IMPROVED MULTI MODEL PREDICTIVE CONTROL FOR DISTILLATION COLUMN

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FOR DISTILLATION COLUMN

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To my beloved ….

Abah *Surhim* (alm.), Emak *Sawi*,

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My wife *Purnawirawati Dwisiwi*

and

My children:

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ABSTRACT

Model predictive control (MPC) strategy is known to provide effective control of chemical processes including distillation. As illustration, when the control scheme was applied to three linear distillation columns, i.e., Wood-Berry (2x2), Ogunnaike-Lemaire-Morari-Ray (3x3) and Alatiqi (4x4), the results obtained proved the superiority of linear MPC over the conventional PI controller. This is however, not the case when nonlinear process dynamics are involved, and better controllers are needed. As an attempt to address this issue, a new multi model predictive control (MMPC) framework known as Representative Model Predictive Control (RMPC) is proposed. The control scheme selects the most suitable local linear model to be implemented in control computations. Simulation studies were conducted on a nonlinear distillation column commonly known as Column A using MATLAB® and SIMULINK® software. The controllers were compared in terms of their ability in tracking set points and rejecting disturbances. Using three local models, RMPC was proven to be more efficient in servo control. It was however, not able to cope with disturbance rejection requirement. This limitation was overcome by introducing two controller output configurations: Maximizing MMPC and PI controller output (called hybrid controller, HC), and a MMPC and PI controller output switching (called MMPCPIS). When compared to the PI controller, HC provided better control performances for disturbance changes of 1% and 20% with an average improvement of 12% and 20% of the integral square error (ISE), respectively. It was however, not able to handle large disturbance of + 50% in feed composition. This limitation was overcome by MMPCPIS, which provided improvements by 17% and 20% of the ISE for all of types and magnitudes of disturbance change. The application of MMPCPIS on a single model MPC strategy produced almost similar performance for both types of disturbances, while its application on MMPC yielded better results. Based on the results obtained, it can be concluded that the proposed HC and MMPCPIS deserve further detailed investigations to serve as linear control approaches for solving complex nonlinear control problems commonly found in chemical industry.

ABSTRAK

Strategi kawalan ramalan model (MPC) telah diketahui efektif dalam mengawal proses-proses kimia termasuk penyulingan. Sebagai ilustrasi, apabila skema kawalan ini digunakan ke atas tiga turus penyulingan lelurus, i.e., *Wood-Berry* (2x2), *Ogunnaike-Lemaire-Morari-Ray* (3x3) dan *Alatiqi* (4x4), hasil yang diperoleh membuktikan keunggulan MPC lelurus berbanding pengawal PI. Namun ianya tidak sama bagi proses yang mempunyai dinamik tidak lelurus, dan pengawal lebih baik diperlukan. Sebagai usaha untuk mengatasi kelemahan ini, satu kerangka kawalan ramalan berbilang model (MMPC) baru yang dinamakan Kawalan Ramalan Model Perwakilan (RMPC) dicadangkan. Skema kawalan ini memilih model lelurus tempatan yang paling sesuai untuk digunakan dalam pengiraan-pengiraan kawalan. Kajian-kajian simulasi ke atas turus penyulingan tidak lelurus yang dikenali sebagai Turus A telah dilaksanakan dengan menggunakan perisian MATLAB® dan SIMULINK®. Pengawal-pengawal tersebut dibandingkan dari segi kebolehan mengikuti titik-set dan nyah-gangguan. Dengan menggunakan tiga model tempatan, RMPC terbukti lebih cekap bagi kawalan servo. Walau bagaimanapun, ia tidak dapat menangani keperluan nyah-gangguan. Kelemahan ini telah diatasi dengan memperkenalkan dua konfigurasi keluaran pengawal: memaksimumkan keluaran pengawal MMPC dan PI (dipanggil pengawal hibrid, HC), dan penukaran keluaran pengawal MMPC dan PI (dipanggil MMPCPIS). Jika dibandingkan dengan pengawal PI, HC telah berjaya menghasilkan prestasi kawalan yang lebih baik dalam menangani gangguan bersaiz 1% dan 20% dengan pembaikan purata 12% dan 20% dalam ralat kamiran kuasa dua (ISE). Namun ianya tidak dapat mengendalikan gangguan sebesar +50% pada komposisi masukan. Kelemahan ini telah diatasi oleh MMPCPIS yang telah menghasilkan pembaikan ISE 17% dan 20% bagi semua jenis dan saiz gangguan. Penggunaan MMPCPIS pada model mpc tunggal menghasilkan prestasi hampir sama bagi kedua–dua jenis gangguan, manakala penggunaannya pada MMPC menghasilkan keputusan yang lebih baik. Berdasarkan keputusan yang diperoleh dalam kajian ini, dapat disimpulkan bahawa HC dan MMPCPIS yang dicadangkan ini sesuai untuk dikaji dengan lebih lanjut sebagai kaedah kawalan lelurus untuk menyelesaikan masalah kawalan tidak lelurus kompleks yang selalu ditemui dalam industri kimia.

