

PROCESS IMPROVEMENT OF PETROLEUM REFINERY FCC LIGHT-ENDS SEPARATION***Dauda Ibrahim**

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Abstract

To satisfy global demand for valuable petroleum refined products such as gasoline and liquefied petroleum gas (LPG) by achieving high recovery in gas concentration unit of a FCC plant, an insightful understanding of which degrees of freedom has significant impact on product (gasoline and LPG) recovery is required. This research proposes a methodology for investigating degrees of freedom that significantly impacts product recovery in gas concentration unit, in addition to proposing operational changes that improves overall process performance. First, a model of gas concentration unit builds in Aspen HYSYS, followed by setting up case study based on investigated degrees of freedom (solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling) from literature, and lastly evaluating each case on the developed model and analysing results. Key findings from this research includes the following trade-offs: increasing operating pressure in primary absorber by 30% (400 kPa), increases gasoline and LPG recovery by 1% and 0.4%; recovery of gasoline and LPG are improved by 0.1% and 0.3% due to increase in recycled gasoline flow rate by 20% (9628 kmol/h); increasing C_6 content of solvent (unstabilized naphtha) by 50 kmol/h (from 15.1 kmol/h), increases gasoline recovery by 7.5%; for ambient cooling medium, varying solvent temperature and pumparound return temperature (inter-stage cooling) does not improve product recovery. Results show that high C_6 composition in solvent has the highest impact on product recovery compared with other degrees of freedom. Proposed methodology and research outcomes can be used by plant operators to optimise performance of FCC light-end separation unit, thereby improving product recovery.

Keywords: Product recovery, Process modelling, Sensitivity analysis, Fluidised catalytic cracking unit.

INTRODUCTION

The global demand for petroleum-based fuels for industrial, transportation and domestic use has been increasing since origination. It has been projected that the world demand of petroleum and other liquids fuel will increase from an old figure of 85.7 million barrels per day in 2008 to 97.6 million barrels per day in 2020 and 112.2 million barrels per day in 2035 (International Energy Outlook, 2011). Petroleum-based fuels include gasoline, kerosene, diesel, LPG, fuel oil etc., which are produced in the refinery via fractional distillation of crude oil, and other conversion and product upgrading processes. To accommodate the increasing demand in petroleum refined products, refiners begin to search for an economic way to use crude oil to produce more valuable products such as gasoline, diesel, liquefied petroleum gas (LPG). A major milestone in the petroleum refining industry was the invention of the first commercial Fluidised Catalytic Cracking Unit (FCCU) in 1942 (Sadeghbeigi, 2000). The FCCU consists of a Fluidised Catalytic Cracking (FCC) reactor and a FCC main fractionator; the FCC reactor uses a catalyst to convert less valuable residual products (such as atmospheric gas oil, vacuum gas oil and other heavy hydrocarbons) into more valuable cracked products, while the FCC main fractionator separates the cracked products into unstabilized gasoline, diesel fuel, light-ends and slurry oil (which is recycled back to the FCC reactor for reprocessing) as shown in Figure 1.

Light-ends from 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. However, a relatively high loss of C_3+ hydrocarbon, around 22%, of the overhead product from the primary absorber has been reported in the literature (Lu *et al.*, 2000), with some valuable products slipping into the fuel gas system, which is undesirable. Therefore, more benefit will be made by the process if recovery of gasoline and LPG is improved. Moreover, improved product recovery will also alleviate the increasing demand of petroleum refined products. The heart of the gas concentration unit is the primary absorber, where the valuable constituents of the hydrocarbon vapour (light-ends) are selectively absorbed. The amount of gasoline and LPG recovered in the process is primarily dependent on the absorption occurring in the primary absorber, therefore improving absorption in the primary absorber increases the overall process yield. Absorption can be improved by (i) increasing the flow rate of solvent i.e. increasing the L/V ratio, (ii) decreasing the solvent temperature and removal of heat of absorption, (iii) increasing column pressure (Smith, 2005). Moreover, another important variable is solvent composition, which determines the performance of the solvent. Solvent performance is defined in terms of solvent capacity, stability, gas solubility, volatility, viscosity and absorption selectivity (Veawab *et al.*, 2013).

