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Biodiesel Production from Waste Cooking Oil: Plant-Wide Control System Design using Integrated Framework Approach

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Abstract— **Control system design for a complete plant with overall control perspective, is referred to as plant-wide control (PWC). Methodologies for this are of vital importance for safe, smooth and economical operation of plants. Increasing material recycles, energy integrations, product purity requirements and environmental regulations pose tough challenges to the smooth and stable plant operation. In this article, PWC structure is developed for a complete biodiesel plant using waste cooking oil as the raw material. Firstly, two process alternatives are developed and optimized for two objectives using the elitist non-dominated sorting genetic algorithm. Then, the better process is determined based on economic and environmental objectives. Later, PWC system is developed, based on an integrated framework of heuristics and simulation (IF), for the chosen process. This method makes effective use of rigorous process simulators and heuristics to aid in decision making while developing a PWC structure. Also, it is simple to apply with minimal computations other than process simulation. Performance of the developed control system is investigated in terms of settling time and deviation from the production target (DPT). The proposed PWC structure is found to be stable and robust in the presence of several expected disturbances.**

I. Introduction

Many plant-wide control (PWC) methodologies have been developed and implemented in several industrial processes [1-7]. Broadly, these methodologies can be classified into heuristics, optimization, mathematical, and mixed approaches [8]. Unlike relatively recent methodologies such as integrated framework of heuristics and simulation (IF) [2,3,8] and economic plantwide control [9-11], earlier methodologies do not make effective use of rigorous process simulators when developing a control system. IF methodology is easy to apply and involves minimal computations other than process simulation using a commercial simulator. Although PWC of industrial processes have been widely studied, PWC of a complete biodiesel process using waste cooking oil (WCO) is hardly found in the literature. Recently, Zhang et al. [12] developed PWC for the biodiesel process, which uses pure vegetable oil, and so they did not include esterification section required for feed with free fatty acids (FFAs). Availability and high cost of pure oil limit its use for biodiesel production. So, this study focuses on the use of waste cooking oil (WCO) or crude oil, where esterification should be carried out to convert FFAs, which otherwise may lead to saponification. We develop PWC of a complete biodiesel plant including esterification and trans-esterification of WCO. This brings down the cost of biodiesel production and also promotes sustainability as the process uses WCO.

The next section presents the design and optimization of the biodiesel process. Section 3 describes the application PWC methodology to the chosen process alternative. Section 4 discusses the performance of the control structure in terms of settling time and deviation from production target (DPT), defined in the Appendix. Finally, the article is concluded by outlining the conclusions.

II. Design and Optimization of Alternative Processes

For this study, biodiesel plant capacity is assumed to be 120,000 metric tons per annum, based on potential WCO in Malaysia. Both steady-state and dynamic simulations of the biodiesel process are developed using Aspen Plus V8.0 and Aspen Plus Dynamics V8.0 respectively. The property model used for these simulations is Dortmund modified UNIFAC. Unlike most previous papers, the present study considers detailed components of palm oil and more realistic kinetics that includes mono- and di-glycerides formation; esterification and trans-esterification are represented by 10 [13] and 96 reactions [14], respectively. Details of the oil constituents can be found in [14]. Composition of oil given in [14] is adjusted to include 6% FFAs. Two process alternatives for biodiesel production from WCO, are simulated and then optimized for two objectives using the elitist non-dominated sorting genetic algorithm (NSGA-II), implemented in MS Excel using VBA. Both the process alternatives use alkali catalyzed trans-esterification, which is more efficient and also used in industry (www.platinumgroup.com.my and www.lurgi.com/website/biodiesel.57.0.html?&L=1). Process 1 is based on the process flow sheet in Sharma and Rangaiah [15], where methanol removal is followed by water washing. Process 2 is based on the process flow sheet presented by Morais et al. [16], where water washing is followed by separation of products. To make these two alternatives comparable, some modifications are made. The two optimized processes are compared for both economic and environmental merits such as maximum profit, minimum heat duty and minimum organic waste. This comparison suggests that process 1 is better than process 2 in terms of both higher profit and lower environmental impact. Hence, process 1 is chosen for PWC study. The optimal values of process parameters, such as reactor temperatures, residence times and feed tray of distillation columns, are determined.

