EFFECT OF HEAT INTEGRATION ON DYNAMIC OF A PALM-OIL FRACTIONATION PROCESS

Arshad Ahmad, Ling Leong Yau
Department of Chemical Engineering,
Universiti Teknologi Malaysia
Skudai, Johor.

ABSTRACT

The needs to satisfy more stringent safety and environmental regulations whilst maintaining the economic objective has drawn more attention towards multiple integration of process units leading to more challenging plantwide process control problem. In order to solve this growing problem, it is desirable to have rigorous models for realistic and large-scale processes. In this paper, the development of dynamic models of a typical palm oil fractionation process using HYSYS.Plant simulation package is presented. Results illustrating the effect of recycle stream and heat integration to process control performance are also displayed and discussed.

Keywords: Plantwide process control, Palm-oil fractionation process, Process Dynamic, Distillation

INTRODUCTION

Plantwide control has captured numerous attentions from both academia and industries over the past few years. This is largely due to the way modern process plants are being designed with multiple heat and material integration and with a small over-design margin which in turn leads to more challenging control problems. This has not been the case in the past because in those days, many plants had utilised surge tanks for buffering the disturbances, and little heat and material integration. Control system designs were often based on the unit operation approaches as suggested by Umeda (1978) for this system had work reasonably well. In general, this approach can be divided into three steps. The first step involves the decomposition of the plant into individual units. Next, one generates the best control structure for each unit. Then, the pieces are combined together into a control system for entire plant. Finally, the conflicts among the individual control structures are eliminated.

The needs to satisfy all the contradicting requirements ranging from capital and operating costs to environmental and safety regulations, and product specifications has led to complex plant design with material and energy recycle and integration. At times, this integration is done without having a complete understanding of its effect on the plant controllability and operability. This leads to a more complex dynamic behaviour resulting in undesired higher product variability. Therefore, the design of an effective control system for regulating the entire plant operation with a good dynamic performance plays a critical role in industrial process control today.

The pioneer work on plantwide control was reported by Buckley (1964), in which he proposed decomposing the problem based on time-scale differences. He also noted that using product quality controllers are much faster compared to material balance controllers. Morari *et al.* (1980) presented a unified formulation for the problem of synthesizing control structures for chemical processes. Since the introduction of the Tennessee Eastman test problem by Downs and Vogel (1993), many researchers have explored the idea of plantwide control strategy using this useful test-bed. For example, Price and Georgakis (1993) proposed a tiered-design framework and derived several guidelines for the plantwide control system design problem of this plant. Then, many researchers used the same test-bed by applying advanced process control such as model predictive control to improve the dynamic behaviour of the process itself.

Another active test-bed, is the HDA process that has been studied extensively by Luyben and co-workers (1998). They proposed a nine-step procedure for plantwide control system design through the use of heuristic and experience. The procedure synthesizes only one control structure and does not consider any alternative structure. Then there is the self-optimising control that was introduced by Skogestad (2000). This is a method for selecting the best pairing of controlled variables and manipulated variables to create a stable process condition that indirectly maximize profits. Groenendijk (2000) proposed a systems approach for evaluating dynamic and plantwide control of a VCM plant that incorporated heat and material integration. This is a simulation-based methodology for assessing the mass integration to find the best flowsheet pairing from controllability point of view. This approach is useful for revamping existing control structures.

As far as the open literature on plantwide control and operation, reports involving palm-oil fractionation process is still non-existent. In fact, oleochemical sector is still an area that is not well investigated by researchers in the process systems engineering. Due to high degree of complexity of the palm oil and its processes involved, the development of a rigorous process model would require the use of commercial simulators. Even then, the development of a reasonably accurate flowsheet that represent the dynamic behaviour of the process would be rather demanding. This has been the underlying motivation for this work.

This paper describes the development of dynamic models for palm oil fractionation process using HYSYS.Plant. The process models that have been developed contain real non-ideal chemical components, realistically large process flowsheet containing standard unit operations and recycle streams and energy integration. Finally, a study on the effect of heat integration based on the dynamic model is presented as an example.