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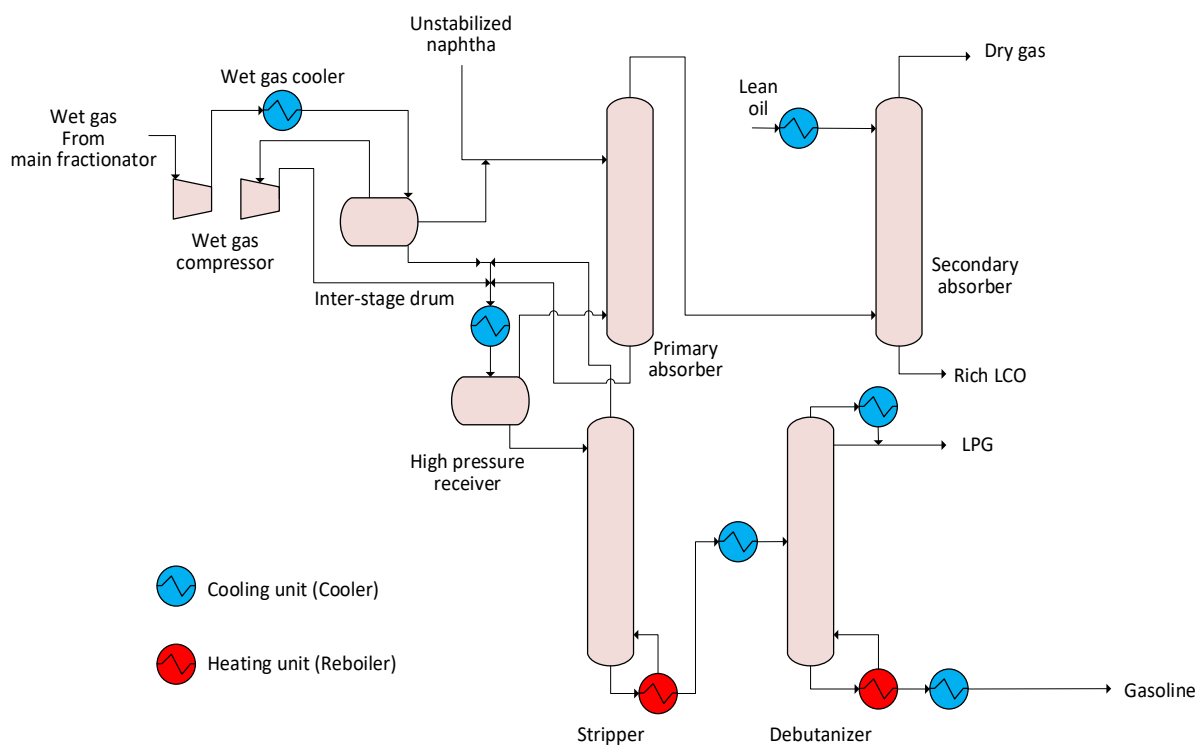


Figure 1. Gas concentration unit (Chang *et al.*, 2012)