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III. PWC BASED ON INTEGRATED FRAMEWORK

PWC system is designed using IF methodology proposed by Konda et al. [17]. This multi-hierarchical methodology has eight levels, where steady-state and dynamic model of the plant are used along with heuristics to make the decisions on control system design. In addition, control decisions based on heuristics are also validated using dynamic simulations. Each level is briefly described as follows.

A. *Level 1.1: Define PWC Objectives*

In the first step, PWC objectives are defined. Note that different objectives may lead to different control structures. Typically, these objectives include product rate, product quality, process/equipment constraints, stable control and environmental constraints. For present plant, PWC objectives are: 1) constant production rate at normal operation with quick and smooth performance for throughput changes, 2) product purity (bio-diesel > 99% as per EN 14214 standards and glycerol > 95%), 3) maintaining reboiler temperature of biodiesel processing distillation column below 250^{o}C and glycerol processing distillation column below 150^{o}C, and 4) maintaining methanol to oil ratio (6:1 molar ratio at normal condition) and methanol split fraction (RTRANS1: RTRANS2: RTRANS3= 0.77:0.12:0.11, at normal condition) to achieve the EN standards for biodiesel.

B. *Level 1.2: Determine Control Degree of Freedom (CDOF)*

Konda et al. [18] proposed the restraining number method to determine CDOF. It uses unit operations in the process flow sheet (without any valves) to determine CDOF. Using this method, CDOF for the biodiesel plant is found to be 83. This large number of CDOF is due to many unit operations and streams involved in the process (see Fig. 1).

C. *Level 2.1: Identify and Analyze Plant-wide Disturbances*

Informed understanding of possible disturbances in the process has a favorable effect on the control scheme development and controller tuning. The steady-state simulator is used to try and test the effect of anticipated disturbance. It is observed that ±10% change in WCO leads to nearly ±10% variation in the recycle streams and product flow rates. Table I shows the anticipated disturbances having considerable effect on the biodiesel plant. Impurities are found to be under control for all disturbances.

TABLE I. ANTICIPATED DISTURBANCES IN THE BIO-DIESEL PROCESS

No.	Disturbance	Magnitude
D1	Feed oil flow rate	+10%
D2		-10%
D3	Catalyst deactivation (pre-exponential factor of reactions converting tri-glycerides to biodiesel)	-10% (RTRANS1, RTRANS2 and RTRANS3)
D4	Dual disturbances	+5 % Feed oil flow rate and D3
D5	Dual disturbances	D2 and D3

D. *Level 2.2: Set Performance and Tuning Criteria*

In this step, settling time is chosen as the performance criteria. At this stage, flow, level, and pressure controllers are tuned based on the guidelines in [19]; other controllers are tuned using Autotuning tools in Aspen Plus Dynamics. Some of the controllers are fine-tuned in the later stage. Controllers having time lags are tuned using closed-loop tuning method. Tyreus-Luyben criteria are used to determine the tuning parameters for such control loops. Controllers having no time lags are tuned using the open-loop tuning method; Cohen-Coon method is used to determine their tuning parameters.

E. *Level 3.1: Product Rate Manipulator Selection*

This step deals with the identification of primary process path from the main raw material to the main product. As the reactor conditions are fixed by optimization, these should not be used as throughput manipulators (TPM). Steady-state simulation can be used to identify other options. Based on the steady-state gain obtained from steady-state simulations, feed oil flow rate is identified as the next best choice for TPM.

