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Combined Internal Model and Inferential Control (CIMIC) for *n*-butane/*i*-butane distillation column

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ABSTRACT

Distillation column control has been studied extensively over the last half century and application of model based controller (MBC) is among that shows potential to be implemented. To date, there are many variations of MBC that had been developed using different techniques. One of them is Combined Internal Model and Inferential Control (CIMIC). CIMIC is a modified version of Internal Model Control (IMC) which utilizes inferential technique in its control scheme. In this study, CIMIC is applied to control an industrial n-butane/i-butane distillation column. The distillation column is simulated using Aspen Plus and validated based on data available from literature. Based on the Aspen model, a linear model of the distillation column is developed and used as process model inside CIMIC. CIMIC control scheme employ the conventional two degree of freedom (2DOF) IMC control scheme with an additional control loop used as inferential to the primary loop. In order to evaluate CIMIC performance, original control schemes i.e. IMC and 2DOF IMC are also being tested as comparison. Based on the tests, CIMIC has demonstrated an improved performance compared to the mentioned control schemes in set point tracking (>30% better), input disturbance (>90% better) and output disturbance (>14% better) rejection tests.

INTRODUCTION

Distillation is a common and important process in Chemical Process Industries (CPI) and refinery plants. It is estimated that 95% of the separation processes in the CPI and refinery plants in the world use the distillation process (Osuolale and Zhang, 2015). In general, distillation is a method to separate a mixture of components of a liquid solution based on the distribution of these components between the liquid and vapour phase. During the distillation process, all the components are present in both phases and are separated based on their boiling points. The distillation operation demands a high usage of energy and operation cost. Díez *et al*. (2009) reported that 60% of the energy used in the chemical industry is consumed by the distillation process alone and Cheremisinoff (2000) stated that the distillation operation cost can contribute up to 50% of the total plant operation cost.

Since the distillation process is a widely used and a costly separation process in the CPI, thus it is important to be properly controlled. Distillation control is a challenging process due to inherently nonlinear of the process, multivariable interaction, non-stationary behaviour and severity of disturbances. The nonlinearity in the distillation process can be caused by many factors such as the non-ideal Vapour Liquid Equilibrium (VLE) of the mixture, the complex column design, the dissimilar materials and heat balance at each tray and the operation problems such as tray wetting and fouling. The multivariable interaction generally happens when both top and bottom products are strictly controlled. The non-stationary behaviour could occur due to changes in tray

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efficiency. Major disturbances such as feed flow rate or composition upsets, subcooled reflux changes, loss in reboiler steam pressure and column pressure upset are usually distressing the distillation column control (Riggs, 2001). Furthermore, acquiring higher product purity (more than 98%) can cause the system to become highly nonlinear. The distillation column is also reported to exhibit multiplicity behaviour in its process (Zheng *et al.*, 1998). Due to such attributes in the distillation process, the controller task in the distillation control scheme has become more complicated and indispensable.

One of the conventional and generally used control schemes in distillation control is Proportional Integral Derivative (PID) controller. Rhinehart *et al*. (2011) estimated that about 95% of all control scheme applications in the CPI are based on PID controller. PID controller is commonly preferred due to its simple design, easy implementation and reliability in linear operating conditions. However, due to the nonlinearity in the distillation system behaviour, the effectiveness of the PID controller is often degraded when achieving its desired output during certain operating conditions and constraints handling (Dutta and Rhinehart, 1999). Gokhale *et al*. (1995) have shown that the conventional PI controller is incapable of properly handling the variation of disturbance in the feed stream.

In order to overcome the conventional basic controller shortcomings, Advanced Process Control (APC) have been proposed for the distillation control (Gokhale *et al*., 1995). The distinctive feature of APC is the application of a computerized system such as a computer based controller, software sensor (soft sensor) and a process model in the controller scheme. This can give a considerable advantage in controlling the nonlinearity process and handling disturbance in the distillation column. Among the APC schemes, Model Based Control (MBC) strategy has attracted a great deal of interest from process control practitioners. The MBC is preferred due to its generic control strategy with a wide selection of process models and controller designs.

One of the well-known MBC techniques is the Internal Model Control (IMC). The IMC scheme appears as an improvement to the traditional feedback control algorithm by utilizing a process model explicitly in the controller design. Compared with other MBCs, IMC provides a more straightforward design and transparent framework for the control scheme development (Saxena and Hote, 2012). In addition, IMC strategy is practised in many CPI processes such as CSTR (Manimozhi and Meenakshi, 2016), combustion (Awais, 2005), batch reactor (Mujtaba *et al*., 2006), pH neutralization (Petchinathan *et al*., 2014) and distillation column (Jin *et al*., 2013).

