MODELLING AND CONTROL OF REACTIVE DISTILLATION SYSTEMS

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Abstract: Reactive distillation (RD) is an integrated functionality of reactor and distillation, which is gradually becoming an important unit operation in chemical process industry. Process modelling is crucial because different applications have different requirements. In this work, the status of the current modelling techniques and their significance on process development are discussed. Strategies to tackle the control problems associated with complex dynamics are also discussed. RD for ETBE production is used as a case study. Set-point optimisation is presented to show the effects on chemical-physical interactions on profitability. Research gaps for future directions are identified.

Keywords: reactive distillation, modelling, control, optimisation

1. INTRODUCTION

Reactive distillation (RD) is gradually becoming an important unit operation in chemical process industry. This integrated operation is a favourable alternative to conventional series of reaction-separation processes for certain reactive systems. It offers reduction in both investment as well as operational costs. Review of the current applications and the potential benefits of RD for many potential systems can be found in (Doherty and Malone, 2001; Sharma and Mahajani, 2003).

The development of RD model has been recently reviewed (Taylor and Krishna, 2000). Like distillation systems, modelling of RD systems may employ equilibrium-stage and rate-based approaches with appropriate modifications and additions to reflect the reactions. The equilibrium-stage model has been used for designing and controlling different applications of RD columns. Recently, a rate-based approach is employed for especially investigating detail column design such as transport inside catalyst (Mohl et al., 2001) and effectiveness of catalyst configurations (Baur et al., 2001). Comparison of both models for several systems can be found elsewhere (Baur et al., 2000; Popken et al., 2001; Beckmann et al., 2002; Jimenez and Costa-Lopez, 2002). Although the rate-based model can represent physically closer to the real systems, it requires the estimation of more empirical models and appears to offer no improvement in accuracy for most systems. *The equilibrium-stage model with proper efficiency (for tray column) or HETP (for packed column) is still a preferable alternative model.*

The control of RD systems is challenging. The process objectives should be obtaining product purity specification (separation performance) while maximising reactant conversion (reaction performance). They are translated into achievable control objectives, which is then integrated into the overall control strategy. Control problems associated with multi-component distillation columns are practically expected in RD control. Control system design basically includes the selection of control configuration and control algorithm. Regarding control configuration, cascade inferential control scheme via measurable temperatures is reasonably adopted due to the lack of reliability of composition controls.

The concepts of established linear control were initially applied for RD systems. For continuous RD systems, general control considerations (Sneesby et al., 1997b) and dual composition and conversion control (Sneesby et al., 1999) have been presented for ethyl tert-butyl ether (ETBE) RD columns. For the same system, the comparison of a variety of one-point control schemes (Bisowarno and Tadé, 2002) and effectiveness of standard control schemes for single and double-feed RD columns (Al-Arfaj and Luyben, 2002b) have been investigated. Besides, linear control has also been implemented on RD systems for several applications (Kumar and Daoutidis, 1999; Al-Arfaj and Luyben, 2002a). *The* *linear controller with proper control configuration can be successfully implemented to avoid control problems associated with multiplicity while achieving the control objectives* on the RD systems.

Recently, nonlinear control is increasingly being investigated for RD systems. Model predictive control algorithm was implemented using reduced order nonlinear models for batch RD systems (Balasubramhanya and Doyle III, 2000). Nonlinear predictive control using neural network has been implemented on a semi-batch RD column for esterification (Engell and Fernholz, 2003). For continuous RD systems, a nonlinear input-output linearizing controller and nonlinear controller (Kumar and Daoutidis, 1999) and a robust PI control scheme (Loperena et al., 2000) have been designed for ethylene glycol system. Analysis of different control configurations and application of nonlinear control strategies have been performed for an ethyl acetate system (Vora and Daoutidis, 2001). Asypmtotically exact input/output-linearisation employing an observer has also been recently investigated (Gruner et al., 2003). Pattern-based predictive control has also been proposed for controlling the product composition (Tian et al., 2003). *The growing nonlinear control applications will certainly continue due to the nonlinear nature of RD systems for higher product competitiveness, tighter safety and environmental regulations.*

This paper aims to review the current progress on modelling and control of RD systems. Set-point optimisation is conducted to show the effects of chemical-physical interactions on profitability. ETBE RD system is used as a case study to emphasise the significance of the discussion. The research gaps and the key contributions in the area are shown in italics within the text of this paper.