F. *Level 3.2: Product Quality Manipulator Selection*

Product purity is one of the important controlled variables for the biodiesel plants as the main product i.e. biodiesel should meet EN 14214 standards. Hence, biodiesel purity and impurity levels, such as tri-, di-, mono-glycerides in the final product should be monitored. Although impurities are found to be below the permissible limit for all disturbances, tri-glyceride content in the final product is found to be sensitive. Hence, reaction conditions are maintained to consume almost all oil. Consequently, methanol ratio is decided in the ratio controllers ‘RC100’ and ‘RC200’ through the cascade loop to maintain FFAs and tri-glyceride impurity in the final product below the permissible limit. Methanol content in the final product is controlled by manipulating the wash water flow rate. Glycerol purity also has to be maintained at its desired value. For this, a cascade loop is implemented to manipulate the reboiler duty of FRAC-4. An additional constraint also has to be satisfied as the reboiler temperature should not increase beyond 150^{o}C to avoid glycerol decomposition.

G. *Level 4.1: Selection of Manipulators for More Severe Controlled Variables*

This step deals with process constraints related to equipments, operation, safety, environment and stability. The important constraints in the biodiesel process are as follows. 1) Reboiler temperatures, T_{FRAC-1} and $T_{FRAC-4} \leq 150^oC$, and T_{FRAC-2} and $T_{FRAC-3} \leq 250^oC$: these temperatures are allowed to vary within acceptable limits. However, the controller becomes active when the reboiler temperature reaches the limit, which is given as the remote set point for the respective controller. 2) Methanol to FFA and Methanol to tri-, di-, mono-glycerides ratios: fresh methanol is manipulated to maintain the required methanol ratio in RC100 and RC200. Similarly, ratio controllers, RC101 and RC201 are also implemented to maintain the ratio of sodium hydroxide and sulfuric acid. 3) Methanol split ratios for the CSTRs: these are maintained using controllers SP200 and SP201. 4) CSTR temperatures: optimal values given by optimization have to be maintained. The reactor duty of each CSTR is manipulated to control the respective reactor temperature.

H. Level 4.2: Selection of Manipulators for less Severe Controlled Variables

This step, in particular, deals with level and pressure controllers. A proper level control is required as level is often integrating. Although P-only controller is enough to

