^7	2.2974×10^4	5.3295×10^4
+0.5% change in feed	37.31	6.9320×10^7	2.9120×10^4	6.7550×10^4
X_D at 0.95 and X_B at 0.01	46.12	4.5805×10^7	2.3650×10^4	5.4723×10^4
X_D at 0.90 and X_B at 0.005	34.76	7.6494×10^7	3.1923×10^4	7.3629×10^4
X_D at 0.95 and X_B at 0.005	34.99	7.4488×10^7	3.0138×10^4	6.9727×10^4

Table 7. Closed loop simulation results for the benzene-toluene system

Control configuration	Exergy eff (%)	Exergy loss (kJ/hr)	Reboiler exergy (kJ/hr)	Reboiler energy (kJ/hr)
<i>LV</i>				
Nominal steady state ($X_D=0.95$; $X_B=0.05$)	47.75	1.573×10^7	9.723×10^6	1.282×10^7
-7.5% change in feed	47.18	1.484×10^7	9.108×10^6	1.182×10^7
+7.5% change in feed	47.89	1.681×10^7	1.062×10^7	1.377×10^7
X_D at 0.988 and X_B at 0.05	44.7	2.073×10^7	1.236×10^7	1.604×10^7
X_D at 0.95 and X_B at 0.01	40.58	3.245×10^7	1.807×10^7	2.329×10^7
<i>DV</i>				
Nominal steady state ($X_D=0.95$; $X_B=0.05$)	47.75	1.573×10^7	9.723×10^6	1.282×10^7
-7.5% change in feed	47.58	1.451×10^7	8.878×10^6	1.171×10^7
+7.5% change in feed	48.16	1.662×10^7	1.044×10^7	1.377×10^7
X_D at 0.988 and X_B at 0.05	44.74	2.098×10^7	1.219×10^7	1.604×10^7
X_D at 0.95 and X_B at 0.01	42.08	3.012×10^7	1.776×10^7	2.329×10^7
<i>LB</i>				
Nominal steady state ($X_D=0.95$; $X_B=0.05$)	47.75	1.573×10^7	9.723×10^6	1.282×10^7
-7.5% change in feed	47.44	1.469×10^7	8.961×10^6	1.182×10^7
+7.5% change in feed	48.15	1.664×10^7	1.045×10^7	1.377×10^7
X_D at 0.988 and X_B at 0.05	44.93	2.055×10^7	1.218×10^7	1.604×10^7
X_D at 0.95 and X_B at 0.01	40.82	3.215×10^7	1.776×10^7	2.329×10^7

Figure Captions:

Figure 1. Schematic diagram of a binary distillation system

Figure 2. Diagrammatic representation of an equilibrium stage

Figure 3. Responses to setpoint changes in top composition (a) and bottom composition (b) for the control structures in the methanol-water system. (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*; dashed-dotted lines: setpoints)

Figure 4. Responses to setpoint changes in top composition (a) and bottom composition (b) for the control structures in the benzene-toluene system. (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*; dashed-dotted lines: setpoints)

Figure 5. Reboiler exergy per time for change in distillate composition (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*)

Figure 6. Reboiler exergy per time for change in bottom composition (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*)

Figure 7. Reboiler exergy per time for increase in feed rate (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*)

Figure 8. Reboiler exergy per time for decrease in feed rate (Solid lines: *LV*; dotted lines: *DV*; dashed lines: *LB*)

Full Paper: An energy efficient distillation control structure selection method using relative exergy array is presented. Relative exergy array is used in assessing the energy efficiency of various control structures. The preferred control structure should have both good operability and good energy efficiency. The proposed method is demonstrated on two binary distillation columns: methanol-water separation and benzene-toluene separation.

Distillation control structure selection for energy efficient operations

F. Osuolale, J. Zhang*

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