(RESEARCH ARTICLE)

Heat-integrated process improvement of petroleum refinery FCC light-ends separation

Dauda Ibrahim *

Department of Chemical Engineering, University of Maiduguri, P.M.B. 1069 Maiduguri, Borno State, Nigeria.

Abstract

This research presents a methodology for investigating and proposing operational changes that can improve product recovery, heat recovery, and economics of a gas concentration unit of a FCC plant. The methodology starts by collecting relevant information from an existing plant, followed by development of a simulation model of a gas concentration unit in Aspen HYSYS. Next, the developed model is used to conduct optimisation studies to identify process changes (solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling) that improves product recovery, heat recovery, and economics. Key findings from this research includes the following trade-offs: increasing operating pressure from 1300 kPa to 1700 kPa increases gasoline and LPG recovery by 1% and 0.4% respectively whilst heat recovery and cooling water demand increases by 0.5% and 0.1%, leading to total benefit increase of 1%; increase in recycled gasoline flow rate by 9628 kmol/h (20% of initial flow rate) improves recovery of gasoline and LPG by 0.1% and 0.3% and heat recovery and cooling water demand by 18% and 77%, but steam generation decreases by 35%, which results to decrease in benefit of 1%; increasing C_6 content of solvent (unstabilized naphtha) by 50 kmol/h increase in gasoline recovery 7.5% and heat recovery and cooling water demand increases by 0.1% and 1%, and benefit increases by 5%; for ambient cooling medium, varying solvent temperature and pumparound return temperature (inter-stage cooling) does not improve product recovery and heat recovery. Analysis of results indicate that high C_6 composition in solvent has the highest impact on product recovery, heat recovery, and process economics compared with other degrees of freedom.

Keywords: Product recovery; Heat recovery; Process modelling; Sensitivity analysis; Fluidized catalytic cracking unit

1. Introduction

The efforts to maximise the use of crude oil to generate more revenue by producing more valuable end products such as LPG, gasoline, diesel etc., and petrochemical feedstock such as propylene, ethylene etc., necessitates the utilisation of different petroleum by-products such as light-ends (wet gas), gas oil etc. 'Light-ends' is a general term referring to all gaseous and more volatile liquid hydrocarbons produced in a refinery (Speight et al., 2009). Light-ends are low boiling hydrocarbons such as methane, ethane, propane, butanes, pentanes and heavier hydrocarbons (Speight et al., 2009) with a small fraction of non-hydrocarbon compounds such as H_2S, CO_2, H_2O etc. The composition of light-end gases depends on source of production (Speight et al., 2009). Refinery units producing light-ends include atmospheric distillation tower, hydrotreaters, hydrocrackers, catalytic reformer, fluidized catalytic cracker and delayed coker (Gary et al., 2007). The latter two units produce light-end hydrocarbons composed of unsaturated hydrocarbons, mainly olefins such as ethylene, propylene, butylenes and pentylene (Gary et al., 2007); while the remaining units produce saturated light-end hydrocarbons.

---

* Corresponding author: Dauda Ibrahim

The separation of cracked hydrocarbons from effluent of a FCC reactor into valuable products such as high-octane gasoline, LPG, diesel, heavy cycle oil (HCO) and slurry takes place in downstream processes which consist of FCC main fractionator and gas concentration unit. The prior separation of reactor products into light cycle oil (LCO), HCO, slurry, unstabilized naphtha and wet gas (light-ends) occurs in the FCC main fractionator (Lucas, 2000), which is part of the fluid catalytic cracking plant. The gas concentration unit or FCC light-ends separation unit separates light-ends into gasoline, LPG and fuel gas (Sadeghbeigi, 2000) in a series of compression, absorption, desorption and distillation processes. A typical gas concentration unit configuration showing all the major equipment involve in the separation process is presented in **Figure 1.**
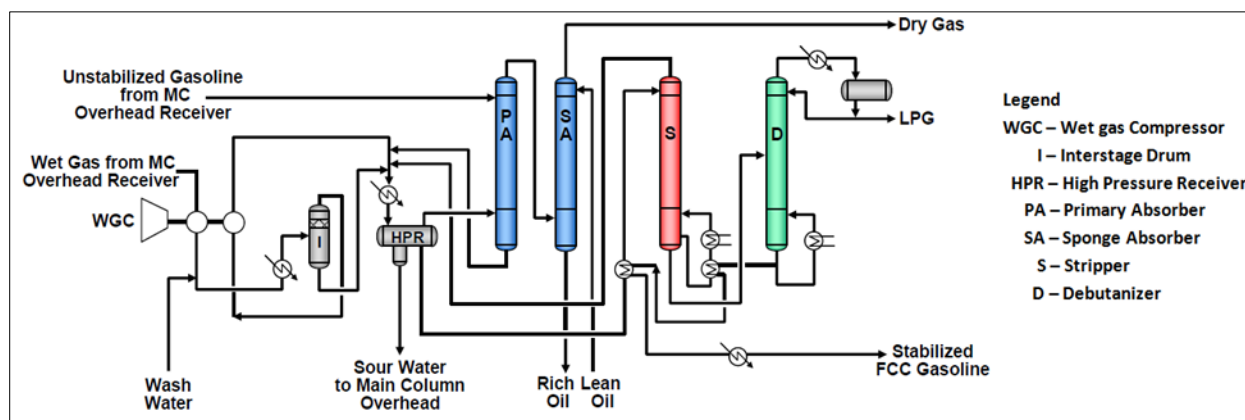
Figure 1 Gas concentration unit of fluid catalytic cracking plant (Wolsehlag et al., 2009)

The gas concentration unit shown in Figure 1 includes a primary absorber, sponge/ or secondary absorber, stripper or de-ethanizer and debutanizer (Wolsehlag et al., 2009). Some gas concentration unit include a gasoline splitter and C_3/C_4 splitter (Sadeghbeigi, 2000). Detailed discussion of the major units and other equipment on the flow sheet can be found elsewhere. The light-ends from the FCC main fractionator contain substantial amount of propane plus heavier (C_3+) hydrocarbons, which are recovered into LPG and gasoline in the Gas Concentration Unit (GCU) of the FCC plant, through a series of compression, absorption, desorption and distillation steps (Dean et al., 2005). The heart of the GCU is the primary absorber, where the C_3+ hydrocarbons are selectively absorbed from the light-ends exiting the FCC main fractionators, and are further separated into high-octane gasoline (pentane and heavier hydrocarbons or C_5+) and LPG. Lu et al., (2000) reported a relatively high loss of C_3+ hydrocarbon, around 22%, of the overhead product from the primary absorber, which is uneconomical and reduces the amount of petroleum refinery end-products derived from C_3+ hydrocarbons. Hence, improving the recovery of C_3+ hydrocarbons in the primary absorber would not only increase refinery net profit but also enable refiners to meet the growing demand for petroleum refined products, for example, gasoline and LPG.

The FCC plant is an energy-intensive process requiring huge amounts of energy to preheat the feed and medium pressure steam for atomizing the raw feed oil (Sadeghbeigi, 2000); also, stripping steam is required in the FCC reactor. On the other hand, the gas concentration unit requires cooling at different locations, which includes compressed wet gas cooler, gasoline cooler, debutanizer overhead condenser etc. (Al-Riyami et al., 2001). Therefore, heat recovery between the FCCU and gas concentration unit reduces the amount of steam required from utility system to satisfy process heating (Wolsehlag et al., 2009), thereby saving huge amounts of energy and greenhouse gas reduction (Bumbac et al., 2008).

Improvement of recovery of various light-end components has been researched by various authors (Shailendra et al., 2012; Schultz et al., 2011; Dean et al., 2005; Haik 2005; Vasconcelos et al., 2005; Lu et al., 2000; Tagamolila et al., 1994). Research in this area has considered revamping the process to improve gasoline and LPG recovery and, in some cases, only recovery of LPG components is enhanced. However, most authors neglect considering the effect of enhancing gasoline and LPG recovery on process heat recovery. Therefore, improving heat recovery could reduce operating cost and increase net profit. Similarly, greenhouse gas emissions could be reduced substantially. Moreover, changes in operational conditions such as temperatures and pressures together with solvent composition affect "heat of absorption" in the primary absorber, which leads to more or less heat being released during absorption. Similarly, the change in flow rate of absorbent affects heat duties and heat recovery in the process. To establish a more economic operation, the enhancement of gasoline and LPG recovery has to be considered simultaneously with heat recovery to reduce utility consumption in the FCC plant.

Some authors focus solely on improving heat recovery in the gas concentration unit. Al-Riyami et al. (2001) implements the pinch methodology to an existing gas concentration unit of a FCC plant to enable heat recovery between the gas concentration unit and the FCC main fractionator. The authors use extracted stream data from an existing heat exchanger network to evaluate the current utility consumption and thereafter use the data for energy targeting and identification of the most promising design among alternative designs generated using SPRINT software. They identified design offers 74% savings in hot utility consumption which corresponds to 27% utility cost saving. Only fixed operating conditions are considered. In order derive more benefit from the process (in terms of product yield and energy cost), the approach should also consider enhancement of gasoline and LPG recovery. Bumbac et al. (2007) consider heat integration between a crude distillation unit and a fluid catalytic cracking unit. The authors extract stream data from the two units to determine the minimum cooling and heating required by those plants. They further explore the option of generating steam for power production using steam a turbine. However, they predicted 9% savings of hot utility and generated 17.5 kW of power. Thus, the approach has not considered improving heat recovery simultaneously with gasoline and LPG recovery in the gas concentration unit.