2. REACTIVE DISTILLATION

RD integrates the conventional functionality of a mixed-continuous reactor and a distillation process. It was initially considered for homogeneous self-catalysed reactions such as esterification and hydrolysis for several applications employing a homogeneous catalyst (Krishna, 2002). RD, which uses a heterogeneous catalysts known as catalytic distillation, were firstly considered for RD in (Spes, 1966), but it then remained uninvestigated and lacked research interests until the 1980s. RD may also be categorized as hybrid and non-hybrid columns (Güttinger and Morari, 1999a; Güttinger and Morari, 1999b). Hybrid RD is used to describe columns, which have separated reactive and separation sections, while the reaction takes place in the whole non-hybrid RD column. *Recently, RD is accepted to describe both self-catalysed reaction and such systems where either homogeneous or heterogeneous catalyst is used to accelerate the reaction.*

2.1 Modelling

Design and control of the complex behaviour of the RD systems requires high-quality mathematical process models. The model has to be revised repeatedly so that it can represent complex phenomena of the system. Two approaches for modelling distillation systems are equilibrium-stage model and non-equilibrium or rate-based approaches. Both of these approaches can be applied to RD system with appropriate modifications and additions to reflect the reactions.

For the equilibrium-stage approach, the main assumption is that the vapour and liquid streams leaving the stage are assumed in phase equilibrium. An efficiency factor (for tray column) or HETP (for packed column) is introduced in the model to closer represent the real systems. However, there are no available methods for estimating either efficiency or HETP in RD due to the influence of reactions on the component efficiencies (Taylor and Krishna, 2000). This limitation leads to the significance of the rate-based model.

The rate-based model follows the philosophy of the rate-based model of a particular distillation column (Krishnamurthy and Taylor, 1985) and rigorous Maxwell-Stefan theory to calculate the interphase mass and heat transfer rates (Krishna and Wesselingh, 1997). The mass transfer rates are calculated directly from fundamental mass transfer models. As a result, physical properties such as surface tension, diffusion coefficient, viscosities, etc. should be specified to calculate the mass and heat transfer coefficients and inter facial areas. Thermodynamic properties are required not only for calculation of phase equilibrium but also for calculation of the driving force for mass transfer (Baur et al., 2000). The effect of nonideal component behaviour should also be considered to calculate the reaction rates and the chemical equilibrium coefficients. The kinetic model of the reaction should be known and the effect of the reaction on the interphase mass transfer rate may need to be considered (Higler et al., 1999).

The selection of the model is crucial as it should be accurate enough to represent the system complexity and feasible for real-time monitoring and control. Different applications also need different requirements for process modelling. For example, rate-based model that is suitable for detail design of the column may not be suitable for real-time control of RD operation. No rate-based model applied for control system design is reported in the literature. On the other hand, simplified (reduced) models that are used for controller design may not be suitable for optimisation and dynamics analysis. Models for dynamics analysis will be different from those for real-time control, because in the latter case timeliness is also a key factor for the system. *Most publications on the modelling of RD systems have not mentioned the purpose and usage of the models.*

It is worth noting that experimental validation is another critical issue that needs to be addressed in the modelling of RD systems. The experimental data can serve for checking the feasibility study, system analysis and design, process modelling, control development and implementation. This important role of the validation has been largely missing in the open literature. *Most publications in the area of modelling and control have been mainly based on computer simulations. The results are largely devalued without the support of (industrial) experimental data.*

In the case study, the significant aspects of modelling of ETBE RD systems involving single column, single column integrated with side reactors, and the whole flow sheet production route, respectively, are addressed as exemplary illustration.

2.2. Control

Control of batch reactive distillation can be found elsewhere (Balasubramhanya and Doyle III, 2000; Fernholz et al., 2000), this section focuses on control of continuous RD systems. Generally, exact model-based control systems do not always work in process industry due to unmodelled process dynamics and uncertain disturbances. Therefore, the control system design should be able to compensate for the process uncertainties. Two aspects of control system design are control configuration and control algorithm.