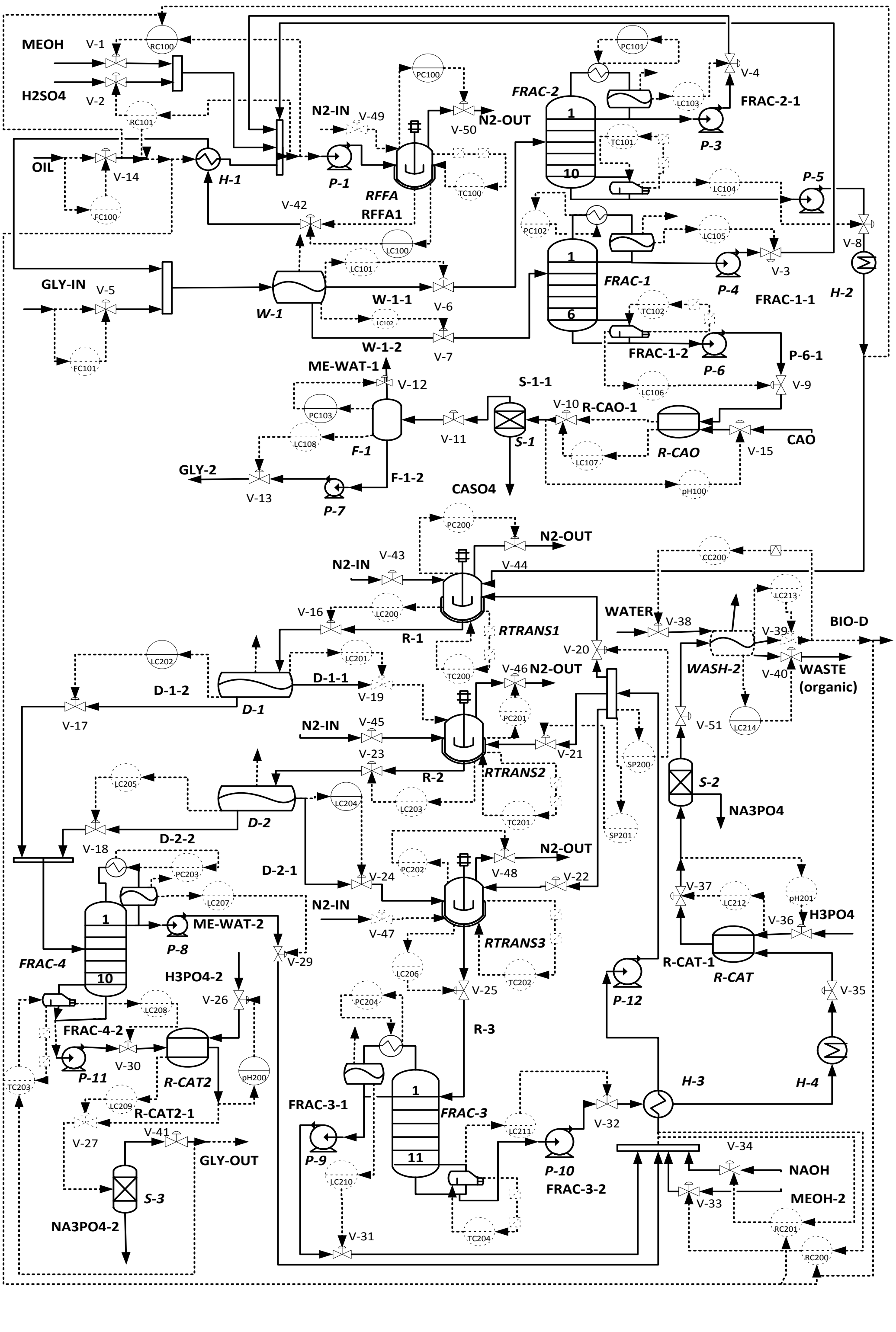

Figure 1. PWC scheme designed for the biodiesel plant using waste cooking oil as the raw material.