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A	-	State (or system) matrix
B	-	Input matrix
C	-	Output matrix
d	-	Disturbance
D	-	Zero matrix
d_{meas}	-	Measurable disturbance
$\bar{\mathbf{e}}$	-	Vector of predicted errors
$e(k + j)$	-	Predicted error based on past moves
E	-	Error
$e_{ff}(k + j)$	-	Predicted error based on past moves for feedforward
F	-	Feed flow rate
G	-	Transfer function
$\mathbf{I}$	-	Identity Matrix
j	-	Sampling instants
J	-	Cost function
k	-	Current time
K	-	Steady state gain
K_c	-	Proportional gain
k_d	-	Discrete dead time
K_p	-	Process gain
K_{cu}	-	Ultimate gain
l	-	Adaptation law
M	-	Control horizon
N	-	Model horizon
N_{data}	-	Number of data
NM	-	Number of model

P	-	Prediction horizon
P_u	-	Ultimate period
Q	-	Weighting diagonal matrices
q	-	Discrete variable
R	-	Weighting diagonal matrices
s	-	Laplace transform variable
T	-	Sampling time
t_s	-	Sampling interval
t_{sw}	-	Switching time or clock period
t	-	Operating time
u	-	System input
w	-	Weight
x	-	State variable
$\bar{y}$	-	Mean of the data
$\hat{y}(k+1)$	-	Future behaviour of the process variable
y	-	System output
y_0	-	Initial condition of the process variable
y_f	-	Effect of past moves
y_m	-	Measurement of y
y_{meas}	-	Actual value of the measured process variable
y_{sp}	-	Set point of y
z	-	Discrete variable
z_F	-	Feed composition

GREEK SYMBOLS

$\Delta\bar{\mathbf{u}}$	-	Vector of controller output moves to be determined.
Δu	-	Control move
γ	-	Weighting factor of output
ε	-	Deviation of y_m from y_{sp}
$\varepsilon(t)$	-	PID controller input

θ_p	-	Dead time
λ	-	Move suppression coefficient
σ	-	Variance of data
τ_{CL}	-	Closed-loop speed of response
τ_D	-	Derivative time constant
τ_d	-	Time constant of disturbance
τ_R	-	Integral (reset) time constant
τ_p	-	Time constant
$\tau_{p,global}$	-	Time constant of global FOPDT model
ω_u	-	Ultimate frequency

LIST OF ABBREVIATIONS

ADMC	-	Adaptive Dynamic Matrix Control
AFMBPC	-	Adaptive Fuzzy Model-Based Predictive Control
ALDMC	-	Adaptive Linear Dynamic Matrix Control
AMPC	-	Adaptive Model Predictive Control
ANMPC	-	Adaptive Nonlinear Model Predictive Control
APC	-	Adaptive Predictive Control
ARMA	-	Autoregressive Moving Average
ARMAX	-	Autoregressive Moving Average Exogenous
ARX	-	Autoregressive Exogenous
BJ	-	Box-Jenkins
CLACE	-	Closed-Loop Control Error adjustment
D	-	Derivative
DMC	-	Dynamic Matrix Control
DTC	-	Dwell-Time Concept
EHAC	-	Extended-Horizon Adaptive Control
EKF	-	Extended Kalman Filter
EOP	-	Effective Open-loop Process
EPSAC	-	Extended Prediction Self-Adaptive Control
FOPDT	-	First Order Plus Dead Time
FP	-	First Principle
GA	-	Genetic Algorithm
GBN	-	Generalized Binary Noise
GPC	-	Generalized Predictive Control
HC	-	Hybrid Controller
HIECON	-	Hierarchical Constraint Control
HW	-	Hammerstein-Wiener