For the control configuration, it includes selection and pairing of the manipulated and measurable controlled variables. Like distillation columns, RD is a multivariable process control system in which pressure and inventory controls can be treated the same as that of distillation control. However, product compositions (or, product composition and conversion) controls for two products RD columns are rather different problems. Inferential control schemes via measurable temperatures are generally preferred because of mainly unreliability of the composition analysers. However, the relationship of the composition and temperatures are frequently non-unique in RD systems (Sneesby et al., 1998; Sneesby et al., 1999) as shown by the presence of multiplicity phenomena. Fortunately, proper selection of the control configuration can avoid problems associated with the input multiplicity.

For the control algorithm, some degree of intelligence in the control algorithms is required to compensate for the process uncertainties and unmodelled process dynamics. Therefore, model-based advanced controller such as adaptive PID, gain-scheduling, multi-model, pattern-based control, fuzzy logic, etc, are promising tools for RD control. Recent published papers have shown that more advanced controllers outperform that of standard linear controller with respect to certain control criteria (Engell and Fernholz, 2003; Gruner et al., 2003; Tian et al., 2003).

Implementation of the advanced controller requires suitable models and hardware including a computer for real-time simulation linked to the control system. Control implementations of the advanced control algorithms have been also missing in the open literature. In practice, standard PI controller may suffice with proper selection of the control configuration (Al-Arfaj and Luyben, 2000). Selection of a suitable control configuration is frequently dominating the control system design of RD systems. The complex RD systems combined with higher product competitiveness, tighter safety and environment regulations will increasingly demand advanced control algorithms.

The present control studies in RD systems have mostly discussed simple column configurations. A complete production route via RD systems mostly involves pre-reactor, downstream separations, or side reactor integrated to the RD column. *The plant wide control of such complete production route is yet to be investigated*. Another critical process development is the scale-up of the RD systems. Scale-up methods, which can guarantee both reaction and separation performances over a broad range of reaction regimes on industrial RD systems, are not readily available (Tuchlenski et al., 2001; Schoenmakers and Bessling, 2003). *Simulation work using a rigorous mathematical process model, which is supported by essential experimental data, can serve as a crucial basis for scaling-up processes*.

In the following case study, some advanced control algorithms for ETBE RD system are discussed to illustrate some of the problems encountered in this area as well as potential solution to these problems.

3. CASE STUDY: ETBE SYSTEM

The complete flow sheet for ETBE production employing a single RD column is depicted in Figure 1 (Sneesby et al., 1997a). It consists of a pre-reactor and a RD column, which may be integrated with side reactors. Therefore, the feed stream of the column is a mixture of isobutylene, ethanol, ETBE, and n-butane, resulting from a pre-reactor, which converts most of isobutylene to ETBE.

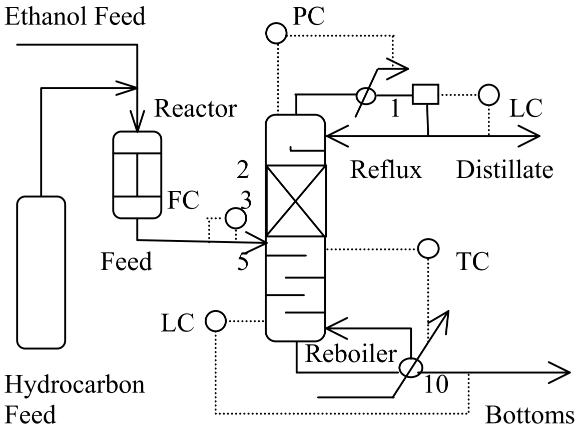


Fig. 1 ETBE RD System with the controllers

Table 1 ETBE RD column characteristics and Inputs

Column Specifications:		
$N_{RE}/N_{RX}/N_{ST}$	3/3/5	
Feed stage	6	
Overhead pressure	950 kPa	
Feed Conditions:	**Range**	**Nominal***
Temperature	30°C	30°C
Rate (l/min)	0.684-0.836	0.76
Comp. (mol)	70–80 mol% conversion in the pre-reactor	0.291 ETBE, 0.091 EtOH, 0.073 iBut, 0.545 nBut
Man. Variables:	**Range**	**Nominal***
L_R (l/min)	2.0 – 2.4	2.2
Q_R (MJ/min)	0.4825-0.555	0.520