Variables such as solvent temperature, operating pressure, solvent flow rate, solvent composition and heat removal (inter stage cooling) can be adjusted independently or simultaneously with other variables to improve absorption. Hence, they are referred to as degrees of freedom. The enhancement of gasoline and LPG recovery in the effluent of an FCC reactor has drawn attention with much research progressing in this area. Dean *et al.* (2005) revamped the primary absorber of the gas concentration unit of an FCC plant to assess the benefit of changing some operating conditions using a case study. They setup three cases: Case 1 increases propylene yield by 10% resulting from increase debutanizer gasoline flow (increase in L/V ratio); Case 2 considers removing heat of absorption from the column by installing a re-contacting drum which raises propylene yield by 15%; Case 3 reduces the temperature of the lean oil and inter cooler return temperature to 21°C using chilled water exchanger and propylene yield increases by 30%. Although the discussion of Dean *et al.* (2005) has taken into account some important degrees of freedom that directly impact on product recovery, in their work only the recovery of propylene was improved, improvement in gasoline recovery is not considered. Haik (2005) investigated increasing refinery profitability through improved recovery of valuable products in the gas concentration unit. He used a case study to examine how absorber pressure, lean oil temperature, inter cooler pump around heat removal, rectified absorber tower reboiler duty and lean oil flow rate can be used to improve propane recovery from the off-gas stream. Haik (2005) applied a combined optimization strategy using the aforementioned variables to achieve 26% improvement in propane (LPG) recovery. Although, the important variables which can be manipulated to improve product recovery were captured in his work, the effect of adjusting each variable, either independently or simultaneously with other variables, on utility consumption and process heat recovery is not considered in his work. Therefore, approach that incorporates improving both LPG and gasoline recovery together with utility consideration will provide a better evaluation of the refinery profitability.

Other authors considered more expensive option of installing additional equipment into the existing process in order to improve the recovery of gasoline and LPG in the gas concentration unit. Shailendra *et al.* (2012) invented a process for enhanced recovery of propylene and LPG from fuel gas produced in a fluid catalytic cracking unit. In their work, they installed an additional unit, called a naphtha stripper, to the existing process which strips off C₄s and lighter components from unstabilized naphtha exiting the main fractionator to obtain a liquid fraction almost free from propylene (< 0.1 mol %) and other LPG components. The authors enhanced the absorption capacity of lean oil to absorb higher amounts of propylene and LPG from fuel gas, leading to an improved recovery of propylene and LPG components, no improvement in gasoline recovery is reported. Schultz *et al.* (2011) modified the absorption recovery process of FCC-produced light olefins such that cracked hydrocarbon separates based on carbon number. The modified process consists of 24 units configured in a complex order, which results to an improved olefin recovery of 98 mol%. Their focus is only on improving C₃s and C₂s recoveries; therefore, there is a need to also consider higher molecular weight hydrocarbon recovery to increase throughput into the refinery gasoline blending pool. Tagamolila *et al.* (1994) invented a process of direct dry gas recovery from fluid catalytic cracking reactor. In their invention, quench and absorption vessels are added to the existing FCC light-end separation process. The FCC reactor has two outlets: the first outlet contains the vapour product from the reactor riser and is routed to the main fractionators, while the second outlet contains products from reactor vessel and is routed to the additional units. They authors used heavy hydrocarbon (light cycle oil) in the quench and absorber vessels to absorb hydrocarbons from the reactor product stream and return the rich LCO stream to the main fractionator, while the leaner stream, that is rich in gasoline is routed to the primary absorber. This approach increases product recovery capacity of the FCC light-end separation; however, it requires a significant modification of the FCC reactor and also involves

structural changes to the gas concentration unit to accommodate the quench and absorber vessels; which will result to a huge capital investment and operability concerns. None of the above authors has considered enhancing the recovery of gasoline fraction of the effluent from the fluid catalytic cracking reactor. The need to simultaneously recover LPG together with gasoline is of great importance in order to add more value to the process. Lu *et al.* (2000) describe a novel design for a gas concentration unit with the aim of minimizing both capital and operating cost. The design uses a complex column known as a combined oil absorption-deethanizer to replace the original absorber and stripper in the existing process; also, the heat exchanger network for the novel design was rearranged and optimized, and the two recycle streams in the existing gas concentration unit were eliminated. The structural changes result to an increase yield of gasoline and LPG with a net profit increase of 44% and substantial reduction in capital and operating cost of 22% and 19% respectively. Nonetheless, this approach is economical only for a new design.