I	-	Integral
IDCOM	-	Identification and Command
IMC	-	Internal Model Control
ISAT	-	In Situ Adaptive Tabulation
ISE	-	Integral Square Error
LMPC	-	Linear Model Predictive Control
MAC	-	Model Algorithmic Control
MIMO	-	Multiple-Input Multiple-Output
MISO	-	Multiple-Input Single-Output
MMAC	-	Multiple Model Adaptive Control
MMPC	-	Multiple Model Predictive Control
MMPCPIS	-	Multiple Model Predictive Control and PI Switching
MMST	-	Multiple Models, Switching and Tuning
MPC	-	Model Predictive Control
MPCI	-	Model Predictive Control and Identification
MPHC	-	Model Predictive Heuristic Control
MRAC	-	Model Reference Adaptive Control
MSE	-	Mean Square Error
MV	-	Manipulated Variable
MVC	-	Multi Variable Control
MWAC	-	Model Weighting Adaptive Control
NARX	-	Nonlinear Autoregressive Exogenous
NLC	-	Non-Linear Control
NMPC	-	Nonlinear Model Predictive Control
NNN	-	Nonlinear Neural Network
NSS	-	Nonlinear State-Space
OE	-	Output-Error
P	-	Proportional
PFC	-	Predictive Functional Control
PEM	-	Parameter Estimation Method
PI	-	Proportional-Integral
PID	-	Proportional-Integral-Derivative
PRBS	-	Pseudo Random Binary Signal
QP	-	Quadratic Programming

QSM	-	Quasi-Sliding Mode
RCGA	-	Real-Code Genetic Algorithm
RMPC	-	Representative Model Predictive Control
RMSE	-	Root Mean Square Error
RTO	-	Real Time Optimization
SEL	-	Catalyst Selectivity
SISO	-	Single-Input Single-Output
SAC	-	Simple Adaptive Control
SANMPC	-	Self-Adaptive Nonlinear Model Predictive Control
SMAC	-	Setpoint Multivariable Control Architecture
SMOC	-	Shell Multivariable Optimizing Controller
SMPC	-	Single Model Predictive Control
SMPCPIS	-	Single Model Predictive Control and PI Switching
SNP	-	Static Nonlinear Polynomial
SOPDT	-	Second Order Plus Dead Time
SSE	-	Sum Square Error
STC	-	Self-Tuning Controller
STR	-	Self-Tuning Regulator
TSK	-	Takagi-Sugeno-Kang

LIST OF APPENDIX

CHAPTER 1

INTRODUCTION

1.1 Motivation of Study

Chemical plants involve many different process units with variety of characteristics. The processes are typically nonlinear, multivariable, and involving high degree of process interactions, giving rise to variety of operational complexities (Bachnas *et al*., 2014). These issues impact the performance of the plant operation system, the heart of which is process control. Since process controllers are generally developed based on linear theories, nonlinearities in process behaviors often cause substandard control performances in some control loops. When coupled with strong interactions between control loops that are oftentimes unavoidable, the control problems become more intricate. This is also exacerbated by the fact that process controllers typically used in the industry are based on linear single-input single-output (SISO) design (Halvarsson, 2010). As such, the desired plant performance cannot be established without human interventions, a requirement which in practice is alerted by the alarm systems. Although this approach is in principle workable and has been in practice for many years since the old days, the needs for better plant performances demand better control strategies.

In some specific plants such as petroleum refinery and various petrochemical processes, model predictive controller (MPC) has been introduced and is now receiving wide acceptance (Potts *et al*., 2014). In these cases, linear models are used generally used for ease of implementation and the fact that linear models are easier to be developed and interpreted. Since process characteristics are in general nonlinear, which can be highly nonlinear depending on the operating zones, the use of such controllers are limited to quite a narrow operating window. This limits the potential of the plant to be operated in more optimal manner. When the operation drifted away from the operating window within which the linear models were developed, the control performances begin to degrade, which may even lead to failure, and again requiring human interventions, which in principle defeated the philosophy of MPC in the first place.

MPC aims at providing accurate automatic control to enable efforts such as real time optimization (RTO), or production of high purity products (Martínez *et al*., 2014). These initiatives often require the plant to operate near process constraints, which in turn demands more accurate nonlinear models to be used within control framework such as MPC to provide accurate prediction of process behavior and explicit consideration of state and input constraints (Liu *et al*., 2015). When this is realized, better control is established, especially when the model is comprehensive and accurate.

Despite these clear advantages, the application of NMPC in the process industry is still limited. One of the key reasons is the fact that the model is more difficult to be fitted and is difficult to be understood by plant operators compared to the linear counterparts. It also requires intensive on-line computations to produce control moves by solving large-scale nonlinear models at each sampling period, and consequently, it is less popular in the industry (Cao, 2005; Magni *et al*., 2009; Ellis and Christofides, 2014).