* Nominal (optimum) operating condition for designing the control system

3.1 Single Column

A pilot scale packed RD column for ETBE production serves as an example for a typical single-feed two-products RD system. The packed column consists of 25 cm rectifying section, 75 cm reactive section, 100 cm stripping section, a total condenser, and an electric partial reboiler, respectively, as shown in Figure 1. The primary and secondary manipulated variables are reboiler duty (Q_R) and reflux rate (L_R), respectively. LV control scheme, which outperforms other control schemes for this column (Bisowarno and Tadé, 2002), is employed. Typical operating data including the operating range are summarised in Table 1.

Equilibrium-stage approach has been applied to model the RD column. Validation with steady state experimental data shows that the model is adequate with proper HETP values of each section as shown in Table 2. Although the HETP cannot be determined *apriori*, M1 - M4 are the models that employ different HETP for each section. M4 result in temperature profile, isobutylene conversion, ETBE

Table 2 Effect of HETP on equilibrium-stage model

Properties	M-1	M-2	M-3	M-4	Exp.
T1 (°C)	-	-	-	-	40
T2 (°C)	69.3	69.4	69.4	69.4	74
T3 (°C)	69.6	70.1	69.7	69.9	77
T4 (°C)	70.4	71.6	71.9	73.1	80
T5 (°C)	71.7	93.5	98.2	110.5	104
T6 (°C)	100.9	148.2	145.9	151.3	155
Conv. mol%	58.7	85.8	88.2	91.3	93.4
X_{ETBE}, wt%	54.4	91.0	90.7	93.7	93
Q_R (kW)	7.50	7.57	7.54	7.55	7.24
Qc (kW)	7.12	7.18	7.16	7.17	na

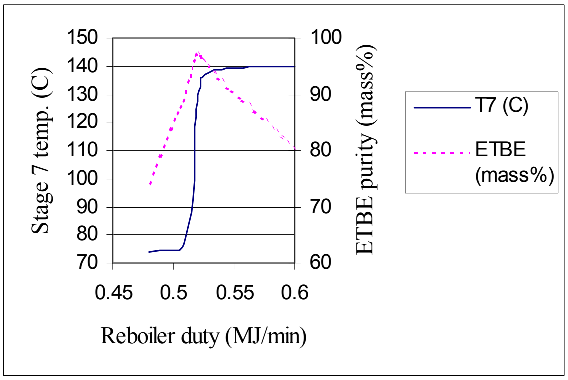
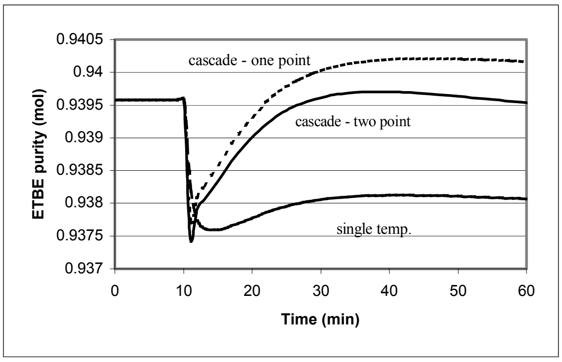


Fig. 2 stage 7 temp./purity vs. Q_R (L_R = 2.2 L/min)

purity, which are reasonably comparable to that of the experimental data. Similar reports have shown the adequacy of the equilibrium-stage model for other applications (Luo and Xiao, 2001; Popken et al., 2001; Jimenez and Costa-Lopez, 2002). The results also indicate the significance of HETP in the modelling of RD column using the equilibrium-stage approach. *Therefore, a method, which can reliably estimate the HETP, is required for obtaining a more rigorous model.*