Other authors used an optimization-based approach to improve C₃₊ hydrocarbons recovery in the gas concentration unit. Lu *et al.* (2004) presents an integrated simulation and optimization approach of the absorption-stabilization system in a gas concentration unit. The authors used net profit as the optimization objective and absorbent flow rate, stripper feed temperature, C₄ content of stabilizer bottom as decision variables. The C₂ concentration in LPG, C₃ concentration in off-gas, feed temperature, C₄ concentration in stabilizer bottom product and absorbent flow rate are constraints. The optimized case eliminates the deethanizer and improves propylene recovery by 5% which corresponds to a net profit rise of 1%. Again, gasoline recovery is not reported to have been substantially improved. Similarly, Vasconcelos *et al.* (2005) set up an online optimization of the gas concentration section of a fluid catalytic cracking plant using factorial design methodology, with the view of attaining optimum operating conditions. The authors developed an optimization model using net profit as objective function with compressor pressure, gasoline recycle flow rate, the main column bottom temperature, gasoline product Reid vapour pressure, top and bottom temperature of debutanizer as constraints. They formulated a concise conclusion that an increase in the bottom temperature of the main column increases the LPG productivity and net profit. Similarly, an increase in the second compressor outlet pressure decreases the productivity and net profit. An increase in the gasoline recycle flow rate decreases productivity and net profit. Their model was able to account for the complex interaction involved in the light-end separation to optimize profit; however, gasoline recovery is not enhanced. So far, however, there has been little attention paid to improvement of gasoline recovery in the gas concentration unit; most authors emphasize approaches which require huge capital investment. There is a need for alternative revamp solution that improves the recovery of both gasoline and LPG in the gas concentration unit, without making significant changes to the existing unit structure. The solution leads to significant savings in capital investment while maximizing benefit and meeting global demand for petroleum refined products. This research critically examines which degrees of freedom have significant impact on yield of gasoline and LPG in the GCU and proposes a methodology for improving yields of gasoline and LPG in the GCU. To improve product recovery, this research model and simulate the gas

concentration section in a FCC plant in Aspen HYSYS using information for a base case plant; investigate, based on the literature, the degrees of freedom in the process that affect the recovery of gasoline and LPG; investigate critically the effect of the identified degrees of freedom on the recovery of gasoline and LPG in the primary absorber; and lastly, formulate general conclusions and to develop an understanding about which degrees of freedom have the most significant impact on yield of gasoline and LPG in the gas concentration unit. Methodology proposed in this work and research outcomes can be utilised by operators to improve the performance of FCC light-end separation 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. Results from effects of degrees of freedom (such as operating pressure, solvent temperature, compositions and flow rate) on product recovery are examined. Section 4 presents conclusions on the effect of the degrees of freedom on enhancing product recovery.

MATERIALS AND METHODS

The method proposed to fully address the research goals presented in Section 1 comprises three main steps, see Figure 2: data collection, process modelling and simulation, and process optimisation. Data collection step gathers all information required to conduct process improvement for gas concentration unit of a FCC plant, for example, process configuration, unit design specification, unit operating conditions, and design and operational degrees of freedom. Step 2, process modelling and simulation, involves developing a simulation model of the gas concentration unit (using data from a base case plant) in Aspen HYSYS. Step 3, process optimisation, examines the effect of degrees of freedom affecting product (gasoline and LPG) recovery using the gas concentration unit model. Details of each step, i.e., data collection, process modelling and simulation, and process optimisation are described in subsections 2.1, 2.2, and 2.3 respectively.

Data collection

In order to conduct the process improvement studies, basic data from an existing FCC plant, obtained from "Samples" files in Aspen HYSYS v7.3 is used to build a model of gas concentration unit in Aspen HYSYS. 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. 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 C₃₊ (gasoline and LPG components) hydrocarbons from light-ends exiting the FCC main fractionators.