In this study, a modified IMC which combine IMC and inferential control technique is proposed. Combined Internal Model and Inferential Control (CIMIC) technique was originally proposed by Häggblom (1996) to ensure a stable operation by rejecting the disturbance in primary variable using the inferential technique. However, the performance of former CIMIC is not as expected. The main reason for this is the selection of primary variable (i.e. tray temperature) that respond almost the same time with secondary variable (i.e. product flow rate). This had diminished the inferential capability and made the controllers counteract with each other. In this current CIMIC, the primary variable is the product composition which generally associate with certain delays. The disturbance in the distillation process is inferred from the secondary variable which is, in this case, the column's most responsive tray temperature. Theoretically, the current CIMIC technique is expected to perform well since the tray temperature is more sensitive (compared to product flow rate) to changes in the distillation process and can be measured instantaneously when compared to product composition. In order to demonstrate the inferential advantage and reduce the controllers' interaction, 10 minute of sampling time is set for the primary variable and 1 minute of sampling time is set for the secondary variable. Moreover, the current CIMIC concept is revisited with a new approach and simpler control scheme compared to the previous one.

Methodology:

Overview:

An overview of the research methodology used in this research is shown in Fig. **1**. Based on the figure, in the early development stage, information regarding the industrial distillation column is gathered and simulated using Aspen Plus software. Then, the simulation model is validated based on the available data (Ilme *et al*., 2001). Next, a linear model of the distillation column is developed using Matlab system identification software. Based on the models, the control scheme for IMC, 2DOF IMC and CIMIC is developed. Controller tuning is carried out to ensure best performance for each controllers.

Distillation Column Simulation:

The industrial scale distillation column is simulated using Aspen Plus software based on the specifications and conditions given in Ilme *et al*. (2001) and Klemola and Ilme (1996) which are shown in Table 1, Table 2, and Table 3 respectively. All the values given are based on distillation column at steady state condition. In this simulation, Peng-Robinson equation-of-state is chosen as thermodynamic property method as the process under consideration deals with high temperature and pressure (Aspen Tech, 2009). For validation purposes, the overall column efficiency is set at 110% which is based on consideration by Ilme *et al*. (2001). The efficiency of a distillation column can reach over 100% due to non-perfect mixing in the liquid phase as the liquid flows across the tray. This usually happens for low volatility distillation column (Kister *et al*., 2007). The full validation results is available in (Muhammad *et al*., 2011). After validated, the model is exported to Aspen Dynamic for dynamic simulation. Recommendations on designing reflux drum and sump are taken from Luyben (2006a) and Kister (1992). The sump and reflux drum dimension are selected based on the heuristic assumption to set up five-minute liquid holdup while the vessel is 50% full when entering and leaving the vessel. The liquid hydraulics and temperature within the stages are calculated by rigorous tray correlations provided in Aspen Plus. The dynamic distillation model is used in linear model identification part.

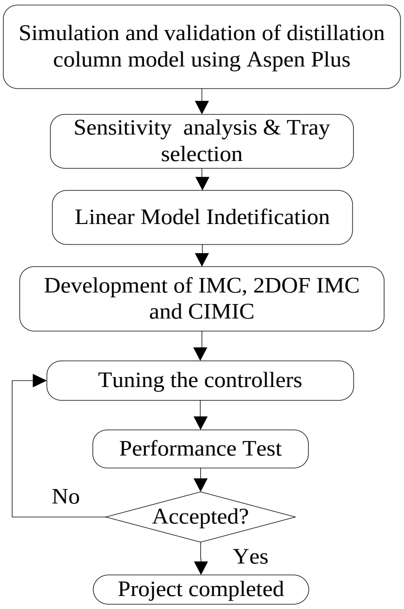

Fig. 1: Overall project methodology

Table 1: Distillation column operation data (Klemola and Ilme, 1996)

Reflux rate (kg/h)	92838	Pressure drop per tray (kPa)	0.47
Reflux Temperature (°C)	18.5	Feed pressure (kPa)	892.67
Column Top Pressure (kPa)	658.6	Boiler duty (MW)	10.24

Table 2: Distillation stream data (Ilme *et al*., 2001)