PROCESS DESCRIPTION

The case study is a typical palm oil fractionation plant in which fatty acids are split. The process consists of five distillation columns connected in series. A simplified schematic diagram of the plant is provided in Figure 1 below. The feed to the plant is Palm Kernel Oil (PKO), which contains fatty acids from C6-C18. In the pre-cut column, light products (C8 and C10) are recovered in the overhead. The bottom stream is fed to the light-cut column where C12 is separated from the rest of its constituents. The bottom product then enters the middle-cut column where C14 and C16 are recovered leaving the C18 to be purified from the rest of the oil constituents in the still-cut column. In the residue cut column, C18 is recovered and recycled back to the previous column. All the distillation columns operate at highly vacuum condition generated using steam ejectors. The columns are packed with structured packing to provide the desired separation properties.

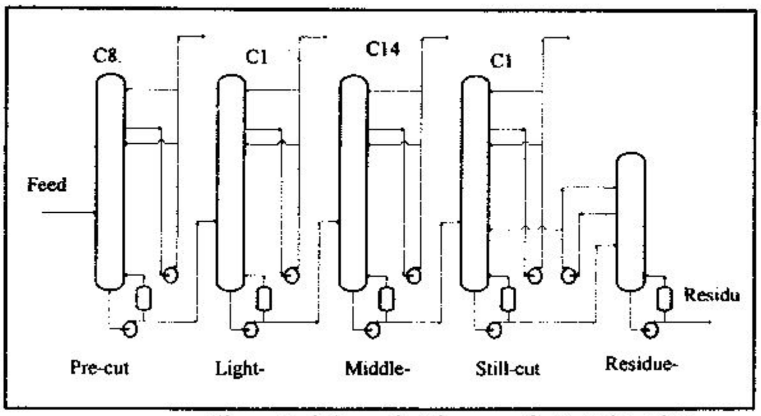

Figure 1: Schematic Diagram of a Fractionation

Due to high operating temperatures, thermal oil is used for heating in the reboilers. A sectional view of a typical column is shown in Figure 2. The column is equipped with a pump-around system. Liquid collected on the trap-out tray is drawn out from a side-draw stream. The stream is split into two - a reflux stream and a stream that goes through an external cooler. The cooled stream is again split into two streams, one returning to the column as a recycle and one as the distillate. This pump-around system provides a means for the vapour in the column to be condensed through direct contact with the cooled liquid from above

MODELLING AND SIMULATION

Simulation of the column was carried out using HYSYS.Plant, a simulator that contains an extensive thermodynamic property package, a library of standard unit operations, logic operations such as PID controllers and a user-friendly flowsheet environment to allow quick modelling of the process. However, due to the non-conventional nature of the palm oil distillation system, the development of the flowsheet for the simulation has not been straightforward. For example, one of the long-chained fatty acid, caprylic acid (C8) was not found in HYSYS Component Properties Library. As such, C8's properties were estimated using HYSYS Hypothetical Component Manager. One has to provide as much data as possible so the estimates will be realistic. Since HYSYS.Plant does not support packed column and the equivalent tray (HETP) calculation was used. Similarly, model of a direct contact internal condenser based on packed column is also not available within the standard library and modifications were implemented.

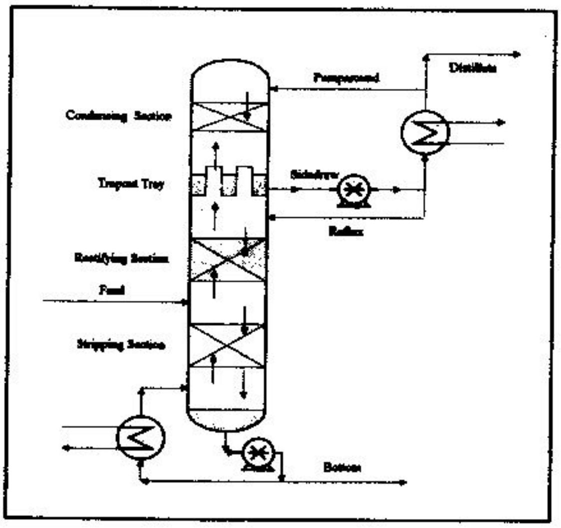

Figure 2: A typical packed column

In the simulation, UNIQUAC activity model was selected for physical property calculations due to the facts that fatty acids (carboxylic acids) are polar component with non-ideal behaviour. The vapour phase fugacity was calculated using Virial model since the system displays strong vapour phase interactions. There are six internally material recycle streams, contributed by each pump-around stream of every column and the sixth by a recycle stream from the residue column to the still-cut column. Therefore, six RECYCLE operations were inserted in the flowsheet.