TABLE II. CONTROL STRUCTURE OBTAINED AND CONTROLLER PARAMETERS FOR THE BIODIESEL PLANT

Controller	Controlled variable	Manipulated variable [valve number shown refers to the valves in Fig. 1]	Parameters: Kc (%/%), τ_i (min)
Esterification Section (21 controllers)			
FC100	Bio-diesel production rate	Inlet oil flow (TPM) [V-14]	0.5; 0.3
FC101	Glycerol flow	Inlet glycerol flow [V-5]	0.5; 0.3
RC100	Methanol to FFAs ratio {remote set point based on composition of FFAs in stream BIO-D}	Fresh methanol flow [V-1]	0.5; 0.3
RC101	Sulfuric acid to FFAs ratio	Inlet sulfuric acid flow [V-2]	0.5; 0.3
PC100	Pressure in RFFA	Outlet N_2 flow [V-50]	20; 10
PC101	Condenser pressure in FRAC-2	Condenser duty in Frac-2	20; 12
PC102	Condenser pressure in FRAC-1	Condenser duty in Frac-1	20; 12
PC103	Pressure in F-1	Vapor flow rate [V-12]	20; 12
TC100	Temperature in RFFA	Heat duty in RFFA	4.6; 9.24
TC101	Reboiler temperature in FRAC -2 {remote set point}	Reboiler duty in FRAC-2	2.9; 9.24
TC102	Reboiler temperature in FRAC-1 {remote set point}	Reboiler duty in FRAC-1	22.94; 2.64
LC100	Level in RFFA	Liquid outlet flow [V-42]	10; 60
LC101	Light phase level in W-1	Light phase outlet flow [V-6]	10; 60
LC102	Heavy phase level in W-1	Heavy phase flow [V-7]	10; 60
LC103	Reflux drum level in FRAC-2	Distillate flow [V-4]	2; 20
LC104	Reboiler level in FRAC-2	Bottoms flow [V-8]	2; 20
LC105	Reflux drum level in FRAC-1	Distillate flow [V-3]	2; 20
LC106	Reboiler level in FRAC-1	Bottoms flow [V-9]	2; 20
LC107	Level in R-CAO	Liquid outlet flow [V-10]	10; 60
LC108	Level in F-1	Liquid outlet flow [V-13]	10; 60
pH100	pH of stream R-CAO-1	Inlet calcium oxide flow [V-15]	1; 20
Trans-esterification Section (32 controllers)			
RC200	Methanol to (TG+DG+MG) ratio {remote set point based on composition of TG in stream BIO-D}	Fresh methanol flow [V-33]	0.5; 0.3
RC201	Sodium hydroxide to (TG+DG+MG) ratio	Inlet sodium hydroxide acid flow [V-2]	0.5; 0.3
PC200	Pressure in RTRANS1	Outlet N_2 flow [V-44]	20; 12
PC201	Pressure in RTRANS2	Outlet N_2 flow [V-46]	20; 12
PC202	Pressure in RTRANS3	Outlet N_2 flow [V-48]	20; 12
PC203	Condenser pressure in FRAC-4	Condenser duty in FRAC-4	20; 12
PC204	Condenser pressure in FRAC-3	Condenser duty in FRAC-3	20; 12
TC200	Temperature in RTRANS1	Heat duty in RTRANS1	31.47; 17.92
TC201	Temperature in RTRANS2	Heat duty in RTRANS2	34.8; 13.96
TC202	Temperature in RTRANS3	Heat duty in RTRANS3	30.8; 15.28
TC203	Reboiler temperature in FRAC-4 {remote set point based on composition of glycerol in stream GLYC-OUT}	Reboiler duty in FRAC-4	1.4; 7.92
TC204	Reboiler temperature in FRAC-3 {remote set point}	Reboiler duty in FRAC-3	3.63; 7.92
LC200	Level in RTRANS1	Liquid outlet flow [V-16]	10; 60
LC201	Light phase level in D-1	Light phase outlet flow [V-19]	34.42; 60
LC202	Heavy phase level in D-1	Heavy phase flow [V-17]	10; 60
LC203	Level in RTRANS2	Liquid outlet flow [V-23]	81.14; 60
LC204	Light phase level in D-2	Light phase outlet flow [V-24]	26.58; 60
LC205	Heavy phase level in D-2	Heavy phase flow [V-18]	10; 60
LC206	Level in RTRANS3	Liquid outlet flow [V-25]	146.74; 60
LC207	Reflux drum level in FRAC-4	Distillate flow [V-29]	2; 20
LC208	Reboiler level in FRAC-4	Bottoms flow [V-30]	2; 20
LC209	Level in R-CAT2	Liquid outlet flow [V-27]	10; 60
LC210	Reflux drum level in FRAC-3	Distillate flow [V-31]	2; 20
LC211	Reboiler level in FRAC-3	Bottoms flow [V-32]	2; 20
LC212	Level in R-CAT	Liquid outlet flow [V-37]	10; 60
LC213	Light phase level in WASH-2	Light phase outlet flow [V-39]	10; 60
LC214	Heavy phase level in WASH-2	Heavy phase flow [V-40]	10; 60
pH200	pH of stream R-CAT2-1	Inlet phosphoric acid flow [V-26]	0.5; 0.3
pH201	pH of stream R-CAT-1	Inlet phosphoric acid flow [V-36]	1; 20
CC200	Methanol composition in stream BIO-D (active only when the limit is exceeded)	Wash water flow rate [V-38]	0.5; 0.3
SP200	Methanol split ratio to RTRANS1	Methanol flow rate to RTRANS1 [V-20]	7.01; 0.59
SP201	Methanol split ratio to RTRANS2	Methanol flow rate to RTRANS1 [V-21]	1.91; 0.59

obtain satisfactory performance, PI controller is implemented to obtain tight control. Based on heuristics, inventory should be controlled in the direction of flow. Therefore in all distillation columns, level in reflux drum and in reboiler is controlled using distillate flow and bottoms flow respectively. Also, liquid levels in CSTRs and phase separators are controlled using liquid outlets as shown Fig.1. The pressure in CSTRs is maintained by manipulating the inert gas outlet flow. The pressure in all distillation columns is controlled using respective condenser duty as suggested by heuristics; these are also verified using dynamics simulations.