Golden and Fulton (2000) describe a low-cost method to improve fluid catalytic cracking unit energy efficiency using a revamp approach. The authors used energy efficient improvement options, process flow scheme changes, while considering equipment limits to identify solutions. Their approach involves evaluating alternative flow schemes together with practical constraints related to both the equipment and unit configuration, but product recovery is not enhanced. Ideally a process revamp should also consider enhancement of gasoline and LPG recovery. Wilson and Sloan (1984) revamped the main fractionator and gas concentration unit to enable conservation of energy through heat integration. In their revamp, they eliminated the debutanizer and absorber/stripper steam reboilers by using the main fractionator intermediate pumparound and top pumparound as heat sources to satisfy these services. Additionally, they install a new (light cycle oil) pumparound which was used together with the top pumparound to provide heat to the incoming feed. They conserved a significant amount of energy in the process with only feed preheat stream requiring heating using external hot utility. Consequently, the process has been more intensively heat integrated. Interactions between the main fractionator and debutanizer, absorbers and stripper could be challenge for operability and control. The approach does not take into account opportunities to increase product yield.

In the literature reviewed so far on improving gasoline and LPG recovery and heat recovery opportunities in gas concentration section of an FCC plant, the authors focus either on LPG recovery or heat recovery opportunities; no research was found to have considered the effect on heat recovery while improving the recovery of gasoline and LPG in the primary absorber. Furthermore, no research was found that critically examined which degrees of freedom have significant impact on recovery of gasoline and LPG. This work proposes a methodology that examines gasoline and LPG recovery in the primary absorber in a gas concentration unit, and also investigates the degrees of freedom that have significant impact on gasoline and LPG recovery, while considering (simultaneously) heat recovery opportunities and process economic. Using information from a base case plant, a model of the gas concentration unit is built in Aspen HYSYS. Degrees of freedom (solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling) are used to optimise the performance of the gas concentration unit, i.e., improving recovery of gasoline and LPG in the primary absorber. Pinch analysis is applied to identify and analyze opportunities for heat recovery between the gas concentration unit and the FCCU. Formulate general conclusions and to develop an understanding degree of freedom that can significantly impact heat recovery and yield of gasoline and LPG in the gas concentration unit.

This paper consists of three sections in addition to introductory section. Section 2 describes the base case process and presents a stepwise procedure/methodology for modelling and simulating the GCU. The procedure used for investigating the impact of degrees of freedom on product and heat recovery is also described. In Section 3, modelling results from case studies proposed in Section 2 are analysed and discussed. In addition, this section examines results from effects of degrees of freedom (such as operating pressure, solvent temperature, compositions and flow rate) on product recovery, heat recovery, and economics. Section 4 presents conclusions on the effect of the degrees of freedom on enhancing product recovery, heat recovery, and economics.

2. Materials and methods / Research methodology

The method adopted to fully address the research goals presented in **Section 1** includes developing a model of the gas concentration unit (using data for a base case plant) in Aspen HYSYS, examining the effect of the degrees of freedom affecting product (gasoline and LPG) recovery using the gas concentration unit model, exploring heat recovery opportunity in the gas concentration unit and applying economic analysis to identify the most profitable case using information from the product recovery and the heat recovery analysis. The proposed methodology is shown in **Figure**

2, which comprises four main steps: data collection, process modelling and simulation, process optimisation and heat integration, and process economic evaluation.
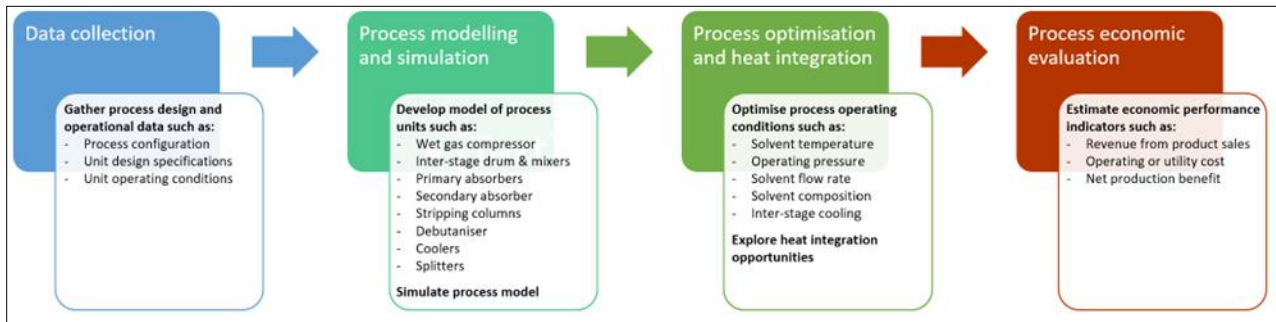

Figure 2 Proposed research methodology for improving product recovery simultaneously with heat integration and economic performance of petroleum refinery FCC light-ends separation

Step 1, data collection, involves gathering relevant process information needed to perform process improvement study, for example, process configuration, unit design specification, and unit operating conditions. In Step 2, data collected in Step 1 are used to build a simulation model of a gas concentration unit in Aspen HYSYS, and the built model is used in Step 3 to conduct optimisation of operating conditions as well as exploring opportunities for heat integration. Step 4 conducts economic evaluation of alternative cases and selects the most cost-effective case. **Subsections 2.1, 2.2, 2.3**, and 2.4 present details of data collection, process modelling and simulation, process optimisation and heat integration, and process economic evaluation respectively.

2.1. Data collection

Process data used in this investigation are collected from a base case plant in Aspen HYSYS and presented in this section. Input data extracted from the base case FCC plant includes the following: process flow diagram of the gas concentration unit, flow rate and composition of streams in the gas concentration unit, operating condition such as temperatures and pressures, and equipment specifications. The process flow diagram of the base case gas concentration unit is shown in **Figure 3**, details of stream flow rate, temperatures, pressures, equipment specifications, and stream compositions can be found elsewhere.
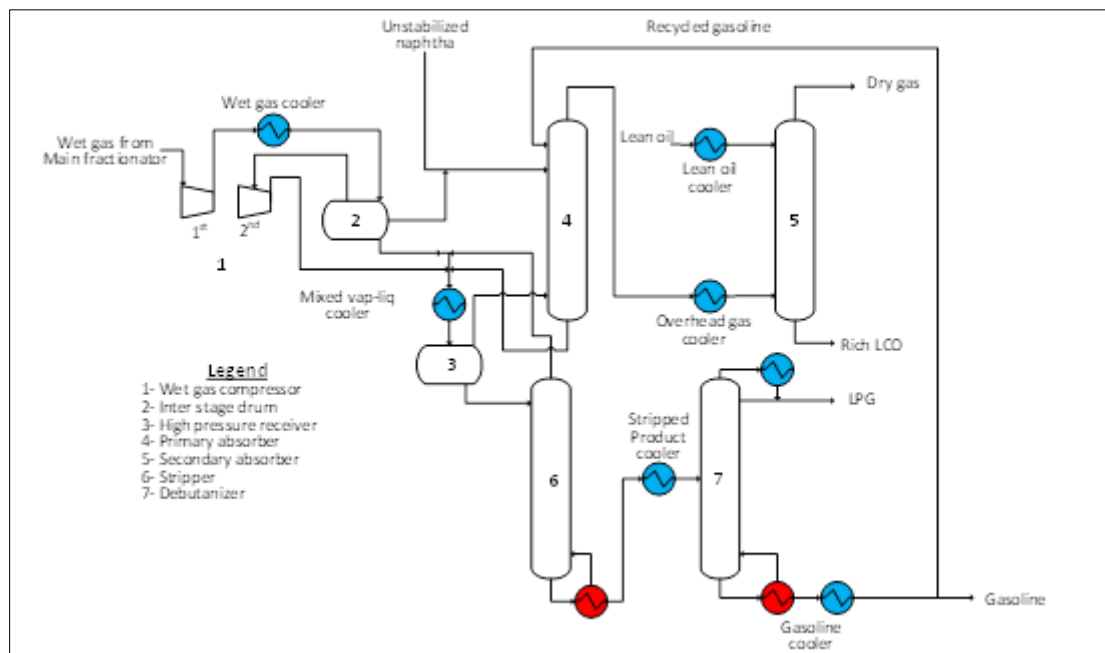
Figure 3 Process flow diagram of a base case gas concentration unit of a fluidised catalytic cracking unit plant.