Components	Feed	Top	Bottom
Propane (wt%)	1.54	4.94	0
i-butane (wt%)	29.5	94.2	0.3
n-butane (wt%)	67.7	0.2	98.1
Isobutene (wt%)	0.13	0.23	0.08
1-butene (wt%)	0.2	0.41	0.1
Neopentane (wt%)	0.11	0	0.17
Isopentane (wt%)	0.77	0	1.12
n-Pentane (wt%)	0.08	0	0.11
Total Flowrate (kg/h)	26122	8123	17999

Table 3: Distillation column specification (Klemola and Ilme, 1996)

Column height (m)	51.8	Downcomer Area (center) (m^2)	0.86
Column diameter (m)	2.9	Tray Spacing (m)	0.6
Number of Tray	74	Hole Diameter (mm)	39
Weir length side (mm)	1.859	Total Hole Area (m^2)	0.922
Weir length center (mm)	2.885	Outlet Weir height (mm)	51
Liquid Flow path Length	0.967m per pass	Tray Thickness (mm)	2
Active Area (m2)	4.9	Number of valves per Tray	772
Downcomer Area (side) (m^2)	0.86	Free Fractional Hole Area (%)	18.82

Temperature Tray Selection:

Inside distillation column, each tray has different temperatures depending on the vapor liquid hydraulic inside the column. There are also other factors that can influence the temperature inside the column such as the temperature of feed, the reaction for reactive distillation and the pressure of the column. In the absence of the concentration analyzer, temperature of the stages can provide a fairly good estimation of the composition. Here, Singular Value Decomposition (SVD) technique as proposed by Luyben (2006b) is used to select the best distillation column stage temperature.

Sensitivity Analysis:

Sensitivity analysis is done to determinate how significance of input towards output of the process. In this study, the selected inputs are reboiler duty and feed composition. The step-test with magnitudes of ±5%, ±10% and ±15% from the nominal values of the input are introduced to the system. The selection of reboiler duty range is made based on predicted operational region of the manipulated variable to produce the desired product composition and the feed composition is based on practical industrial disturbance range as suggested by Bettoni *et al*. (2000).

Linear Model Identification:

Linear model identification is carried out by introducing a step response into the open loop distillation system in the Aspen Dynamic to obtain the input-output profiles. In the test, a gain of +5% from nominal reboiler duty value is introduced to the system. The value chosen is adequate to cover the linear operation region of the distillation column. Matlab System Identification is used to develop the transfer function model of the distillation column

Internal Model Control (IMC):

The development of CIMIC control scheme begins with IMC. The IMC is a control scheme which utilizes a process model to estimate process disturbance and an inverse model as a controller to overcome the disturbance. The linear model developed in previous section is used as process model in IMC control scheme. In order to imitate a practical application, a perturbation of 15% in the process gain and time constant is introduced (Shamsuzzoha and Lee, 2008). The IMC controller development guidelines is available by Bequette (2003). The IMC control scheme is shown in Fig. **2**.

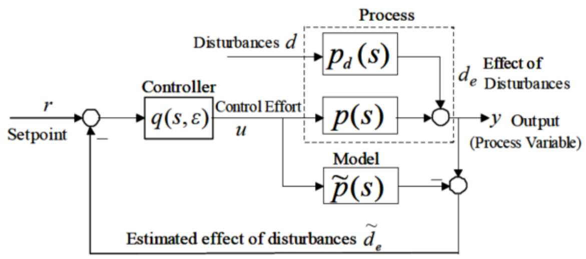

Fig. 2: IMC control scheme (Brosilow and Joseph, 2002)

From Fig. **2**, a closed loop relationship between output y(s), set point r(s) and disturbance d(s) can be described as follows:

$$y(s) = \frac{p(s)q(s)}{1+\left(p(s)-\tilde{p}(s)\right)q(s)} r(s) \tag{1}$$

$$y(s) = \frac{\left(1-\tilde{p}(s)q(s)\right)p_d(s)}{1+\left(p(s)-\tilde{p}(s)\right)q(s)} d(s) \tag{2}$$

2DOF IMC:

A two degree of freedom (2DOF) control system is referred to the number of closed loop transfer functions that can be adjusted independently. Hence, 2DOF IMC is introduced to optimize the set point tracking and disturbance rejection problem separately. Therefore, trade-off between controller performance in set point tracking and disturbance rejection can be handled effectively compared to IMC. The full control scheme of 2DOF IMC is shown in Fig. **3**. Based on the figure, feedback controller, qq_d is designed to reject disturbance while setpoint filter, q_r is designed to shape the response of manipulated variable (MV) to the desired output. The development of feedback controller and set point filter are based on Brosilow and Joseph (2002) work. Technically, both of the blocks are inverse model based controllers which are similar to the IMC controller. However, Brosilow and Joseph (2002) suggested the introduction of the q_d term to be introduced in the disturbance rejection controller. The tuning parameter is used to make the controller more robust towards the model mismatch error.