I. Level 5.0: Control of Unit Operations

Control of individual unit operations is dealt with in this step. Basic control of the most common processes is well established as given in [19]. All level and pressure control loops are already decided in the previous steps. Temperature control of CSTRs and distillation columns is also taken care of in the level 4.2. Overall, unit-wise inventory is observed to be well regulated; hence, no more control loops are implemented. In addition to the level control in the neutralization reactors, pH of outlet stream is controlled using inlet calcium oxide in R-CAO and using inlet phosphoric acid in both R-CAT and R-CAT2.

J. Level 6.0: Check Component Material Balance

It is necessary to ensure that the component inventory is well regulated. Plant-wide accumulation of all components should be calculated and observed. If required, unit-wise accumulation can be determined to investigate if further improvements are required. Component balances are therefore checked to ensure minimal accumulation. Negligible accumulation suggests that the inventory is well regulated.

K. Level 7.0: Effects of Integration

The dynamics of the process should be studied for the anticipated disturbances, both with and without recycles closed. It is done by observing (i) the overall accumulation profile of WCO in a complete plant, (ii) dynamics of process and (iii) effect on important process parameters such as conversion, production rate etc. Fig. 2 shows that the accumulation is relatively more when there is recycle.

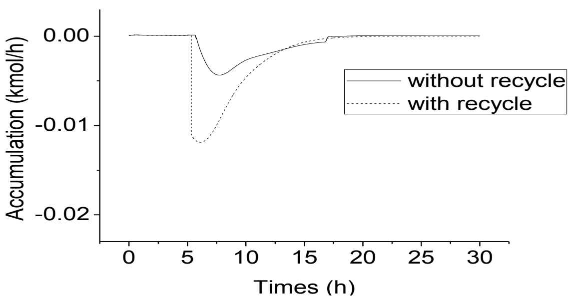

Figure 2. WCO accumulation due to disturbance D1, with and without recycle.

No significant change is noticed in terms of settling time of biodiesel flow rate, which suggests that plant dynamics are not significantly affected. Conversion and product flow rate are not affected after closing the recycles as the parameters affecting these such as temperature, pressure and methanol to oil ratio in CSTRs, are already taken care in the previous steps. Also, note that the change in the WCO flow rate leads to proportionate variation in the recycle streams, as found in level 2.1. To conclude, the effects of integration are not severe, and hence no modification is warranted in the control scheme.

L. Level 8.0: Enhance Control System Performance with Remaining CDOF

If required, remaining CDOFs can further be used to enhance the control structure performance. As the developed control structure is adequate, no further improvement is required. The obtained control structure by the above IF methodology is presented in Fig. 1. Table II presents all the controllers and their tuning parameters. The percentage opening of control valves for the base case operation is about 50%. However, as this model is based on pressure-flow solver (i.e. pressure driven simulation), where pressure depends on upstream conditions, valve opening may marginally deviate from the design opening, as was found in [12].

IV. EVALUATION OF CONTROL SYSTEM

PWC based on IF is developed and successfully implemented in the biodiesel plant, as shown in Fig. 1. Plant performance is tested for the disturbances D1 to D5 (Table I). It is quantified in terms of settling time (i.e. time required for the production rate to reach within 1% of the target) and absolute DPT, as described by Vasudevan and Rangaiah [20]. Initially, the plant is allowed to run for 5 h, after which the disturbances are introduced, one at a time. Table III shows the results for the disturbances D1 to D5. For this plant having capacity of 120,000 tons per annum, the settling time for all disturbances is about 10 h, which is in line with the settling time of about 10-20 h for the biodiesel plant having the approximate capacity of 200,000 tons per annum [12]. Disturbance D3 has a very small DPT as -10% change is introduced only in the pre-exponential factor of reactions converting tri-glycerides to biodiesel. DPT can be expected to be larger if the change is made in all reactions producing biodiesel. DPT for other disturbances is comparable (Table III).