The process flow diagram in **Figure 3** shows a base case gas concentration unit from a FCC plant, which is used for recovering valuable amount of C_3+ (gasoline and LPG components) hydrocarbons from light-ends exiting the FCC main fractionators. The light-end or wet gas from the FCC main fractionator is compressed, cooled and separated in the inter-stage drum. The liquid phase from inter-stage drum contains a large fraction of C_5+ hydrocarbon which is mixed with unstabilized naphtha from main fractionator and fed to primary absorber as solvent (lean oil); while the vapour phase is further compressed and mixed with stripper overhead product and primary absorber bottom product. The mixed stream is cooled and fed to high pressure receiver. The vapour phase from high pressure receiver contains a large fraction of C_5- compounds and is return to primary absorber as gas feed while liquid phase containing predominantly C_3+ hydrocarbons with small amount of C_2- and non-hydrocarbon gases are fed to the stripper. Overhead gas from primary absorber is feed to secondary absorber to recover C_5+ hydrocarbons using light cycle oil from main fractionator; rich light cycle oil is returned back to the main fractionator while non-hydrocarbons plus unrecovered hydrocarbon leaves the column as dry gas. Stripper bottom product stream is cooled and separated into LPG and gasoline in debutanizer; gasoline product is cooled and portion of the stream is recycled back to primary absorber, which serve as solvent (leaner in C_3s and C_4s) to enhance C_3s and C_4s hydrocarbon recovery.

2.2. Process modelling and simulation

In order to meet the research goals, basic data from an existing FCC plant, obtained from the "Samples" files in Aspen HYSYS v7.3 is used to build the model of the gas concentration unit in Aspen HYSYS. The model is then used to examine the effect of degrees of freedom on gasoline and LPG recovery, considering heat recovery. The performance indicators used for the analysis include gasoline and LPG recovery and fraction of gasoline and LPG in dry gas. Amount of gasoline and LPG recovered determines the revenue from the process, whereas fraction of gasoline and LPG in dry gas indicates lost hydrocarbon.

2.3. Process optimization and heat integration

2.3.1. Process optimization

As discussed in **Section 1**, the key degrees of freedom affecting recovery of gasoline and LPG discussed in literature includes solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling. These degrees of freedom are used to set up five cases, which are investigated separately to gain insights into the impact of each degree of freedom or operating condition on product recovery and heat recovery, in addition to finding optimal operating conditions. For each case, the degree of freedom is varied within a range that is set by the allowable operating conditions of the process equipment and properties of the absorption solvent. The range constitutes four different values of the degree of freedom, which are termed Test 1, Test 2, Test 3 and Test 4. Each test value is then applied on the gas concentration unit model in Aspen HYSYS, by adjusting the operating conditions. The resulting effect on recovery of gasoline and LPG, and fraction of gasoline and LPG in dry gas are recorded. For each case, the test with the highest recovery of gasoline and LPG is selected for heat recovery analysis and economic evaluation, because high product recovery maximizes revenue, heat recovery minimizes utility cost, and therefore benefit is maximized. For the sake of brevity, only the range for each degree of freedom is presented.

The **temperature** range used in this investigation includes 15℃, 25℃, 45℃ and 55℃ (Test 1, Test 2, Test 3 and Test 4 respectively); the base case solvent temperature is 35.4℃. For **operating pressure**, the range used includes 900 kPa, 1100 kPa, 1500 kPa and 1700 kPa, (Test 1, Test 2, Test 3 and Test 4 respectively); the base case operating pressure is 1300 kPa. **Solvent flowrate:** (1) effect of unstabilized naphtha on gasoline and LPG recovery is investigated by varying the stream flow rate in the range of 400 kmol/h, 450 kmol/h, 600 kmol/h and 650 kmol/h (Test 1, Test 2, Test 3 and Test 4 respectively); the base case flow rate is 552.4 kmol/h, (ii) debutanized gasoline is examined by varying the stream flow rate in the range of 19256 kmol/h, 28884 kmol/h, 38512 kmol/h and 57768 kmol/h which correspond to 40%, 60%, 80% and 120% of the Base case recycle flow rate (48140 kmol/h) (Test 1, Test 2, Test 3 and Test 4 respectively). **Solvent composition:** (i) components selected for the current study include n-pentane, 1-pentene, cyclopentane and benzene which belong to paraffinic, olefinic, naphthenic and aromatic hydrocarbons respectively. Therefore, increasing the flow rate of n-pentane, 1-pentane, cyclopentane and benzene represents as Test 1, Test 2, Test 3 and Test 4 respectively. For each test, the flow rate of the specific hydrocarbon (such as n-pentane, 1-pentene etc.) in the unstabilized naphtha stream is increased by 50 kmol/h (from the base case flow rate 10.1 kmol/h, 4.8 kmol/h, 0.9 kmol/h and 0 respectively), (ii) the lightest component to be absorbed in the current study is propane (C_3); hence, the effect of carbon number is investigated by increasing the flow rate of hydrocarbons which include pentane (C_5), hexane (C_6), heptanes (C_7) and octane (C_8), (Test 1, Test 2, Test 3 and Test 4 respectively). **Interstage cooling:** heat removal is carried out using two pumparounds with flow rates of $64 \ m^3/h$ (Haik, 2005) and the draw stage temperatures of 46℃ (Tray 5) and 50℃ (Tray 8). The return stages are Tray 4 and Tray 7 respectively, and the return temperatures are set

accordingly. The return temperature range includes 22℃, 26℃, 30℃ and 34℃ (Test 1, Test 2, Test 3 and Test 4 respectively).

2.3.2. Process heat integration

In this work, the extracted stream data for the selected test in each case is applied in SPRINT. The outputs include composite curve, grand composite curve, heat recovery and utility demand. The grand composite curve is used to explore steam generation options in the process. Hence, trade-offs between product recovery and heat recovery or utility demand is established. The trade-offs provide insights into which degree of freedom can be adjusted to derive benefit from the process in term of improved product recovery and heat recovery. Similarly, the gasoline and LPG recoveries from the selected test in each case are used to determine the revenue from the process, and the utility demand is used to determine the operation cost. The recent market price of each product and utility is used for estimating the revenue and operating cost. Two scenarios are evaluated; (i) steam generation is not considered, (ii) steam generation is considered. In both scenarios, the net benefit is estimated by subtracting operating cost from revenue. The benefit from the cases is compared, in order to provide insight into which degree of freedoms can be adjusted to maximized benefit.

Heat recovery analysis involves the following;

- Extract process stream data from the Aspen HYSYS simulation
- Use the stream data to develop composite and grand composite curve
- Determine heat recovery and utility requirements from the composite curve and
- Explore options for steam generation using the grand composite curve.

Hence, for each case in **Section 2.3.1**, the most beneficial tests in terms of high recovery of gasoline and LPG are selected for heat recovery analysis. The process stream data used in the analysis is extracted from the corresponding Aspen HYSYS simulation according to established methods (Linnhoff, 1998; Smith, 2005; Kemp, 2007). The extracted stream data is then applied in SPRINT software to determine the heat recovery and utility requirements using composite. The opportunity for generating MP steam for process heating is also explored using grand composite curve. The data in **Table 1** is used for the analysis.

Table 1 Data for heat recovery analysis

Parameter	Value
ΔT_{min} (℃)	10
Cooling water temperature (℃)	25
Cost of cooling water (£/ kWh)	0.000025
Cost of MP steam (£/ kWh)	0.011290

Therefore, for each case, the performance indicator obtained in **Section 3.3** together with heat recovery and utility requirements obtained in **Section 3.4** is used to consider trade-offs between product recovery and heat recovery, and between product recovery and utility requirements. These trade-offs provide general insights into the degrees of freedom that can adjusted to significantly improved product recovery and heat recovery.

2.4. Process economic evaluation

The economic evaluation involves determining benefit using selling price of gasoline and LPG together with cost of cooling water and compressor power demand. Benefit is determined and used for comparing the cases, in order to provide general insights into which degrees of freedom can be adjusted to maximize benefit. The procedure used to perform the economic evaluation is explained below;

- Step 1: Determine the revenue from product sales

The products include gasoline and LPG; the flow rate of each product is obtained from the case study results in **Section 3.3**. The current market price of the product/ utility (Whatgas, 2012; Quarterly energy price, 2013; Pitt, 2013) in **Table 2** is used to evaluate the net revenue.

Table 2 Market price of product and utility

Product/ utility	Value
Gasoline (pence/litre)	136.27
LPG (pence/litre)	70.12
Steam at 15 bar (£/kg)	0.008188
Electricity (pence/kWh)	8.51
Cooling water (pence/t)	2.5

*Calculated from fuel cost as shown in appendix

- Step 2: Determine the operating or utility cost required to produce the products

The utility required to produce the products after heat recovery include cooling water, compressor power demand and steam (at 1000 kPa and 1351 kPa); the data in **Table 3** present the details of the steam requirement by the FCC plant.

Table 3 FCC and Main Column Steam Requirement

Name	Mass flow rate (kg/h)	Temperature (°C)	Pressure (kPa)
FCC steam	4 597	200	1000
MC steam	5 500	240	1351
Diesel steam	370	240	1351

The data in **Table 3** is obtained from the FCC plant accompanying the gas concentration unit in Aspen HYSYS "Samples files". Data in **Table 3** are unaffected by change in degrees of freedom in the gas concentration unit. The cooling water and compressor power requirements for the case are obtained from **Section 3.4** and **Section 3.3** respectively. Note that the power requirements for all cases are constant, because the compressor pressure for all cases is fixed at 1802 kPa. Therefore, the utility costs in **Table 2** are used to evaluate the operating cost for each case.