The relationship between the ETBE purity and the reboiler duty reveals input multiplicity phenomena as shown in Figure 2. Based on the sensitivity analysis, stage 7 temperature is found to be the most appropriate measured variable to infer the ETBE purity (Sneesby et al., 1997b) so that the problems associated with input multiplicity can be avoided. The choice of the stage 7 temperature is also justified via dynamic simulations (Sneesby et al., 1999). Figure 2 also shows the relationship between the stage 7 temperature and the reboiler duty. Hence, inferential control is adopted to control the ETBE purity. The nonlinear process gain ($\Delta T_7/\Delta Q_r$) is large around the nominal operating condition and becomes small outside this range. This inferential control scheme can be extended to directly or indirectly infer the reactant conversion (Sneesby et al., 2000; Tadé and Tian, 2000).

The main objective of the control system is to keep the controlled stage 7 temperature close to the set-



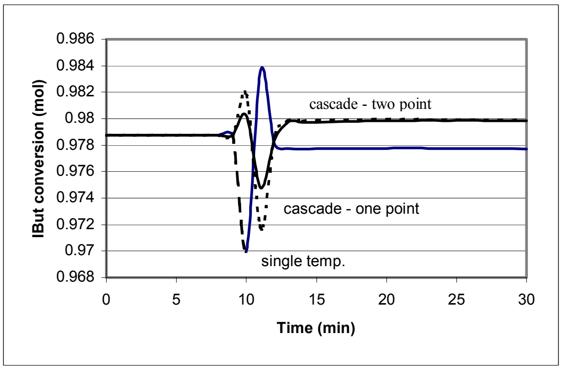


Fig. 3 Rejection of +10% step change in the feed rate

points despite the presence of disturbances. The most significant disturbances are changes in the feed flow rate and in the feed composition. The second objective is a sufficiently fast set-point tracking. These two objectives must be achieved for the entire operating range of the RD column.

Multivariable control using inferential multiple temperatures of an ETBE RD column is used to demonstrate the significance of considering both conversion and composition control. Using steady-state least square regression, the multiple temperature measurements to infer the ETBE purity and isobutylene conversion is shown in equation 1. The available temperatures, which are top column (T_2), top reactive section (T_3), bottom reactive section (T_5), mid-stripping section (T_7) and reboiler temperature (T_{10}), are used for inferential control to avoid several problems associated with single composition control such as effects of noise, pressure and nonkey components variation (Mejdell and Skogestad, 1993).

$$\text{Conversion} = 0.745648 + 0.011224*T_2 - 0.00997*T_3 - 0.00142*T_5 + 0.001119*T_7 + 0.000815*T_{10} \quad (1a)$$

$$\text{Purity} = 0.21096 - 0.00204*T_2 - 0.00589*T_3 + 0.000834*T_5 - 0.00057*T_7 + 0.0082*T_{10} \quad (1b)$$

Standard PI controllers are employed and the effectiveness of the proposed control configuration is compared to that of one-point (purity) control configurations. Fig. 3 shows that *better control scheme (cascade control using multiple inferential temperatures) has better rejection ability than the standard inferential single temperature control scheme.*

A good-rigorous RD model is mostly too complex for control design. *A pattern-based predictive control (PPC) scheme incorporating standard PI controller has been developed to alleviate the model requirement* (Tian et al., 2003). This control scheme consists of two main parts: a nonlinear transformation and a pattern-based predictor. The PPC system outperforms the standard PI control.

However, implementation of the control algorithms on an industrial system or on experimental rig is yet to be investigated. Extra time delay, which is a common and difficult problem in control, may be introduced due to improper implementation. *There is no publication on implementation of control systems on RD systems.*

3.2 Single column with Side Reactors

The application of industrial RD is still limited for certain reactive systems, mainly etherification (MTBE), esterification (methyl acetate), and alkylation (ethylbenzene or cumene)(Tuchlenski et al., 2001). Detail economic comparison reveals that significantly simpler flowsheet of the RD technology is offset by the higher cost of the RD column compared to that of conventional technology for toluene disproportionation system (Stitt, 2002) and butyl acetate system (Jimenez and Costa-Lopez, 2002). Successful commercialisation of RD technology requires specific hardware designs, which seldom correspond to those of conventional distillation. For example, high liquid hold up in reactive section, which is required for maximising conversion, does not agree with the requirement of high interfacial area for good separation (Krishna, 2002).