In running the simulation, the conditions of the streams are assumed first. Then the iterative calculations were carried out automatically until the values in the calculated streams matched those in the assumed stream within specified tolerances. In doing this, proper initial values should be chosen for each stream, otherwise the system might converge to some undesirable values due to the non-linearity and unstable characteristics of the process. (HYSYS Manual, 2000)

Hysys flowsheeting has to major environments: the Main and the Column. The Column is a special type of sub-flowsheet in HYSYS. A Sub-Flowsheet contains equipment and streams, and exchanges information with the Parent Flowsheet through the connected streams. From the Main Environment, the Column appears as a single, multi-feed multi-product operation (see Figure 5). When columns are modelled for steady-state condition, pressure profiles, numbers of trays and the feed tray number must be specified in addition to the specification of the inlet streams conditions. Additionally, two specifications must also be provided for columns having both the reboiler and the condenser installed using variables such as their duties, the column reflux rate, draw stream rates and the composition fractions. In this study, the reflux ratio and the overhead composition mole fractions have been used. The tray sections of the columns are calculated with the HYSYS Tray Sizing Utility. Though the tray diameter, weir length, weir height, and the tray spacing are not required in steady state modelling, they are required for an accurate and stable dynamic simulation.

Model of a column with a direct contact internal condenser is also not available within the HYSYS library. As a result, modifications of the HYSYS standard column template were carried out to produce the equivalent configuration as shown in Figure 3. A close match between the simulation and the industrial data was obtained despite the unavailability of an exact column template. The effect of disturbance is shown in Figure 4 when the feed C8 mass flow is increase 10%.

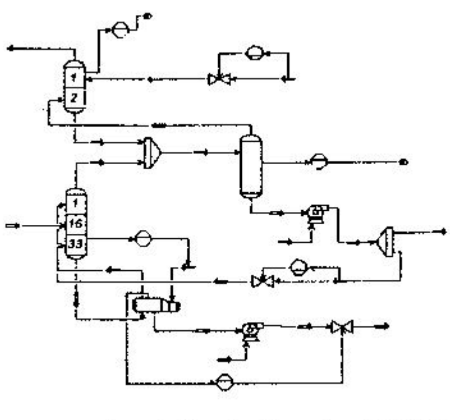

Figure 3: Pump-around modelling for the palm-oil distillation process

EFFECT OF HEAT INTEGRATION

Heat integration is a possible cause of ill performance of the control loops operation. Although the heat integration is a way to reduce operating cost, a complete understanding its effect on the process controllability is a must. This section studied the effect of heat integration for the first two columns. In this study, distillate stream from Light-Cut column was used to heat up the feed to the process. A make-up cooler was used to increase the temperature up to 210 °C. This configuration was named Case 2 and shown in Figure 6. Figure 5 shows the alternative case referred to as Case 1 that is a set-up of the column without any integration and is used to compare the integration effect on process operability.