- **Step 3: Determine the net production benefit**

The net production benefit is determined by subtracting the operating cost (determined in Step 2) from revenue (determined in Step 1). The benefit is evaluated for two different scenarios, namely heat recovery with steam generation and heat recovery without steam generation. For the former scenario, the credit obtained from steam generated from the process is added to the net revenue, while in the latter scenario, cooling is satisfied by cooling water and no steam heating is required.

3. Results and discussion

3.1. Effect of degrees of freedom on gasoline and LPG recovery

A simulation model of the gas concentration unit is developed and presented in **Figure 4**. Flowrates of key process streams are summarised as follows: gasoline and LPG recovered corresponds to 691.9 kmol/h (67470 kg/h) and 1099 kmol/h (55830 kg/h), dry gas flow rate is 703.4 kmol/h (14440 kg/h), which contains 2.97 kmol/h (166 kg/h) of C_3+ hydrocarbons (gasoline and LPG components). From the simulation results, the amount of C_3+ in dry gas can be reduced by recovering more gasoline and LPG, thereby increasing process profitability.
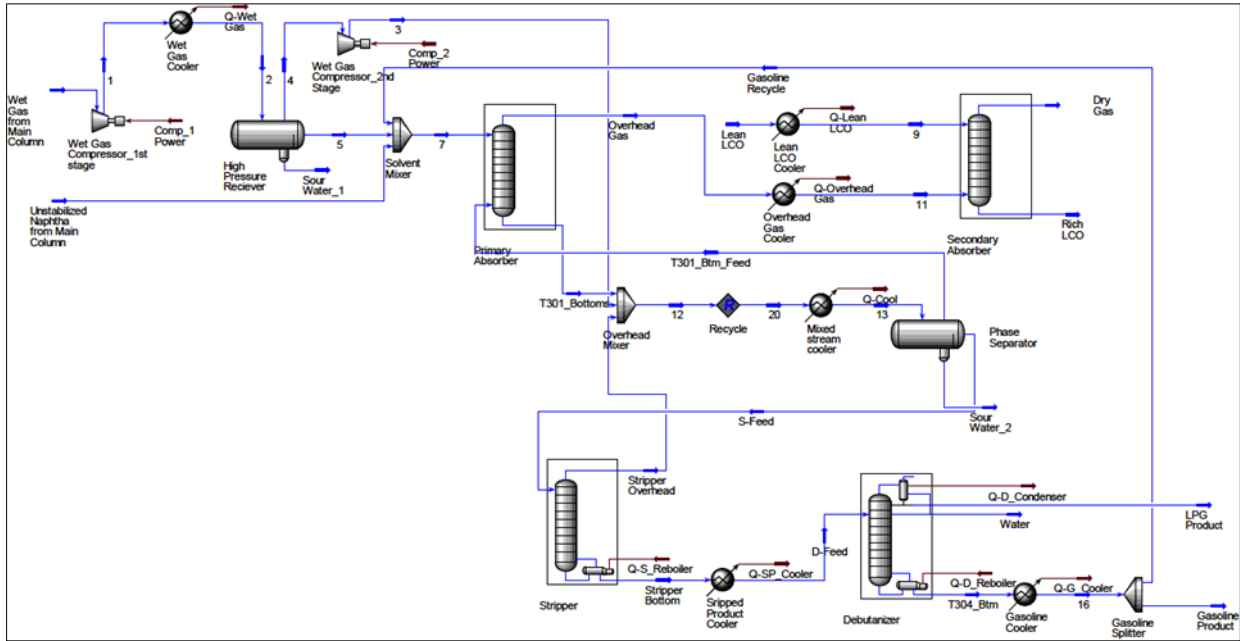
Figure 4 Simulation model of the base case gas concentration unit of a fluidised catalytic cracking unit

Table 4 shows the performance of selected modifications in each case study and/or optimal degree of freedom leading to improved gasoline and LPG recoveries. Aside product recovery, this work applies heat integration analysis and economic evaluation (in addition to product recovery) in order to improve the performance of a gas concentration unit of a FCC plant. Hence, **Sections 3.2 and 3.3** present effect of degrees of freedom on heat recovery and process economics respectively.

Table 4 Result for best performance for case studied

Case	Degree of freedom	Gasoline flow rate		LPG flow rate		Dry gas flow rate
		(kg/h)	(kmol/h)	(kg/h)	(kmol/h)	C_3+** (kmol/h)
Base	Base	67470	691.9	55830	1099	2.9649
1	Temperature	67470	691.9	55830	1099	2.9649
2	Pressure	68220	698.9	56050	1103	2.2774
3	Flow rate	67400	692.4	56010	1102	2.4555
4	Composition	72300	743.4	55780	1099	2.8402
5	Inter stage cooling	67470	691.9	55830	1099	2.9649

**Gasoline and LPG components in dry gas

3.2. Effect of degrees of freedom on heat recovery

This section presents and analyses the case study results of the investigation set up in **Section 2.3**. The results include extracted stream data, composite curve, grand composite curve, utility requirement, heat recovery and steam generation. It should be noted that the presented results are only for the selected modifications in each case study in addition to the base case.

3.2.1. Base case

The extracted stream data from the FCC plant and gas concentration unit is presented in **Table 5**; shaded rows indicate streams extracted from the FCC main fractionator, while the unshaded rows indicate streams extracted from the gas concentration unit. Streams #1 to #12 are heat sources and Stream #13 and #14 are heat sinks (debutanizer and stripper reboiler). Note, data on FCC feed preheat is not provided in the base case FCC plant, only data on FCC steam demand is provided, therefore steam generated in the process will be used in the FCC reactor.

Table 5 Extracted stream data for Base Case

Stream No.	Stream Name	Supply Temp. (℃)	Target Temp. (℃)	Heat load (ΔH) (kW)	CP (kW/K)
1	MC Condenser	120.4	41	37722	475
2	PA H-Naphtha	145.4	82	8894	140
3	PA LCO	250	187	7881	125
4	PA HCO	371	272	5867	59
5	PA Quench	345	257	11136	127
6	PA Subcool	345	257	916666	10417
7	Wet Gas Cooler	92.35	35	6164	108
8	LCO Cooler	206.9	35	7942	46
9	Phase Sep_Cooler	50.35	40	48950	4729
10	Stripper P_Cooler	177.3	127	180900	3596
11	Gasoline Cooler	177.8	35	456600	3197
12	Debutanizer Cond	65.9	51.1	35060	2369
13	Debutanizer Reb	163.4	177.8	21200	1472
14	Stripper Reb	149.6	177.3	503900	18191

However, there is no heat integration scheme in the existing FCC plant; all cooling and heating needs are satisfied by external utility such as steam and cooling water. Utility cost dominates the cost of running an FCC plant (Dean et al., 2000). Minimizing external utility demand reduces the cost of utility and maximizes profit.

For the current study, external utility requirement is minimized by considering heat recovery opportunities in the FCC plant using pinch methodology. The composite curve and grand composite curve obtained from SPRINT using the data in **Table 5** are presented in **Figures 5 and 6** respectively.
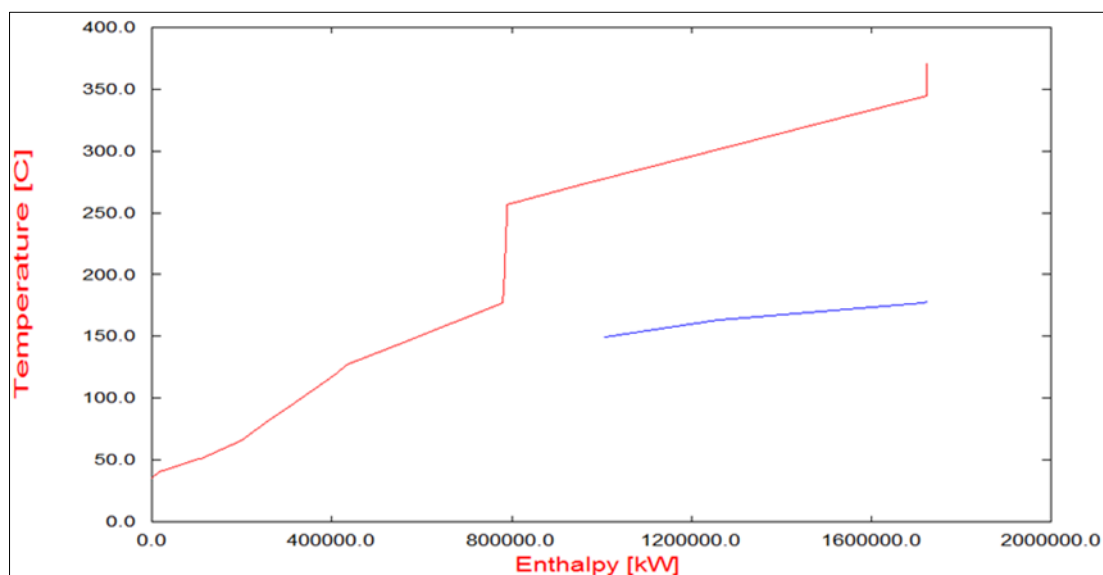
Figure 5 Composite curves for Base case FCC plant. Red line denotes hot composite curve whilst blue line denotes cold composite curve