The concept of side reactor is introduced to overcome hardware design limitations(Jakobsson et al., 2001). The side reactor has potential to reduce the requirement of catalyst loading in the reactive section. As a result, shorter reactive section may be employed, which may lead to the reduction of column cost if less amount of catalyst forces a decrease in the diameter or height of the column. The side reactor can also be treated the same as the pre-reactor, which is therefore more convenient for shut down operation and catalyst replacement. *For the process modelling, equilibrium-stage model of the RD column and stirred tank reactor of the side reactor are employed. A series of these reactors can be used to model a tubular or a fixed bed reactor.*

Table 3 show the column specifications and simulation results of the three RD columns considered (Bisowarno et al., 2003). The SR-1 design employs side reactor and has fewer reactive stages

Table 3 ETBE RD column with side reactor

Property		B	SR-1	SR-2
RD Column:				
$N_{RE}/N_{RX}/N_{ST}$		7/7/14	7/4/14	7/4/14
Condenser temp, C		55.6	55.7	55.8
Reactive section temp, C		56.6-90.4	57.2-73.6	57.2-76.5
Reboiler temp., C		143.9	140.9	143.8
Bottom comp.	EtOH	4.75	7.56	4.92
Mol%	ETBE	*95.22*	92.40	95.01
	DIB	0.02	0.03	0.07
Distillate	EtOH	0.02	0.08	0.07
comp. mol%	iBut	0.03	0.07	0.12
	nBut	99.94	99.85	99.80
iBut conversion, mol%		99.91	99.82	99.67
Condenser duty, kW		2667	2670	2671
Reboiler duty, kW		3082	3024	**2798**
Side Reactor:				
iBut conv. mol%		NA	-61.8	-47.2
Input comp.	EtOH	NA	6.47	7.68
Mol%	iBut		0.06	5.91
	nBut		93.05	62.81
	ETBE		0.42	23.60
Input rate, kmol/hr			37.73	185.9
Output comp.	EtOH	NA	6.71	17.98
Mol%	iBut		0.30	4.10
	nBut		92.81	59.95
	ETBE		0.16	11.87
	DIB		0.01	**6.10**
Output rate, kmol.hr			37.83	194.9

than that of the basic column (B). The SR-2 design has the same stages as the SR-1 but the feed stream from the pre-reactor is fed to the side reactor instead of at the bottom of the reactive section. Table 3 shows that the side reactor can be integrated into the RD column to reduce the number of reactive stages. However, modification of the basic RD configuration should be done to produce the same ETBE purity and isobutylene conversion as that of the basic RD design. The effluent of the pre-reactor is fed to the side reactor instead of at the bottom of the reactive section.

Although the experimental validation is yet to be conducted, the results show the feasibility of side reactors (integrated with the RD column) to potentially reduce the capital costs. Control analysis of this RD system involving a single column integrated with side reactors is yet to be investigated.

3.3 Flow sheet production

A RD system for ETBE production route consists of a pre-isothermal reactor and a RD column with or without side reactors. Analysis and design of the plant wide control is yet to be conducted. There are only a few published literature about design and control analysis of the production route using RD column *(Luo and Xiao, 2001).*

Table 4 Relative Values

Relative values	Feed	Distillate	Bottoms
EtOH (\$/tonne)	-	-250	450
IBut (\$/tonne)	-	150	120
ETBE (\$/tonne)	-	0	800
Inert-C4 (\$/tonne)	-	150	120
Overall (\$/tonne)	250	-	-
Energy costs			
Heating (\$/kW)	0.03		
Cooling (\$/kW)	0.02		

3.4 Optimisation

RD operation is usually justified on steady state process results, which may not be realisable in dynamic operation, especially if regular disturbances occur. Besides, the operating conditions, which maximise reactant conversion, do not coincide with the conditions to maximise the product purity. An alternative solution is to maximise an objective function, which is based on economic considerations and depends on both parameters (purity and conversion). Single RD column for ETBE production is used to demonstrate the optimisation using rigorous-dynamic simulations in what follows. This column is the same as the basic column (B) in the previous section, which consists of 7 rectifying stages, 7 reactive stages, and 14 stripping stages, respectively.