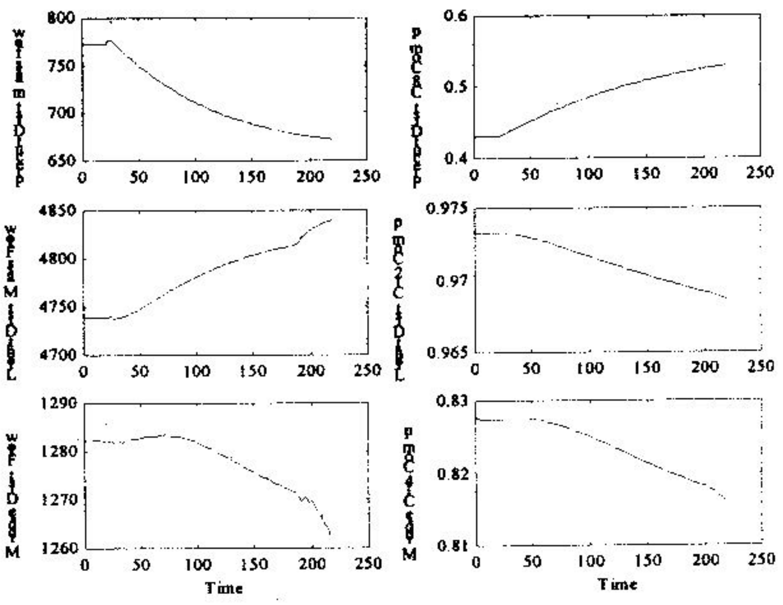

Figure 4: Dynamic behaviour of the process for step change in C8 for 10% in mass flow at the feed.

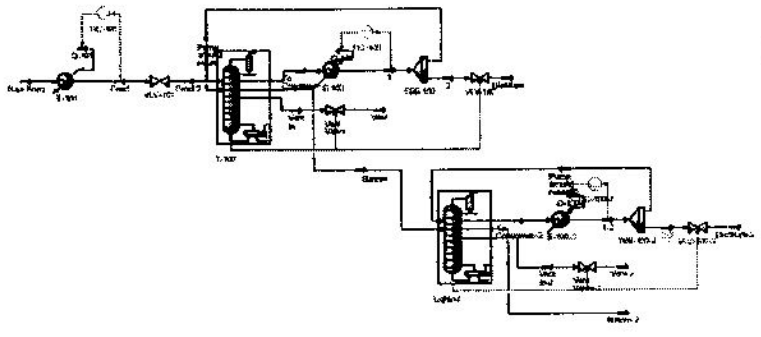

Figure 5: Case 1: Column Configuration without heat integration.

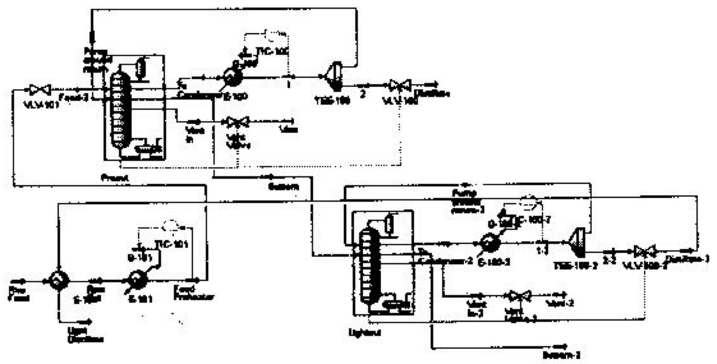

Figure 6: Case 2: Column Configuration with heat integration.

A temperature disturbance on Raw-feed stream was introduced from 30 °C to 40 °C. The results for case 1 and case 2 are shown in Figure 7 and Figure 8 respectively. The graphical illustrations of the results clearly expressed the decrease in the heat duty as a result of heat integration. However, heat integrated scheme demands longer time for the response to settle to its ultimate value.