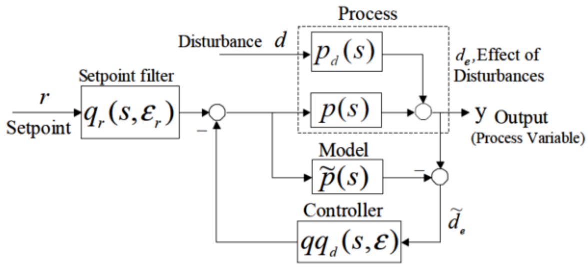

Fig. 3: 2DOF IMC (Brosilow and Joseph, 2002)

CIMIC:

The development of CIMIC is based on work of Häggblom (1996) and Decoupling and Disturbance Rejection (DRD) scheme by Sandelin *et al*. (1991). Based on Fig. **4**, CIMIC structure is fundamentally the modification of the 2DOF IMC with the additional inferential loop. The dashed line represents the process of the distillation column which also incorporates disturbance. In this scheme, the primary CV loop, y is typically associated with a large time delay and hard to measure such as concentration. The secondary loop, v is typically associated with a variable which has a fast sampling time and easy to measure such as temperature. Since the primary CV loop response is delayed, any disturbance that occurs in the system will have a significant impact on the process due to the delayed controller response. Thus, in order to resolve this matter, the disturbance in the primary variable is inferred from the performance of the secondary variable. Besides that, the second variable is normally more sensitive to disturbance and its behavior can be monitored directly.

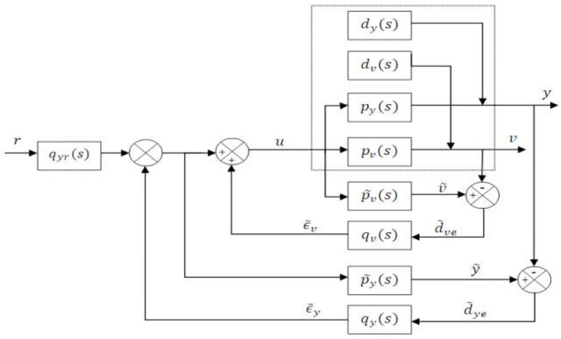

Fig. 4: CIMIC control scheme

The purpose of inverse controller for both the primary and the secondary loop is to invert the model mismatch into appropriate controller effort deviations. In addition, for perfect model case, the model mismatch between the actual process and process model would be explicitly due to the effect of the disturbance. These deviations, $\tilde{\epsilon}_y$ and $\tilde{\epsilon}_v$ would be summed up to counter the disturbance effect in the process. The q_{yr} block is particularly used as the servo control loop. In this work, the bottom product concentration (*n*-butane) is selected as the primary control variable and the secondary variable is the tray temperature.

Tuning:

All tuning parameters used in IMC, 2DOF IMC and CIMIC are obtained using IMCTUNE software (Brosilow and Joseph, 2002). The tuning parameters is calculated based on Mp tuning method (Stryczek *et al*., 2000).

RESULTS AND DISCUSSIONS

Tray Temperature Selection:

Tray temperature selection is done based on the Singular Value Decomposition (SVD) analysis and the result is shown in Fig. **5**. Based on Fig. **5**, the highest Usvd magnitudes for the reboiler duty and the reflux ratio are located at tray 29 and tray 61 respectively. Usvd refers the tray sensitivity measurement matrix where the highest value refers to the most sensitive tray temperature towards the respective MV. These imply that both the trays are the most sensitive to temperature change. However, in this work, temperature tray 68 which has the second highest Usvd value for the reboiler duty is used instead of tray 29 because in general practice, the tray temperature used to infer or control the bottom composition is usually selected near to the bottom stream which is dynamically correlated with each other during constant pressure (Hoffman *et al*., 2006). Thus, it is more appropriate to select tray 68 instead of tray 29.