The objective function should reflect the profitability of the process. The relative value of each component in the feed, bottom product, distillate product, and value of energy is shown in Table 4. The values are given in dollars but the units and magnitudes are chosen arbitrarily. The high negative value of ethanol in distillate reflects the potential to poison downstream catalysts. The reduced value of C4 components in the bottom product reflects the effect on gasoline pool volatility. The objective function for optimisation does not consider the fixed operating costs and it can be expressed as shown in equation 2.

$$\text{Profitability (P)} = \text{Product value} - \text{Feed Cost} - \text{Energy Cost} \quad (2)$$

Based on the values shown in Table 4, the profitability can be derived as shown in equation 3.

$$P = (450\ x_{EtOH,B} + 120\ x_{iBut,B} + 800\ x_{ETBE,B} + 120\ x_{inert,B})\ B + (-250\ x_{EtOH,D} + 150\ x_{iBut,D} + 120\ x_{inert,D})\ D - 250\ F - 30\ Q_R - 20\ Qc \quad (3)$$

The RD column can only operate within the constraints imposed by the equipment design such as column capacity (flooding), maximum duties of the reboiler and condenser. Restrictions on the

Table 5 Simulation results of Optimisation

	Maximum Profitability	Maximum ETBE Purity
Set-points		
Top pressure (kPa)	650	725
Stage 7 temp. (oC)	104	113
Reflux rate (l/min)	17.5	21
Key Results		
ETBE purity (wt%)	95.1	95.7
iBut conv. (mol%)	94.6	95.4
Profitability ($/m3)	1744	1430
Constraints		
Q_R (MW	5.10	5.64
Qc (MW)	*3.90*	4.35
Flooding factor (%)	74	*80*
$X_{EtOH,D}$ (wt%)	0.18	0.38
C4 in bottom (wt%)	0.45	0.36

composition products due to downstream processing and blending requirements should be included. The chosen constraints are shown below.

$Q_R < 6$ MW (4a)
$Qc < UA(Tc – Ta)$ (4b)
Vapour flooding factor < 80% (4c)
Downcomer flooding factor < 80% (4d)
$x_{EtOH,D} < 0.025$ (4e)
$x_{iBut,B} + x_{inert,B} < 0.020$ (4f)
$x_{ETBE,B} > 0.900$ (4g)

Although the primary manipulated variables can be optimised directly, the set-point optimisation is preferable. It can then be used within the existing control structure to reject disturbances. The multivariable optimisation was undertaken within Aspen CM using a feasible-path successive quadratic programming routine. The steady state solutions of the optimisation are shown in Table 5.

Table 5 show that the maximum profitability can be achieved not at the maximum ETBE purity. At the maximum profitability, the condenser duty constraint is active. This result indicates that the lower overhead pressure, which consequently condenses the overhead vapour at lower temperature, increases the profitability. However, the chemical equilibrium assumption is not satisfied at lower pressure, which results in slower reaction rate and lower purity. The maximum unconstrained profitability is achieved at the overhead pressure of 500 kPa. On the other hand, increasing the overhead pressure can ease the condenser limit and increases the ETBE purity, but introduces a flooding limit.

This optimisation framework, which employs controller set-point that can be updated automatically or manually, can be used to develop a supervisory control system for the RD column.

4. CONCLUSIONS

The current status of modelling of RD systems has been discussed. The equilibrium-stage approach is still an interesting alternative. However, a method, which can reliably estimate efficiency or HETP, is required for obtaining a more rigorous model. Stirred tank reactor may be used to model side reactor, which can be integrated with the RD column. A series of these reactors can be used to model a tubular or a fixed bed reactor.

Regarding control of RD systems, selection of control schemes including the measurable inferential temperatures has crucial role on the overall control performance. To alleviate the complex model requirement, a pattern-based predictive control has been proposed. Further work need to be done in this area.

The optimisation shows that maximum profitability does not coincide with maximum product purity. This set-point optimisation framework can be extended for supervisory control scheme for RD system.

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