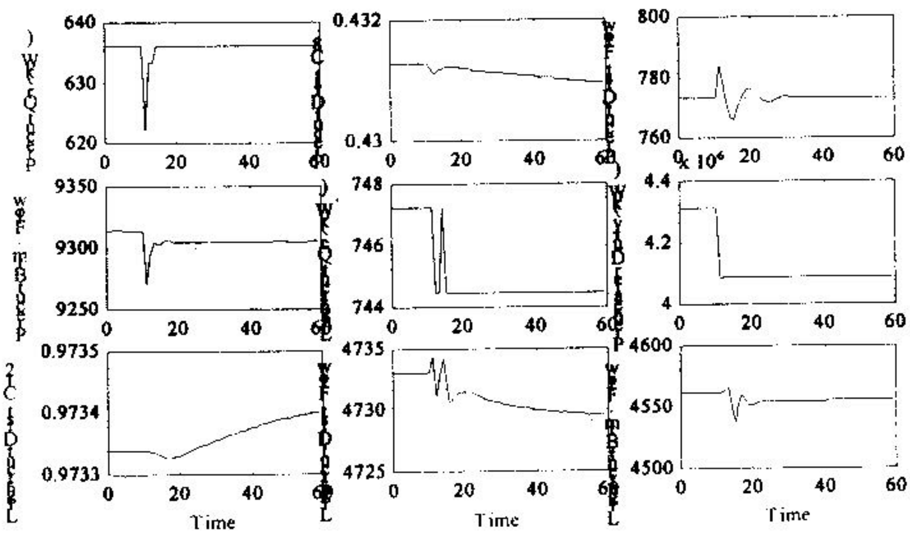

Figure 7: Responses of the process variables to step change of Raw-Feed temperature from 30 °C to 40 °C for case 1.

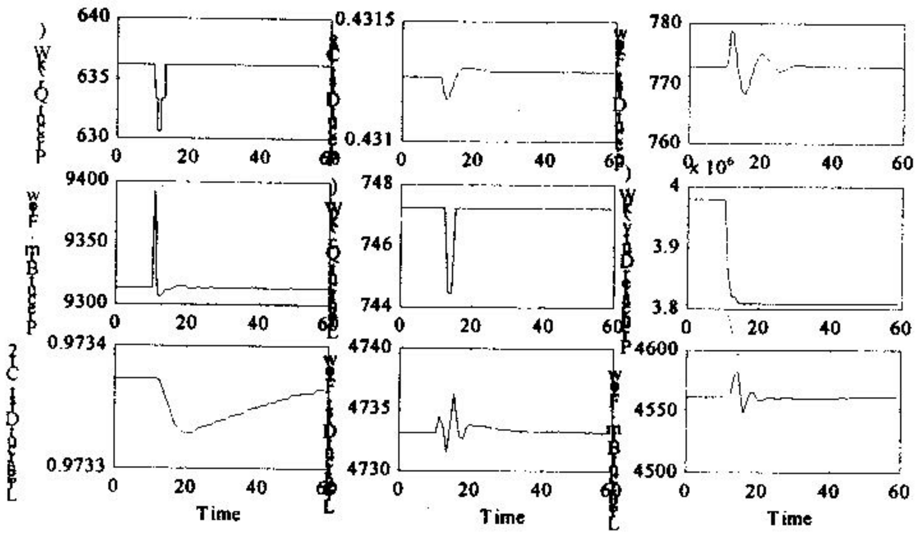

Figure 8: Responses of the process variables to step change of Raw-Feed temperature from 30 °C to 40 °C for case 2.

The effects of pre-heater control loop to the process are shown in Figure 9 and Figure 10. Here the setpoint of the preheater was increased from 210 °C to 220 °C. Similar trends as of figures 7 and 8 can be observed. Significant degradation of control performances are noticeable in most control loops proving the significant interaction resulting from the heat integration exercises. Therefore, a complete understanding of the process behaviour should be favoured before any integration is performed, a failure of which may lead in poor performance. At the limit, serious interaction may even lead to process instability.

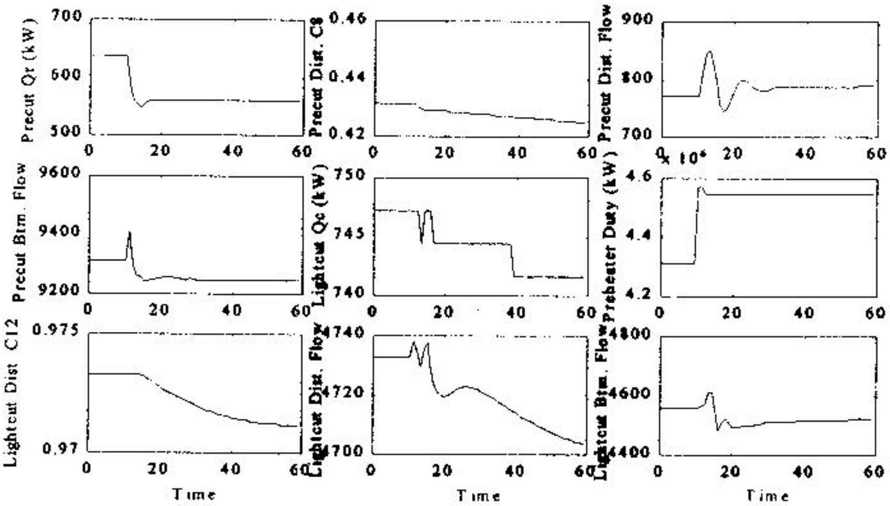

Figure 9: Responses to step change of Pre-heater Controller from 210 °C to 220 °C for case 1