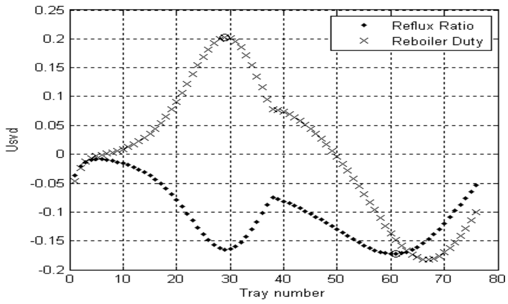

Fig. 5: Usvd vs. tray number plot

Sensitivity Analysis:

In this study, the effect on bottom product composition (*n*-butane) and temperature tray 68 is evaluated based on changes in the reboiler duty and feed composition.

Effect of Reboiler Duty:

The reboiler step test result for *n*-butane purity (XB) and tray 68 temperature (T68) is shown in Fig. **6**. Based on the results, the negative step tests in XB produce larger deviations than the positive step tests. Since the maximum bottom composition of the distillation column model design is only around 98.4%, any positive gain approaching this value will become saturated. On the contrary, a different outcome is produced by T68 where it is much more sensitive to positive steps tests rather than negative step tests. This is due to the effect of increasing vapor boiling up rate in the vapor-liquid transfer in the column which is much more significant in the higher purity region (i.e. the positive step). The distillation column system is known to be more sensitive and nonlinear at a higher purity region (Fuentes and Luyben, 1983). From an overall view, both XB and T68 share a similar trend in the positive and negative step test. This signifies a correlation between XB and T68 output which can justify the application of the inferential technique between the two parameters.

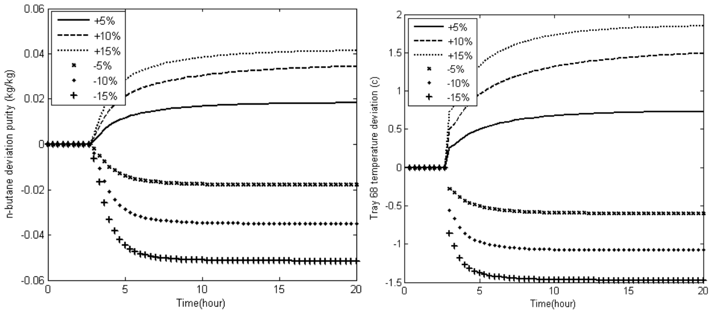

Fig. 6: Step test results for bottom product purity (left) and temperature Tray 68 (right) by manipulating the reboiler duty at ±5%, ±10% and ±15% change from the nominal condition

Effect of Feed Composition:

The effects of disturbance in the feed composition for XB and T68 are shown in Fig. **7**. The figure shows that the increase of the *n*-butane amount in the feed stream would favor the increase in the bottom product composition and vice versa. Since the *n*-butane is a heavy key component, the additional availability of this component in the rectifying section would shift the VLE profile and thus influence the column's temperature as well. In this case, the T68 profile trend is similar with the XB profile trend which indicates the strong relationship between the T68 and XB response with the feed concentration effect.

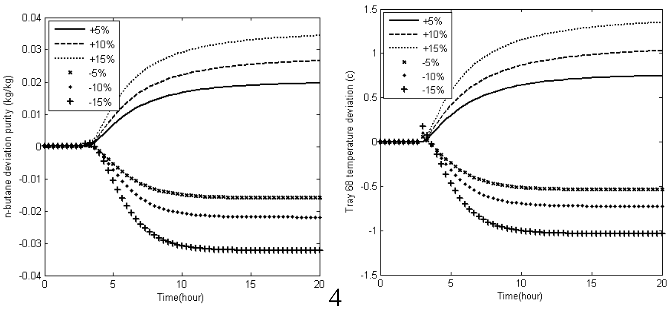

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Fig. 7: Step test results for bottom composition (left) and Tray 68 temperature (right) by manipulating the *n*-butane feed composition at ± 5%, ±10% and ±15% change from the nominal condition

Based on the presented results, XB and T68 parameters have displayed asymmetric response profiles. This observation shows that distillation column under consideration is a nonlinear system (Pearson, 2003). In addition, Muhammad *et al*. (2011) has concluded that the particular distillation column behavior is also affected by multiplicity phenomena which increase the degree of nonlinearity of the process.

Linear Model Identification:

Based on reboiler duty ±5% step test results from sensitivity analysis, a symmetric linear response is observed throughout the process. Thus, it indicates that the process is behaved linearly in the specified region. The results for model identification are tabulated in Table 4. The letter P represents pole, Z is zero and D is delay parameter in a general transfer function model respectively.