Figure 5 shows that the process requires cooling of 1008 MW at 35°C, which can be supplied by cooling water at ambient temperature of 25°C (for minimum approach temperature of 10°C). Furthermore, the amount of heat recovery in the process is 716 MW. It should be noted that FCC plant requires only cooling after heat recovery, i.e. the process exhibits 'threshold problem' behaviour

The grand composite curve is used to explore opportunity for steam generation in the FCC plant. As shown in **Figure 6**, medium pressure (MP) steam is generated at a rate of 379 MW from the process, which leads to reduction of cooling water to 629 MW from 1008 MW. Therefore, process steam generation reduces the cooling water requirement by 62%, which will in turn reduce utility cost. Information from the heat integration analysis such as cooling water demand, process steam generated and heat recovery are used in **Section 2.3** to determine utility cost, which is used to evaluate benefits. A similar heat integration analysis is carried out for the four case studies discussed in **Section 2.3**; in each case, the aforementioned information is summarized and used in benefit evaluation.
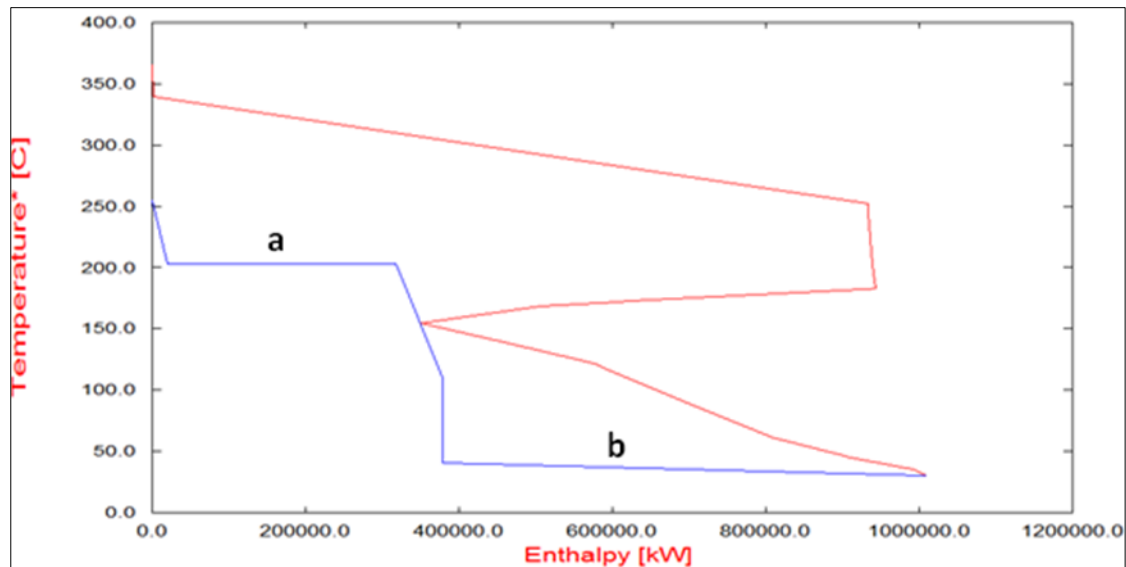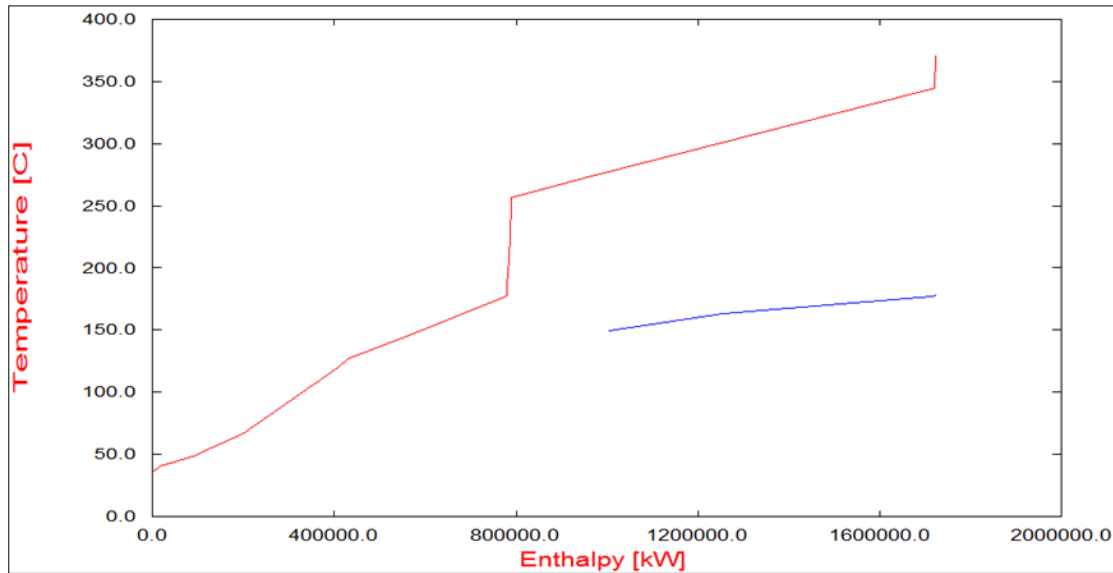

Figure 6 Grand composite curves for base case FCC plant. a, denotes steam generation; b, denotes cooling duty

### 3.2.2. Case 1 – Solvent temperature to the primary absorber

The heat integration analysis for Case 1 is the same as the base case because of cooling at ambient temperature. Therefore, no heat recovery analysis data is presented for Case 1.

3.2.3. Case 2 – Operating pressure in the primary absorber

The composite curve for Case 2 shown in **Figure 7** indicates that 719 MW of heat is recovered from the process, which reduces the cooling water requirement to 629 MW from 1004 MW. Hence, cooling demand is reduced by 37% as a result of heat integration. Therefore, comparing the current case to the base case, heat recovery is reduced by 1% while cooling water demand is increased by 0.1%; which results from increase product flow rate due to improvement of gasoline and LPG recovery.

Figure 7 Composite curves for Case 2- Operating pressure of primary absorber. Red line denotes hot composite curve whilst blue line denotes cold composite curve
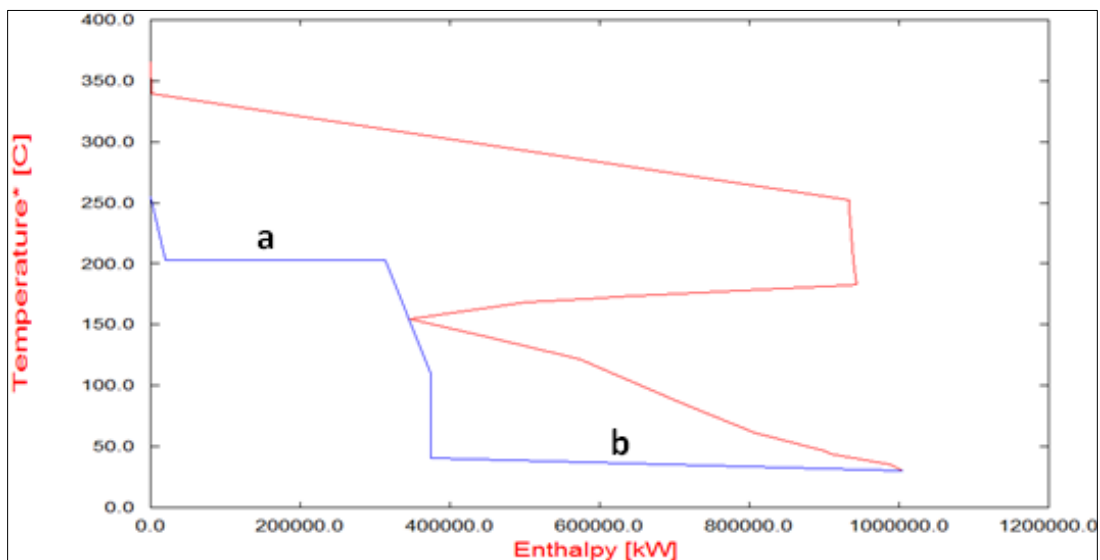

Figure 8 Grand composite curves for Case 2- Operating pressure of primary absorber. a, denotes steam generation; b, denotes cooling duty