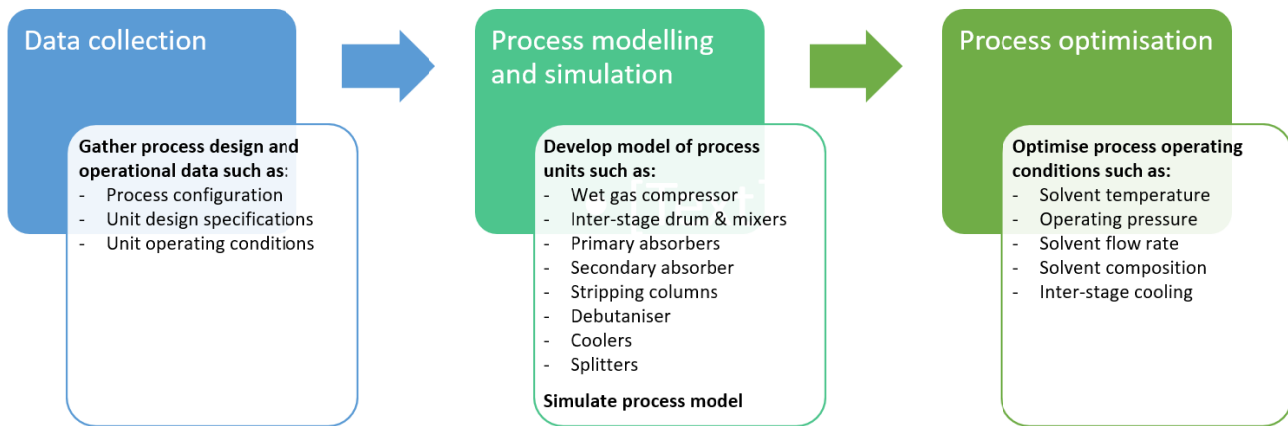


Figure 2. Proposed research methodology for improving the performance of petroleum refinery FCC light-ends separation

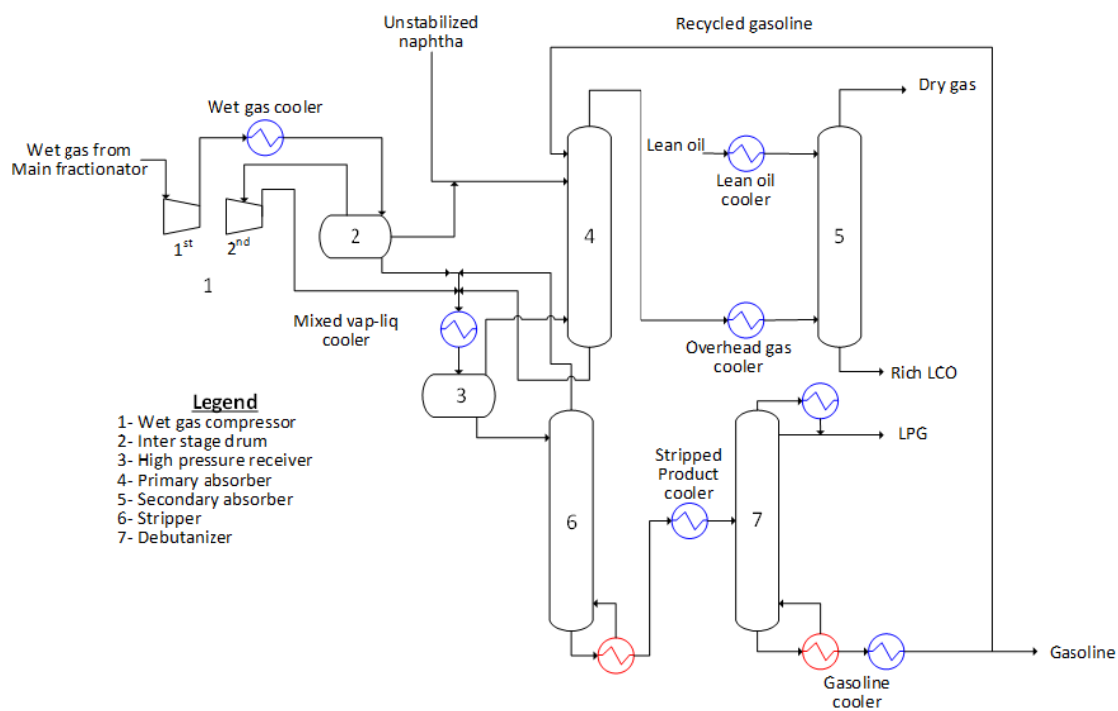


Figure 3. Process flow diagram of a base case gas concentration unit of a fluidised catalytic cracking unit plant

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_3 s and C_4 s) to enhance C_3 s and C_4 s hydrocarbon recovery.