Based on Table 4, it can be observed that model P2Z produces the best response for both the distillation column composition model (XB) and the temperature model (T68) based on highest best fits values. The term P2Z shows that the transfer function model has two poles and a single zeros. Based on the results, the transfer function model for the bottom composition XB model is P2Z (referred after this as Model Y) and P2Z for tray temperature T68 model is P2Z (referred after this as Model V). Both transfer function model parameters are shown below:

Table 4: System identification results for Model XB and Model T68

Model XB		Model T68	
Models	Best Fits %	Models	Best Fits %
P2Z	97.88	P2Z	97.7
P1D	93.05	P1Z	97.61
P1	93.05	P1D	87.40
P2	91.47	P2	83.93

Transfer function Model Y:

$$\frac{3.9846e^{-5}(1.919s+1)}{(3.611s+1)(0.67771s+1)} \tag{3}$$

Transfer function Model V:

$$\frac{0.0015863(1.0771s+1)}{(3.1218s+1)(0.01355s+1)} \tag{4}$$

Controller Scheme Design:

The CIMIC controller scheme is developed based on the transfer function model with a modified 2DOF IMC design. Here, the results for all the controller designs are presented below:

IMC Controller

i. The tuning parameter used here is 0.063767

IMC Controller:

$$\frac{(2.447s^2+4.289s+1)}{(7.646e^{-5}s+3.985e^{-5})}\frac{1}{(0.063767s+1)} \tag{5}$$

2DOF IMC

i. The set point controller parameter is the same as the IMC
ii. The disturbance rejection controller Y tuning parameter is 0.1198

Disturbance rejection controller Y:

$$\frac{(2.447s^2+4.289s+1)}{(7.646e^{-5}s+3.985e^{-5})}\frac{1}{(0.11988s+1)}\frac{(0.22539s+1)}{(0.11988s+1)} \tag{6}$$

Linear CIMIC

i. The set point controller and disturbance rejection controller Y parameters for CIMIC are the same as 2DOF IMC

ii. The disturbance rejection controller V tuning parameter is 0.003912

Disturbance rejection controller V:

$$\frac{(0.0423s^2+3.135s+1)}{(0.001709s+0.001586)}\frac{1}{(0.003912s+1)}\frac{(0.00781s+1)}{(0.0039127s+1)} \tag{7}$$

The controller scheme performance is evaluated based on the set point tracking and disturbance rejection test. The Integral Absolute Error (IAE) analysis is used as a tool to quantify each controller performance.

Set point Tracking Test:

The set point tracking test overall result is shown in Fig. **8**. Based on the figure, CIMIC (or LCIMIC) performed better than 2DOF IMC and IMC with faster settling time. Moreover, during the step up test at t=5 hours, CIMIC showed an overdamped response compared with the 2DOF and the IMC slight overshoot behaviour as shown in Fig. **8** (right). This observation is in line with IAE value obtained where the CIMIC gives the smallest value (IAE =0.0689) followed by 2DOF IMC (IAE = 0.1018) and IMC (IAE = 0.1037). The controller output for the set point tracking test can be seen in Fig. **9**. From the figure, it can be observed that all the controllers produced a similar response. However, from a closer observation in Fig. **9** (right), it can be noticed that CIMIC had exerted a faster controller action to track the appropriate MV value compared to 2DOF IMC and IMC. Since reboiler duty energy consumption is large, a small difference in the MV value will result to a significant amount of energy.

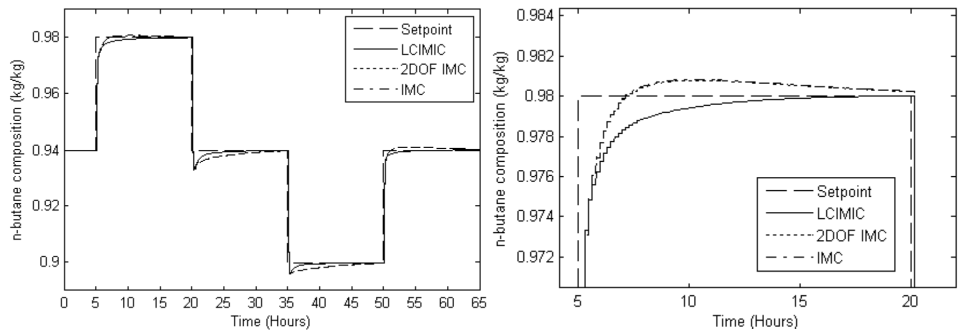

Fig. 8: Set point tracking response results for CIMIC, 2DOF IMC and IMC (left) Set point tracking result for a step up change at t=5 hours (right)