TABLE III. PERFORMANCE OF PWC DESIGNED BY IF METHODOLOGY

No.	Performance based on	
	Settling time (h)	*DPT (kg)*
D1	9.1	1643.6
D2	10.1	1614.5
D3	9.2	21.43
D4	9.3	1662.7
D5	10.5	1984.8

Fig. 3 depicts the accumulation of oil in the presence of disturbances D1 and D2. For brevity, the accumulation of only WCO is observed and shown. If required, accumulation for the complete plant as well as for individual unit operations can be monitored to check if any inventory loop has been left out. Fig. 3 clearly shows that the accumulation of WCO reaches to zero after certain amount of time. TG impurity in biodiesel due to the disturbances D4 and D5 is shown in Fig. 4. TG impurity is observed to be below its permissible limit as per the EN standards even though a small rise is observed for the increased flow rate of WCO. Additionally, the control scheme is found to provide satisfactory performance for -20% change in WCO; these are not shown here for brevity. Overall, the plant is stable and performing well under the PWC designed by IF methodology.

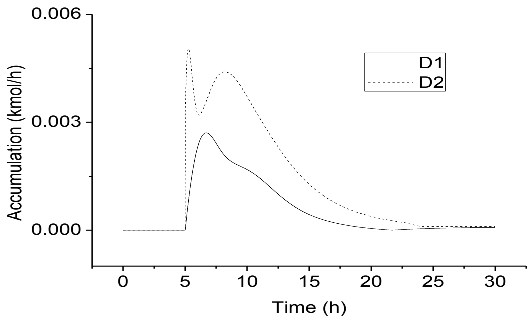

Figure 3. WCO accumulation due to disturbances D1 and D2.

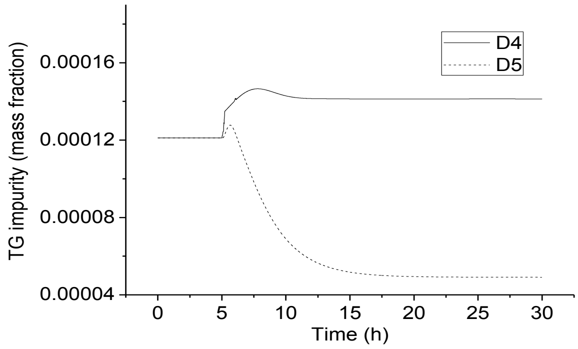

Figure 4. Tri-glyceride impurity in biodiesel due to disturbances D4 and D5.

V. CONCLUSION

In this study, PWC for the biodiesel process using WCO is investigated. First, multi-objective optimization using NSGA-II is carried out to determine the better process alternative, and also to find the optimal values of process design and operating variables. PWC for the chosen process is developed based on the integrated framework of heuristics and simulation (IF), and then implemented successfully. The performance of the designed control system is investigated in terms of settling time and deviation from the production target. The control system is found to provide smooth and stable control. Future work will study control system design by another methodology for comparison with that presented in this article.

APPENDIX

DPT: DPT is an indirect economic measure in terms of production rate. Smaller is the DPT, better is the control system. Vasudevan and Rangaiah [20] defined DPT as:

$$DPT = \int_{0}^{t_s} (P_A - P_T)dt$$

Here, P_A is actual production rate, P_T is production target and t_s is settling time.

REFERENCES

[1] G. Y. Zhu and M. A. Henson, "Model predictive control of interconnected linear and non-linear processes," *Ind. Eng. Chem. Res.,* vol. 4, pp. 801-816, 2002.

[2] S. Vasudevan, G. P. Rangaiah, and N. V. S. N. M. Konda, W. H. Tay, "Application and evaluation of three methodologies for plant-wide control of the styrene monomer plant," *Ind. Eng. Chem. Res.,* vol. 48, pp. 10941-10961, 2009.