The grand composite curve for Case 2 shows that 375 MW of MP steam is generated from the process; cooling water demand is reduced to 629 MW from 1004 MW, which represents reduction of 63%. The heat recovery analysis for operating pressure shows a slight increase of process heat sinks as shown in **Figure 8**. The increase of heat sink is caused by increased amount of hydrocarbons that are separated in the stirpper and debutanizer reboilers, which is accompanied by increased heat duty.
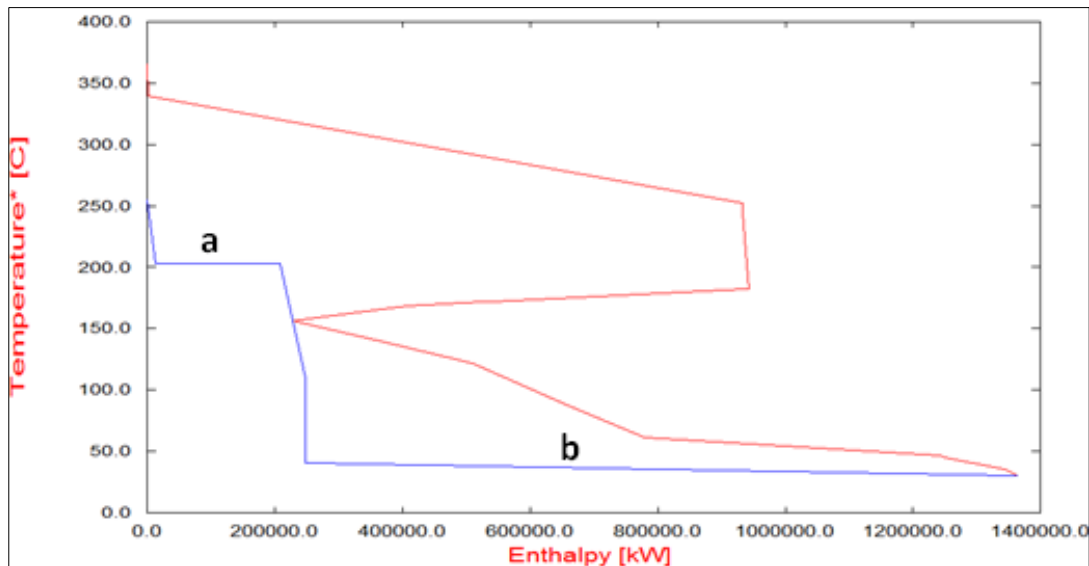
3.2.4. Case 3 – Solvent (unstabilized naphtha) flow rate

The results for Case 3 shown in **Figure 9 and 10** indicate that 843 MW of heat has been recovered in the process and the cooling water requirement is 1364 MW. There is a noticeable decrease in heat recovery of 35% and increase in cooling water demand of 18% as compared to the base case. The heat recovery difference is caused by increase of process heat sinks resulting from increase heat duty in the stripper and the debutanizer reboilers required for separation of C_2- and C_3+ in stripper, and C_4- and C_5+ in debutanizer. Furthermore, the increase in debutanized gasoline flow to 57768 kmol/h from 48140 kmol/h caused large cooling duty in exchangers, thereby increasing cooling water requirement.
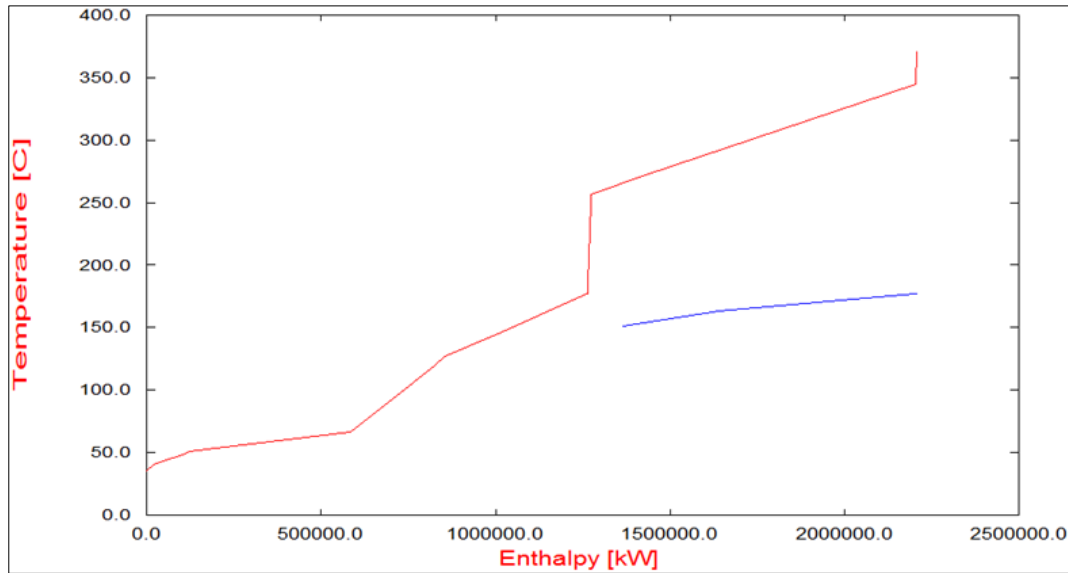
Figure 9 Composite curve for Case 3- Solvent (unstabilized naphtha) flow rate. Red line denotes hot composite curve whilst blue line denotes cold composite curve
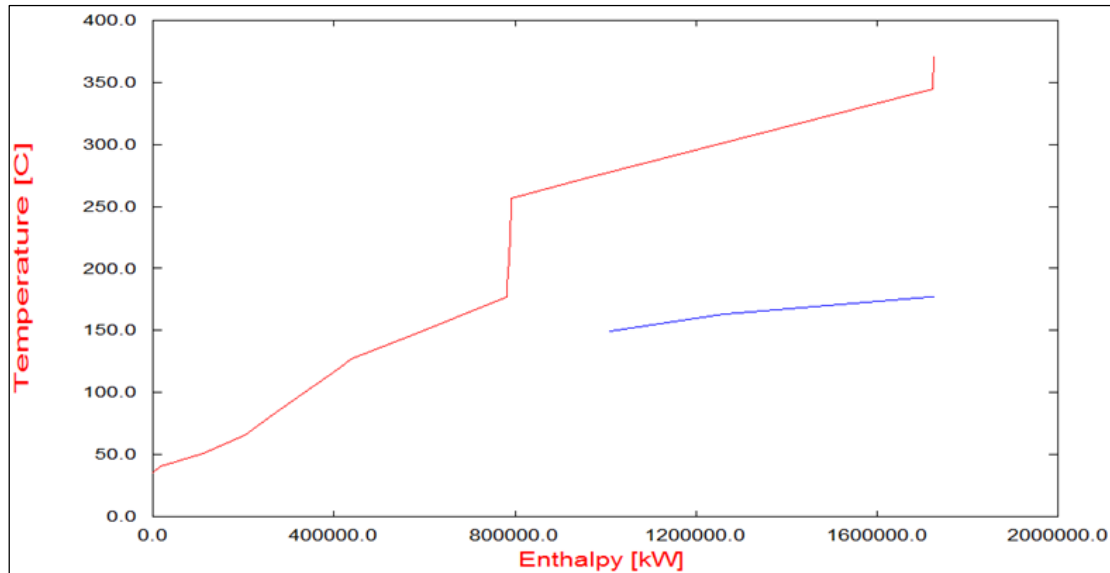
The quantity of steam generated from the process is 249 MW, which reduces cooling water requirement by 18% i.e. from 1364 MW to 1115 MW. The comparison of Case 3 to the base case indicates 35% reduction of heat recovery and 77% increase of cooling water, which results from increase duty of heat sinks and heat source respectively due to increase product flow rate.

Figure 10 Grand composite curve for Case 3- Solvent (unstabilized naphtha) flow rate. a, denotes steam generation; b, denotes cooling duty

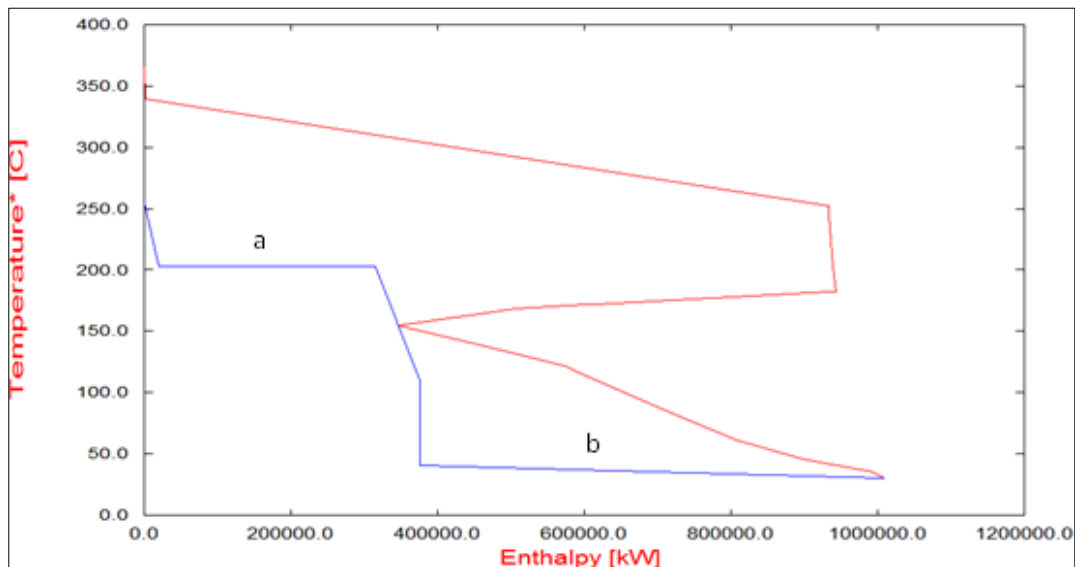*3.2.5. Case 4 – Solvent composition (increased C_6 content)*

The composite curve for Case 4 shown in **Figure 11** indicates that the heat recovery and cooling water requirement are 717 MW and 1009 MW respectively. Comparing the case result to the Base case, it is observed that both heat recovery and cooling water requirement has increased by 0.1%, which are approximately equal.