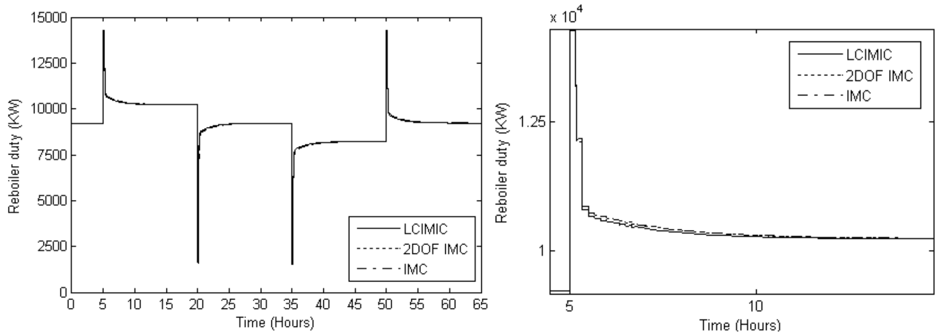

Fig. 9: Set point tracking MV profile results for CIMIC, 2DOF IMC and IMC (left) MV profile for step up at t=5 hours (right)

The overall performance of CIMIC, 2DOF IMC and IMC for set point tracking test is predicted to be almost the same since the controller has the same set point controller parameters. However, the additional inferential loop in CIMIC control scheme has led to a faster response. This is evident because the inferential technique in CIMIC is used to infer a faster secondary variable. Hence, CIMIC controller can act faster based on the inferential response.

Input Disturbance:

Input disturbance usually happens at the beginning of the process such as perturbation in the MV due to certain situations. The value selected for the disturbance is equivalent to 10% of the process input. In this work, the disturbance is introduced at t = 1 hours for 5 hours as shown in Fig. **10**. Based on the figure, CIMIC scheme displays its profound ability to effectively reject the disturbance compared to the other control schemes. The reason for the poor performance for IMC and 2DOF IMC is because the effect of input disturbance is not considered by the controllers. Since IMC and 2DOF IMC only perform based on model mismatch, the effect of input disturbance is not explicitly included in the mismatch. Thus, the controller is unable to produce the correct gain to compensate the input disturbance (Wassick and Tummala, 1989). Since CIMIC has another loop to infer the disturbance from the tray temperature profile, the existence of the disturbance at the beginning of the process can be rejected. To reject the input disturbance promptly, CIMIC controller must exert a sudden extent of effort if compared to the other controllers but it is still within the operation constraint as shown in Fig. **10** (right).

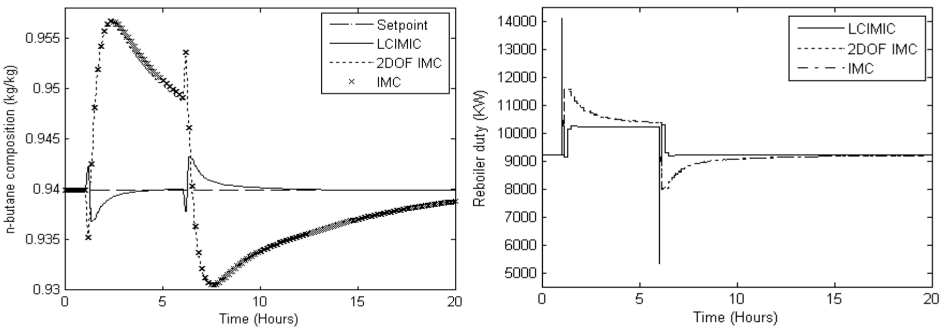

Fig. 10: Input disturbance rejection results for CIMIC, 2DOF IMC and IMC (left) Input disturbance rejection MV profile for CIMIC, 2DOF IMC and IMC (right)

Output Disturbance:

Generally, the occurrence of the disturbance at the process output is related to the perturbation that occurs inside the process such as tray condition, sudden drop in column pressure and equipment failure which can affect the process response directly or indirectly. Thus, a step disturbance of 10% from the nominal value of the process output stream is introduced at t=1 hours for 14 hours as shown in Fig. **11**. Based on the figure, all the control schemes can equally reject the output disturbance using almost the same magnitude of the reboiler duty as shown in Fig. **11** (right). However, the controller needs at least 14 hours to fully reject the disturbance. Since *n*-butane/*i*-butane distillation process has a slow process gain, thus, such characteristic is inherited by the linear model and controller as well.