[3] S. Vasudevan and G. P. Rangaiah, "Integrated framework incorporating optimization for plant-wide control of industrial processes," *Ind. Eng. Chem. Res.*, vol. 50, pp. 8122–8137, 2011.

[4] G. Herrmann, S. K. Spurgeon and C. A. Edwards, "Model-based sliding mode control methodology applied to the HDA plant," *J. Process Contr.*, vol. 13, pp. 129-138, 2003.

[5] A. C. B. Araujo, M. Govatsmark and S. Skogestad, "Application of Plant-Wide Control to the HDA Process. I - Steady-State Optimization and Self-Optimizing Control. *Control Eng. Pract.*, Vol. 15, pp. 1222-1237, 2007.

[6] M. L. Luyben, B. D. Tyreus and W. L. Luyben, "plant-wide control design procedure," *AIChE J.* vol. 43, pp. 3161-3174, 1997.

[7] M. A. Al-Arfaj and W. L. Luyben, "Plant-wide control for tame production using reactive distillation," *AIChE J.* vol. 50, pp. 1462–1473, 2004.

[8] S. Vasudevan, N. V. S. N. M. Konda and G. P. Rangaiah, "Plant-wide control: methodologies and applications," *Rev. Chem. Eng.*, vol. 25 (S6), pp. 297-337, 2009.

[9] S. Skogestad, "Control structure design for complete chemical plants," Comput. Chem. Eng., vol. 28, pp. 219-234, 2004.

[10] R. Jagtap, N. Kaistha and S. Skogestad "Plantwide control for economic operation of a recycle process with side reaction," *Ind. Eng. Chem. Res.*, vol. 50, pp. 8571-8584, 2011.

[11] S. Skogestad, "Economic plantwide control," In Plantwide control-recent developments and applications , G. P. Rangaiah and V. Kariwala, Ed. Wiley, 2012, pp. 229-251.

[12] C. Zhang, G. P. Rangaiah and V. Kariwala, "Design and plantwide control of a biodiesel plant," In *Plantwide control-recent developments and applications* , G. P. Rangaiah and V. Kariwala, Ed. Wiley, 2012, pp. 293-317.

[13] M. Berrios, J. Siles, M. A. Martín, A. A. Martín, "Kinetic study of the esterification of free fatty acids (FFA) in sunflower oil," *Fuel*, vol. 86(15), pp. 2383-2388, 2007.

[14] Aspen Technology. Aspen Plus - Aspen plus biodiesel model (Examples). 2012.

[15] S. Sharma and G. P. Rangaiah, "Multi-objective optimization of a bio-diesel production process," *Fuel, vol.* 103, pp. 269-77, 2013.

[16] S. Morais, T. M., A. A. Martins, G. A. Pinto and C. A. V. Costa, "Simulation and life cycle assessment of process design alternatives for biodiesel production from waste vegetable oils," *J. Clean Prod.*, vol. 18(13), pp. 1251-1259, 2010.

[17] N. V. S. N. M. Konda, G. P. Rangaiah and P. R. Krishnaswamy, "Plant-wide control of industrial processes: an integrated framework of simulation and heuristics," *Ind. Eng. Chem. Res.*, vol. 44, pp. 8300-8313, 2005.

[18] N. V. S. N. M. Konda, G. P. Rangaiah, P. R. A. Krishnaswamy, "Simple and effective procedure for control degrees of freedom," *Chem. Eng. Sci.*, vol. 61, pp. 1184-1194, 2006.

[19] W. L. Luyben, *Plant-Wide Dynamic Simulators in Chemical Processing and Control*. New York: Marcel Dekker, 2002, ch. 3.

[20] S. Vasudevan and G. P. Rangaiah, "Criteria for performance assessment of plant-wide control systems," *Ind. Eng. Chem. Res.*, vol. 49, pp. 9209-9221, 2010.