Figure 11 Composite curve for Case 4– Solvent composition (increased C_6 content). Red line denotes hot composite curve whilst blue line denotes cold composite curve

The amount of steam generated from the process is 375 MW as shown in **Figure 12**; hence the cooling water requirement is reduced to 633 MW from 1009 MW, which represents 37% reduction. Therefore, comparing the Case 4 result to the Base case, the MP steam generation is reduced by 1% and cooling water demand is increased by 1%.

Figure 12 Grand composite curve for Case 4– Solvent composition (increased C_6 content). a, denotes steam generation; b, denotes cooling duty

3.2.6. Comparison of utility demand and heat recovery for case studies

Figure 13 presents a summary of results from the heat recovery analysis in this section. It should be noted that Case 1 and Case 5 results are the same as the Base case.
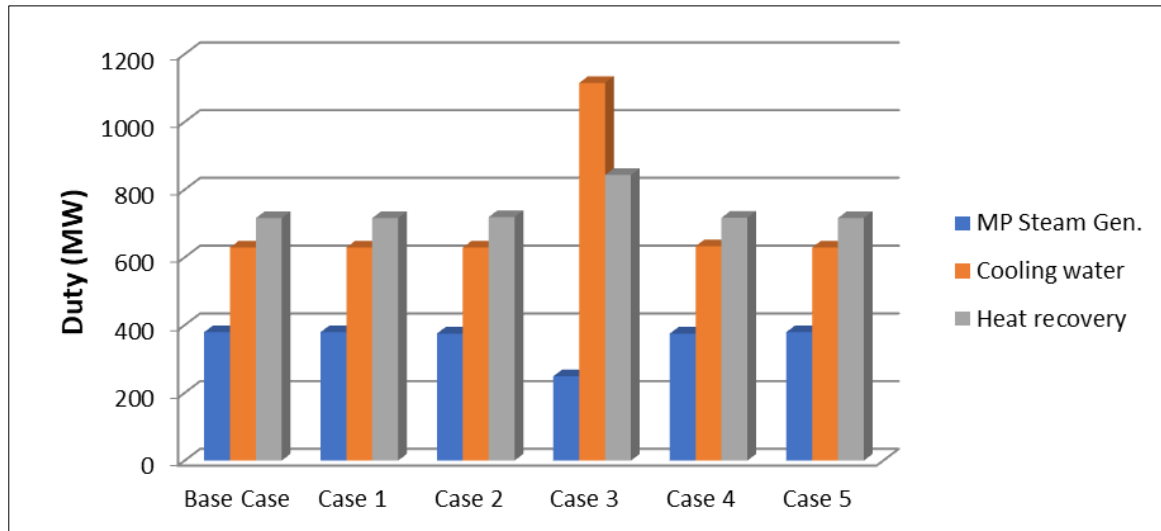
Figure 13 Comparison of heat recovery analysis results

Based on the result shown in **Figure 13**, the following deductions are drawn.

- Increase in operating pressure by 400 kPa in the primary absorber increases heat recovery and cooling water by 0.5% and 0.1%, while steam generation is decreased by 1%.
- Increasing flow rate of recycled gasoline by 9628 kmol/h (20% of base case flow), increases heat recovery and cooling water demand by 18% and 77%, while steam generation is decreased by 35%.
- Increasing the flow rate of C6 hydrocarbons by 50 kmol/h in the unstabilized naphtha stream, increases heat recovery and cooling water demand by 0.1% and 1%, while steam generation is decreased by 1%.

Therefore, change in the flow rate of recycled gasoline have the most significant impact on heat recovery, steam generation and cooling water requirements in the FCC plant, which is largely caused by increase in process streams flow rate due to increase recycled gasoline flow. Change in operating pressure has the least impact, followed by change in solvent composition as shown in **Table 4**.

3.3. Economic Analysis

The results obtained by applying the economic evaluation procedure in **Section 3.5** is presented, analysed and discussed in this section. Thus, the analysis is carried out in two scenarios namely:

- steam generation is not included in the analysis;
- steam generation is included. For the former, the analysis examines the extent to which heat recovery can reduce utility cost, while the latter scenario examines how steam generated and used in the process can further reduce utility cost. It should be noted that the current analysis links the effect of change in degrees of freedom on product recovery and heat recovery and or utility requirements.

3.3.1. Scenario 1 – Heat recovery without steam generation

The effect of change in degrees of freedom on gasoline and LPG recovery is analysed in **Section 3.1**; similar analysis is also performed on heat recovery in this section. Therefore, to identify the potential benefits of varying each degree of freedom, the utility cost is used together with net revenue from product sales to evaluate profit as shown in **Figure 14**.
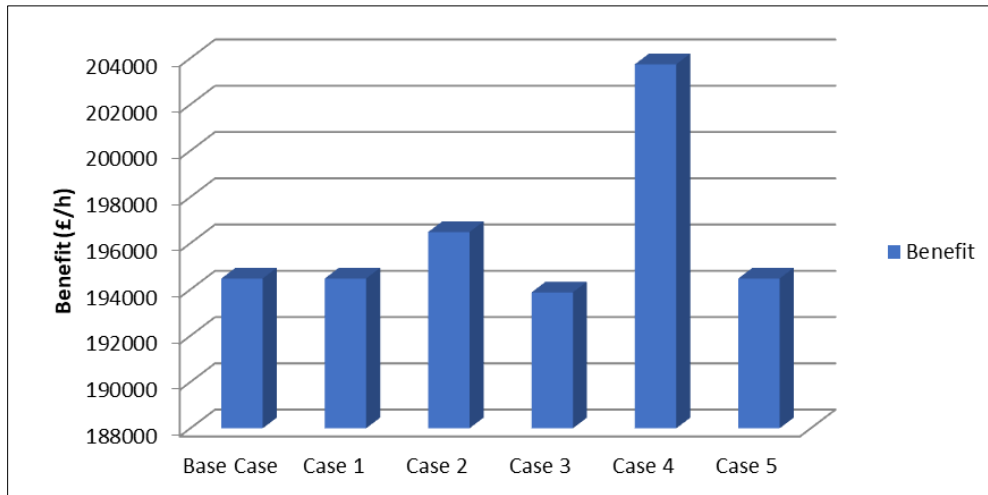
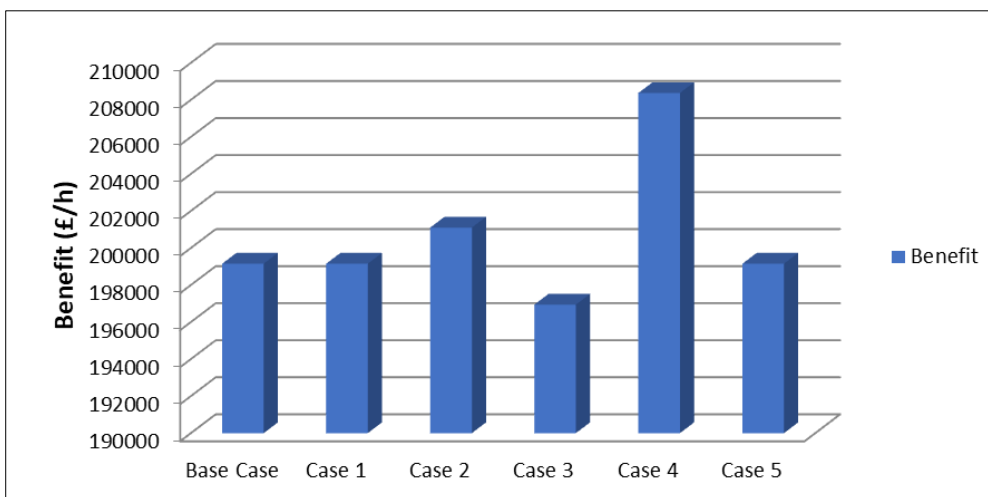**Figure 14** Net production benefits without steam generation

The following deductions are drawn based on **Figure 14**;

- Increase in operating pressure by 400 kPa, increases benefit by 1%
- Increasing recycled gasoline flow by 9628 kmol/h decreases benefit by 0.3%, which results from increased utility cost due increased exchanger duties. Furthermore, increase recycle gasoline flow rate impact more on LPG recovery than gasoline, and LPG has less market value based than gasoline.
- Increasing the composition of C6 hydrocarbon in solvent by 50 kmol/h increases benefit by 5%.
- For ambient cooling medium, no increase in benefit is observed. Similarly, for inter stage cooling using pumparound return temperature of 22°C-34°C, no increase in profit is observed in the current study.

Thus, increasing the composition of C_6 hydrocarbons in solvent (unstabilized naphtha) has the highest impact on benefit, which is largely contributed by the increase of gasoline and LPG recovery.