Process modelling and simulation

Base case gas concentration unit of FCC plant shown in **Figure 3** is a complex arrangement of units including absorbers, stripper, debutanizer, coolers, heaters, separators and compressors. Modelling the gas concentration unit at once can cause convergence problem; hence, the modelling is carried out one step at a time. The process modelling is performed using SI unit set.

Defining simulation basis

Selection of an appropriate property package is crucial for accurate prediction of the behaviour of a particular process (Yela, 2004). Property package is selected based on nature of process, components, process temperatures and pressures etc (Yela, 2004). The current process constitutes hydrocarbons (both pure components and pseudo components), non-

hydrocarbon gases and water. Hence, a sufficient property package for the simulation is Peng-Robinson (Chang *et al.*, 2012).

Feed streams data

Feed streams are defined by specifying flow rate, composition and a minimum of two process conditions such as vapour fraction, temperature, pressure, heat flow etc. The unspecified stream conditions are then calculated by the software. The current simulation has three feed streams all coming from the main fractionators, which include wet gas, unstabilized naphtha and lean LCO, see Figure 2.2.

Installing and defining process equipment

Wet gas compressor (First stage): wet gas compressor (first stage) is the first equipment in the gas concentration unit. The unit can be modelled using either centrifugal or reciprocating model. The former has high volumetric flow rate and high discharge pressure while the latter is limited to low volumetric flow rate with high discharge pressure (Smith, 2005). Hence, the centrifugal model is used for the current simulation due to high flow rate and discharge pressure involved. The inlet stream condition is fully defined; unit parameters specified include adiabatic efficiency (%) and outlet pressure (kPa). **Wet gas cooler:** The cooler model has some key design parameter such as pressure drop, temperature difference and heat duty, with inputs such as input and output stream conditions. Specifying three variables enables the software to estimate the remaining variables. For the current simulation, the inlet and outlet temperature are specified together with pressure drop. The inlet stream condition is fully defined from the compressor simulation. **Inter-stage drum:** key design parameters of the three-phase separator model include inlet pressure and vapour outlet pressure, to determine flashing of the liquid stream. A three-phase separator is used for separating mixed liquid-vapour stream into heavy liquid, light liquid and vapour stream. Both inlet and vapour outlet pressure drops are set to 0 kPa. The inlet stream condition is known from wet gas cooler simulation. **Wet gas compressor (Second stage):** second stage wet gas compressor is installed and defined in the same approach as the first stage wet gas compressor. The inlet stream condition is fully defined from inter-stage drum simulation. **Solvent mixer:** mixer model has two parameter options namely 'equalize all pressure' and 'set outlet to lowest inlet'. Therefore, the second option is used for the current simulation since all the inlet streams pressure is known.

Primary and secondary absorber: inputs required for modelling the primary and secondary absorber include top pressure, bottom pressure, and no. of stages. Note that the top and bottom temperature are optional inputs. **Wet gas mixer:** wet gas mixer is modelled in a similar approach to the solvent mixer. **Recycle block:** recycle block compares the guess value (outlet) and the estimated value (inlet) during simulation and the block converges when the guess value equal estimated value. The inputs (guess values) used for the current simulation includes temperature, pressure and molar flow rate, which are 50°C, 1266 kPa and 78310 kmol/h respectively. **Mixed stream cooler:** mixed stream cooler is modelled using the same manner as the wet gas cooler. **High pressure receiver:** high pressure receiver is modelled in a similar way to the inter-stage drum, described above. Thus, the inlet stream condition is fully defined from 'Mixed stream cooler' simulation. **Lean oil**

cooler and overhead gas cooler: inlet stream conditions of the two units are fully defined. **Stripping column:** reboiled absorber model is used for modelling the stripper. Therefore, the data include unit design parameter and product specification – no. of stages, top stage pressure (kPa), boil-up ratio, bottom temperature (°C), H₂S fraction in bottom product. **Stripped product cooler:** inlet stream condition is fully defined from the stripper simulation.