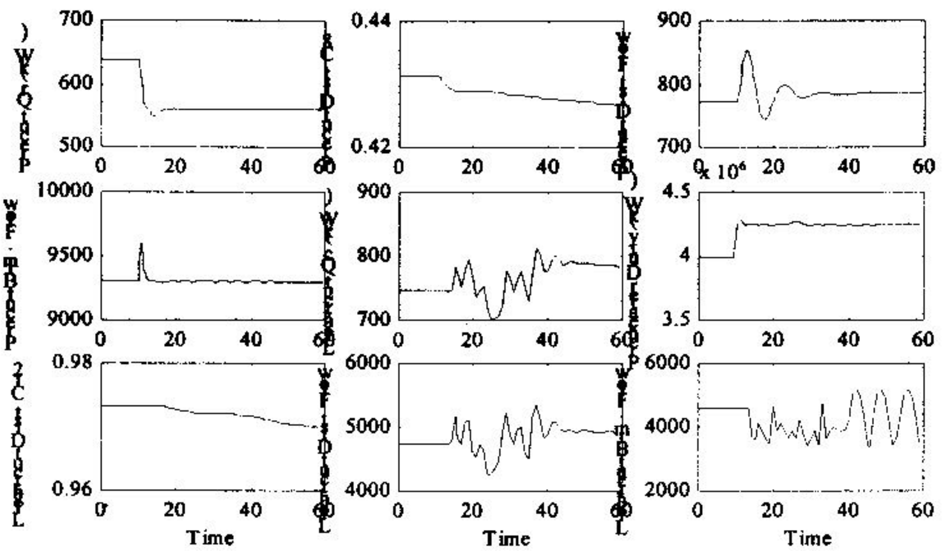

Figure 10: Responses to step change of Pre-heater Controller from 210 °C to 220 °C for case 2.

CONCLUDING REMARKS

The results presented here have pointed out a number of important conclusions. The dynamic simulation of an industrial oleochemical process has been successfully implemented using a commercial simulator. Thus, the model is applicable for further plantwide control system design. An application of this model is also presented to analyse the effect of heat integration on the dynamic characteristic of the process. The results show that with the aid of rigorous model, a better understanding of the process dynamic could be accomplished. As a result, a trade-off between the design and control could also be assessed qualitatively in the early stage of design.

REFERENCES

1. Umeda, T., Kuriyama, T. and Ichikawa, A, A logical structure for process control system synthesis, Proc. of IFAC Congress (Helsinki), 271-278 (1978).
2. Downs, J.J. and Vogel, E.F., A plantwide industrial process control problem, Computers & Chem. Eng., 17(3), 245-255 (1993).
3. Buckley, P.S., Techniques of Process Control, John Wiley & Sons (1964).
4. Morari, M., G. Stephanopoulos and Y. Arkun, Studies in the synthesis of control structures for chemical processes. Part I: Formulation of the problem, process decomposition and the classification of the control task. Analysis of the optimizing control structures, AIChE Journal 26(2), 220–232 (1980).
5. Price, R.M. and C. Georgakis, A plant wide regulatory control design procedure using a tiered framework, Ind. Eng. Chem. Res. 32(11), 2693–2705 (1993).
6. Luyben, W.L., B.D. Tyreus and M. Luyben, Plantwide Process Control, McGraw-Hill (1998).
7. Skogestad, S., Plantwide control: The search for the self-optimizing control structure. J. Proc. Control, 10, 487-507 (2000).
8. Gronenendijk, A.J., Dimian, A.C., and Iedema, P.D., System approach for evaluating dynamics and plantwide control of complex plants, AIChE Journal, 46(1), 133–145 (2000).
9. HYSYS.Plant User's Manual (2000). Hyprotech Ltd., Calgary, Canada.