3.3.2. Scenario 2 – Heat recovery with steam generation

This scenario examines the effect of change in degrees of freedom on profit, taking into account product recovery, heat recovery, steam generation and cooling requirement. The steam generated in the process is used to offset steam requirements by the FCC reactor and main fractionator, and surplus steam generated from the process is used as credit. The total utility cost and revenue are used to evaluate profit for each case study, see **Figure 15**.

Figure 15 Net production benefits with steam generation

Comparison of benefit for case studies in **Figure 15** shows that Case 2 and Case 4 results to benefit increase of 1% and 5%, while Case 3 decreases profit by 1%. Based on the above comparison, it can be concluded that increase in C6 content of solvent (unstabilized naphtha) has the greatest impact on benefit.

In summary, no improvement of gasoline and LPG recovery is observed with solvent temperature at or above ambient condition, but significant improvement is observed at low temperature, which requires expensive cooling medium. An increase in primary absorber operating pressure improves gasoline and LPG recovery, but the pressure increase is limited by the capacity of the wet gas compressor. High operating pressure improves process heat recovery, but limits the opportunity for steam generation. Gasoline and LPG recovery may be enhanced by increasing unstabilized naphtha flow rate or recycled gasoline flow rates. Unstabilized naphtha improves gasoline recovery substantially, while recycled gasoline improves LPG recovery compared to gasoline. Increased flow rate improves heat recovery, but requires high utility demand. The carbon number of hydrocarbon in solvent was explored to enhance gasoline and LPG recovery; but a trade-off exists between carbon number and product recovery. Solvents dominated by high carbon number hydrocarbons (e.g. C_7, C_8) recovers more gasoline than LPG, while solvent dominated by low carbon number hydrocarbons (e.g. C_5, C_6) recovers more LPG than gasoline. Therefore, three carbon atoms heavier than C_3 is the best (i.e. C_6) for gasoline and LPG recovery. Increase of C_6 in solvent (unstabilized naphtha) has the highest impact on overall benefit compared with other degrees of freedom.

4. Conclusion

This research work investigates degrees of freedom/ operating conditions that can impact gasoline and LPG recovery as well as heat recovery in a gas concentration unit of a FCC plant. The methodology proposed in the work improves simultaneously gasoline and LPG recoveries and heat recovery. The effect of degrees of freedom (solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling) on product recovery and heat recovery has been investigated by adjusting each degree of freedom independently and the corresponding improvement in product recovery and heat recovery are monitored, thus providing insight into which degrees of freedom have high influence on product recovery and heat recovery. This study reports 1% and 0.4% increase in gasoline and LPG recovery by increase of operating pressure from 1300 kPa to 1700 kPa, which is assumed to be within the wet gas compressor capacity of 1802 kPa. The modification increases heat recovery and cooling water demand by 0.5% and 0.1%, whereas steam generation is reduced by 1%. Total benefit is increased by 1%. Recovery of gasoline and LPG are improved by 0.1% and 0.3% due to increase in recycled gasoline flow rate by 9628 kmol/h (20% of initial flow rate). The current modification increases heat recovery and cooling water demand by 18% and 77%, but steam generation decreases by 35%; which results to decrease in benefit of 1%. Thus, it can be concluded that modifying recycled gasoline flow rate to maximize benefit is feasible only when the market value of LPG is considerably high. Hydrocarbon solvent containing high C_6 hydrocarbon is identified as the best, for recovery of both gasoline and LPG. Thus, the current study reports 7.5% increase in gasoline recovery by increasing C_6 content of solvent (unstabilized naphtha) by 50 kmol/h. Heat recovery and cooling water demand increases by 0.1% and 1%, and benefit increases by 5%. In conclusion, increase of C_6 (60-70* pseudo component) composition in unstabilized naphtha entering the primary absorber has the highest impact on benefit compared with other degrees of freedom. In the current study, the degrees of freedom affecting gasoline and LPG recoveries are investigated independently. This approach limits the improvement of gasoline and LPG recoveries to the best value of only one degree of freedom. Future work should consider influence of two or three degrees of freedom simultaneously which could provide an improved performance. The current research is limited to exploring the influence of degrees of freedom on gasoline and LPG recovery in the gas concentration unit, assumed within capacity limit of equipment. It is recommended that research be carried out that will explore the degrees of freedom more widely, to search for more potential benefits. The proposed work would consider process revamps, and require more detailed capital cost analysis, in addition to heat recovery analysis and investigation of effects of degrees of freedom on recovery of gasoline and LPG.

Compliance with ethical standards

Disclosure of conflict of interest

No conflict of interest to be disclosed.

References

[1] Al-Riyami B. A., Klemes J., & Perry S. (2001). Heat integration retrofit analysis of a heat exchanger network of a fluid catalytic cracking plant. Applied Thermal Engineering, 1449-1487.

[2]   Birol F., (2011). International energy outlook. Paris: International energy agency.

[3]   Branan C. (2002). Rule of Thumb for Chemical Engineers. Houston, Texas: Gulf Publishing Company, pp 99-100.

[4]   Bumbac G., Ciornei C., Turcu A., & Muntean O. (2008). New Solution for Thermal Coupling between a Crude Distillation Unit and a Catalytic Cracking Unit. Revue Roumaine de Chimie, 635-641.

[5]   Chang A. F., Pashikanti K., & Liu Y. A. (2012). Refinery Engineering: Integrated Process Modelling and Optimization. Weinheim, Germany: Wiley-VCH Verlag & Co. KGaA.

[6]   Dean C. F., Golden S., & Rich P. (2005). Increasing FCC propylene yield. Saudi Aramco Journal of Technology, 44-51.

[7]   Gary J. H., Handwerk G. E., & Kaiser M. J. (2007). Petroleum Refining: Technology and Economics. Baco Raton: Taylor and Francis Group.

[8]   Golden S. W., & Fulton S. (2000). Low-cost methods to improve FCCU energy efficiency. PTQ, 1-9.

[9]   Haik S. M. (2005). Improved propane recovery increases refinery profitability. Oil and Gas Journal, 47-51.

[10]   Kemp I. C. (2007). Pinch Analysis and Process Integration, Second Edition: A User Guide on Process Integration for the Efficient Use of Energy. Oxford: Elsevier.

[11]   Linnhoff B. (1998). Introduction to Pinch Technology. Cheshire: Linnhoff March, pp 5-11.

[12]   Lu E., Luo M., & Zhang H. (2004). The Integrated Simulation and Optimization for Absorption-stabilization System and Gas Process Unit. Unpublised.

[13]   Lu E., Zhang H., & Zhu X. (2000). A novel design for vapor revovery units. Computers and Chemical Engineering, 1317-1322.

[14]   Lucas A. G. (2000). Modern Petroleum Technology. West Sussex: John Wiley & Sons Ltd, pp 46-47.

[15]   McKee C., (2013). Quarterly Energy Prices. York: Department of Energy and Climate Change.

[16]   Mehra R., 1987. U.S, Patent 4696688.

[17]   Pitt M. (2013, January 8). IChemE Education Subject Group. Retrieved August 1, 2013, from http://ed.icheme.org/costutil.html

[18]   Sadeghbeigi R. (2000). Fluid Catalytic Cracking Handbook. Houston: Butterworth-Heinemann.

[19]   Schultz A., Wegerer A., Towler P., 2011. U.S Patent 8007661 B1.

[20]   Shailendra A., Asit K., Yogesh G., Shubhangi J., Minaz M., Sukumar M., Gobind T., Manoj Y., 2012. U.S Patent EP2449059 A2.

[21]   Smith R. (2005). Chemical Process Design and Integration. Chischester: Wiley, pp 181-184.

[22]   Speight J. G. (1999). The Chemistry and Technology of Petroluem. New York: Marcel Dekker.

[23]   Tagamolila C., Haun E. , Lomas D., Zimmermann J., 1994. U.S Patent 5360533.

[24]   Treybal R. E. (1981). Mass-Transfer Operations. Singapore: McGraw-Hill, pp 275-282.

[25]   Vasconcelos C. J., Filho R. M., Spandri R., & Woif-Maciel, M. R. (2005). On-Line Optimization Applied to Large Scale Plants. European Symposium on Computer Aided Process Engineering, Elsevier.

[26]   Veawab A., Aroonwilas A., Chakma A., Tontiwachwuthikul P., 2013. Solvent Formulation for CO2 …… Separation from Flue Gas Streams. Unpublished, pp 1-9.

[27]   Whatgas. (2012). Retrieved August 2, 2013, from http://www.whatgas.com/petrol-prices/index.html

[28]   Wilson J. W., & Sloan H. D. (1984). Energy Conservation Revamps in Fluid Catalytic Cracking Systems. Industrial Energy Technology Conference Volume 1, (pp. 312-318). Houston, Texas.

[29]   Wolschlag L. M., Couch K. A., Zhu F., & Alves J. (2009). UOP FCC Design Advancements to Reduce Energy Consumption and CO2 Emissions. Unpublised, 19-22.

[30]   Yela S. (2009). Framework for Operability Assessment of Production Facilities:An Application to a Primary Unit of a Crude oil refinery. Unpublised.