Debutanizer: The distillation column model is used for modelling the debutanizer. Input data include unit specifications such as no. of stages, feed stage, condenser pressure, reboiler pressure, liquid rate, and reflux rate. After the debutanizer converge successfully, some key product specifications are then specified to ensure product quality meet consumer requirements, i.e., bottom product (gasoline) Reid vapour pressure (kPa) and C₅₊ mole fraction in LPG. Consequently, the reflux ratio and condenser temperature are no longer specified. The new active specifications become bottom product Reid vapour pressure and C₅₊ mole fraction in LPG. **Gasoline cooler:** inlet stream condition is fully defined from debutanizer simulation. **Splitter:** Tee model for splitter requires key parameter such as outlet stream flow rates. For the current simulation, only the flow rate of recycled gasoline is specified while the flow rate of gasoline product is determined by the software. The current approach enables improvement in gasoline recovery to be determined easily. Thus, the recycled gasoline flow rate specified is 48140 kmol/h, which is obtained from the Base case gas concentration unit. **Flow sheet convergence:** The 'recycled gasoline' stream is connected to the inlet of the Solvent mixer in Step 5. Therefore, the recycle block re-iterates until convergence is attained; during the converging process, the stripper, debutanizer, secondary absorber and primary absorber are all recalculated. Figure 4 shows the converged process flow diagram of the Base case gas concentration unit, which is used for the present investigation.

Process optimisation

The degrees of freedom affecting gasoline and LPG recovery discussed in Section 1 are used to develop five cases which are investigated independently. The detailed procedure for investigating each case is presented in the following sub-subsections.

Case 1 – Solvent temperature

The tendency for a solvent to dissolve or absorb a solute depends on the solubility of the solute in the solvent. Solubility decreases with increase temperature (Smith, 2005). Decreasing temperature of solvent during absorption increases the amount of solute dissolved and hence, improve gasoline and LPG recovery. Therefore, the temperature range for the case is selected such that expensive utilities (both heating and cooling) are avoided, also high temperatures are avoided to prevent rise in solvent vapour pressure which decreases gas solubility and results to loss of solvent during absorption. The temperature range used for the investigation includes 15°C, 25°C, 45°C and 55°C (Test 1, Test 2, Test 3 and Test 4 respectively); the base case solvent temperature is 35.4°C. A cooler or heater is installed to the gas concentration unit model in Aspen HYSYS, in order to adjust the temperature of the solvent entering the primary absorber. The key specification of the cooler or heater

includes pressure drop (set to 0 kPa). For each test, the outlet temperature of the cooler or heater is set accordingly.

Case 2 – Operating pressure

The operating pressure of an absorption process impacts directly on solubility of solute in a solvent. High solubility of solute in solvent is achieved at high pressure (Smith, 2005). Increasing the pressure in an absorber increases the partial pressure of the solute (Mehra, 1987) resulting in high recovery of solute. Operating at high pressure brings benefit to a process by enhancing recovery of solute. Moreover, while enhancing recovery in an existing process (revamp), the maximum increase in pressure is limited by the capacity of the gas compressor (Dean *et al.*, 2005) to avoid additional capital investment. Therefore, the operating pressure range is set according to the capacity limit of the wet gas compressor which is 1802 kPa. The operating pressure range used for the investigation 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. Each test pressure is examined by varying the operating pressure in the primary absorber of the gas concentration unit model. The pressure drop in the column is 22 kPa, which is equivalent to 2.4 kPa per tray (9 trays).