The capability of the linear controllers to reject output disturbance is anticipated since such disturbances can lead to an additional model mismatch in the control scheme. If a perfect process model is available, then such mismatch is solely due to the disturbance in the process. Based on the quantitative error test, the CIMIC produced a better performance with the IAE of 0.0346 compared with the 2DOF IMC (IAE = 0.0364) and IMC (IAE = 0.0370). Thus, the addition of the inferential loop in CIMIC has again helped the controller to properly reject the disturbance. The 2DOF IMC is found to perform slightly better than the IMC since the disturbance is compensated using a different controller. Unlike the IMC scheme, both the set point tracking and the disturbance rejection cases are handled by a single controller only.

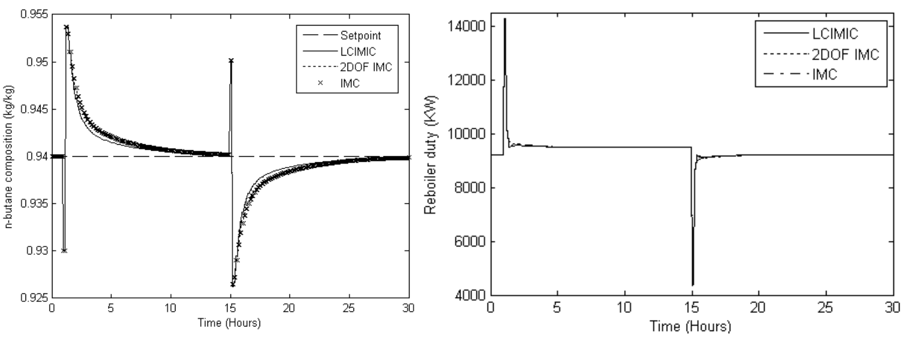

Fig. 11: Output Disturbance rejection results for CIMIC, 2DOF IMC and IMC (left) MV profile for CIMIC, 2DOF IMC and IMC output disturbance rejection test (right)

Results Summary:

A summary of the results from the set point tracking, input disturbance and output disturbance test is presented in Table 5. It can be observed that CIMIC produced a better performance than 2DOF IMC and IMC in all the tests conducted. In the set point tracking test, CIMIC has improved the response profile response by 32.3% and 33.5% compared to 2DOF IMC and IMC, respectively. In rejecting the outside disturbance, CIMIC has able to exceed others by 14.0% and 16.8%, respectively. The most noteworthy improvement is in rejecting the internal disturbance which is 93.5% and 93.7%, respectively. This shows that the application of inferential technique inside CIMIC has given advantage in term of predictive ability in rejecting the disturbance (Dumpa *et al*., 2016). Similar application of modified IMC to improve disturbance rejection capability is also reported in the literature (Jin and Liu, 2015). Furthermore, the concept of 2DOF IMC technique has provided a flexibility in adjusting appropriate tuning parameters for the set point tracking and disturbance rejection cases (Saxena and Hote, 2012).

Table 5: Summary results for CIMIC, 2DOF IMC and IMC performance

Control Scheme	IAE		
	Step Test	Input Disturbance	Output Disturbance
CIMIC	0.0689	0.0080	0.0497
2 DOF IMC	0.1018	0.1236	0.0578
IMC	0.1037	0.1274	0.0597

Conclusions:

A model based control (MBC) strategy, namely CIMIC, has been developed to control *n*-butane/*i*-butane distillation process. CIMIC is a modified 2DOF IMC with inferential technique. CIMIC has the advantage of handling slow sampling measurement control loop and has an effective disturbance rejection capability. In this study, steady state and dynamic model of *n*-butane/*i*-butane distillation column was successfully developed using Aspen software and validated with industrial data. The *n*-butane/*i*-butane distillation process offers a unique challenge as it has a multicomponent feed stream, low relative volatility and slow dynamics. A linear transfer function model of the distillation column was developed using Matlab System Identification and was then used in the development of IMC, 2DOF and CIMIC control scheme. Based on the performance test, CIMIC has shown its ability to control the distillation model properly especially in rejecting internal disturbance. Thus, it gives an encouraging opportunity to further develop CIMIC using nonlinear model to handle nonlinear process.

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