A promising solution to overcome this issue is to employ a multi-model MPC or Multiple MPC (MMPC). In this case, the models are basically consisting of an array of linear models in MIMO configuration at each certain range of controlled variables (CVs) or output variables. Some of the advantages of this strategy include its simplicity in modeling, better predictability, and ease of maintenance. However, since it is essentially a linear MPC (LMPC), it is still subjected to all limitations of the typical LMPC. The hope is that, if sufficient number of models is used, and each model in an accurate representation of the process within their region operation, accurate control can be established. This is of course subject to a condition of availability of an effective switching algorithm such that the changes in process models to be used for control computations are managed effectively and efficiently.

1.2 Problem Statements

The potential of MPC in managing control problems in chemical industry has been reported in academia as well as industry. However, despite the rapid development in nonlinear control, its application has been hampered by the needs for highly skilled process control engineers to deal with nonlinear modelling problems (Praprost *et al.*, 2004). As such, linear models are preferred. A good compromise strategy is to go along the idea of multi-model MPC where multiple linear models are used. To enable effective and efficient implementation of this strategy, the following issues should be answered:

- How to implement LMPC in several variations of linear multivariable models? Are its controlling performances better than those of conventional controller?
- How to control a nonlinear dynamic model using MMPC?
- If the MMPC is confronted with problems in controlling nonlinear dynamic model, what is the suitable approach to improve it?

1.3 Objective and Scope of Work

The aim of this research is to investigate the use of multi model predictive control approach in controlling distillation process. Distillation process is chosen as the case study because of its widespread use in the chemical process industry and the fact that it is nonlinear, interactive and often require high purity product requirement. As such, it is a good testbed to study the issue mentioned in the above. In particular, the work is limited to following key objectives:

(i) Evaluation of linear MPC on linear multivariable distillation control problems.

(ii) Control of nonlinear distillation process using MMPC approach.
(iii) Investigation and analysis of the effects of MMPC parameters tuning.
(iv) Investigation and analysis of effects of controller output configurations.

To facilitate the study, the study employs four distillation models, i.e. Wood-Berry (linear 2x2 system), Ogunnaike-Lemaire-Morari-Ray (linear 3x3 system), Al-Atiqi (linear 4x4 system) and Skogestad's nonlinear distillation model (Column A).

1.4 Contribution of Work

This work illustrates the developmental and implementation issues of MMPC in addressing nonlinear distillation control problems. The findings have illustrated that an MMPC strategy employing linear models, referred to as linear MPC (LMPC), has provided better control performance compared to conventional control when tested on multivariable linear distillation control problems (Ahmad and Wahid, 2007a,b; Wahid and Ahmad, 2008). The strategy adopted was to select the best LPMC as a candidate to build MMPC, and is called representative model predictive control (RMPC). In this control scheme, the substitution between one model to another is specified by the CV value changes from one condition to another.

Although, RMPC was able to outperform the conventional control for setpoint tracking (Wahid and Ahmad, 2009; Wahid *et al*., 2013), it still has a limitation on large disturbance rejections. This is improved by introducing a strategy called hybrid controller (Wahid and Ahmad, 2015). However, the hybrid controller fails when the system was subjected to large disturbance change in the form of changes in the feed composition (50%). Although in practice, disturbances are expected to be smaller, large disturbances were used in this study to evaluate the ability of the controller at extreme conditions. This limitation was successfully overcome by controller output switching strategy involving both MMPC and PI controller. This control scheme was able to deal with both setpoint tracking and disturbance rejection problems.

1.5 Thesis Organization

Following this introductory chapter, the literature review is presented in chapter 2, which lays the theoretical basis of research as well as the review of closely related literatures. Important issues in MPC and NMPC, evaluation of their advantages and disadvantages along with potentials of MMPC for solving nonlinear processes problem are discussed. The research gap is subsequently identified and addressed as well in this chapter. This is followed by simulation studies on application of linear MPC to three well-known distillation test-beds, i.e. Wood-Berry (2x2 system), Ogunnaike-Lemaire-Morari-Ray (3x3 system), and Al-Atiqi (4x4 system).

The advantages of linear MPC are assessed in comparison with conventional controller for the setpoint tracking and disturbance rejection. Next, the thesis explores the control of nonlinear distillation process based on Column A in chapter 4. This chapter focuses on the usage of linear MPC in the MMPC configuration to solve nonlinear problems.

As an extension to available MMPC approach, in chapter 5 a control strategy called controller output configurations. This is followed by the conclusion and recommendation for future works, presented in chapter 6.

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