Case 3 – Solvent flow rate

The flow rate of an absorbent has influence on the amount of absorbed solute. The liquid to vapour (L/V) ratio in an important design parameter for absorption processes (Smith, 2005). Increasing the flow rate of solvent increases the solute recovery (Mehra, 1986). There are two solvents used for gasoline and LPG recovery in the primary absorber, namely unstabilized naphtha and recycled gasoline (debutanized gasoline). Unstabilized naphtha contains a high fraction of C₅+ hydrocarbons which can absorb heavier hydrocarbons; while debutanized gasoline is leaner in C₄- hydrocarbons, hence it has high capacity to absorb high amount of C₄- hydrocarbons. Therefore, the effect of the two solvents on gasoline and LPG recovery is examined separately in order to analyse the potential benefits of adjusting the flow rate of each solvent. The change in flow rate directly affects the liquid to vapour (L/V) ratio in the primary absorber; hence high gasoline and LPG recovery are expected at high L/V ratio. The 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. The flow rate range is usually set according to the capacity limit of unstabilized naphtha pump (Haik, 2005), but the capacity limit for the current study is unknown. Thus, a 17% increase or decrease of flow rate from the base case values is assumed; Haik (2005) used 38% increase in solvent flow rate in his study. The 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). The maximum increase in flow rates is 20% above the base case value, which is assumed to be within the capacity limit for the current study.

Case 4 – Solvent Composition

The chemical nature of lean oil (solvent) is an important factor in light-end recovery due to the molecular interactions

involved during the absorption process. Absorption of hydrocarbon depends strongly on molecular interactions (Haik, 2005) between the absorbate and solvent, therefore hydrocarbon solvents have greater capacity to absorb chemically similar compounds (Mehra, 1987); hence, lean oil that is leaner in paraffinic component tends to absorb greater amount of paraffinic component from the vapour phase so long as the vapour pressure of the lean oil is lower than the column operating pressure. Lean oil used for light-end recovery contains different hydrocarbons which are classified as paraffinic, olefinic, naphthenic and aromatic hydrocarbon. Hence, varying the flow rate of each class of hydrocarbon affects the solvent performance, thereby affecting gasoline and LPG recovery. The 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-pentene, 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)

Similarly, the recovery of gasoline and LPG is also affected by carbon number (Haik, 2005) of hydrocarbons present in the solvent. Choice of solvent for hydrocarbon recovery ranges between 2 or 3 carbon atoms heavier than the lightest component absorbed (Haik, 2005). The optimum lean oil is approximately three carbon atoms heavier than the lightest component to be absorbed (Branan, 2002). The lightest component to be absorbed in the current study is propane (C₃); hence, the effect of carbon number is investigated by increasing the flow rate of hydrocarbons which include pentane (C₅), hexane (C₆), heptanes (C₇) and octane (C₈), (Test 1, Test 2, Test 3 and Test 4 respectively).

Components such as hexane, heptane and octane are not present as pure components in the unstabilized naphtha stream. Hence, pseudo-components with similar properties (see Table 3.10) to hexane, heptane and octane were used instead. The pseudo components are already in the oil characterization; pseudo components in the unstabilized naphtha stream are presented in **Table 1**.

Table 1. Physical properties of selected pure and pseudo components

Property	Components					
	Pure nC6	Pseudo 60-70*	Pure nC7	Pseudo 90-100*	Pure nC8	Pseudo 120-130*
Boiling point (°C)	68	65	98	95	125	125
Molecular weight (g/mol)	86.19	84.74	100.20	103.60	114.23	119.0

*Pseudo-component name as represented in the oil characterization

For each test, the flow rate of the specific hydrocarbon (such as 60-70*, 90-100* etc.) in the unstabilized naphtha stream is increased by 50 kmol/h. (from the base case flow rate 10.1 kmol/h, 15.1 kmol/h, 27.4 kmol/h and 37.6 kmol/h respectively)

Case 5 – Inter-stage cooling

Absorption is accompanied by a release of heat at the interface between gas and solvent (Treybal, 1981), resulting from latent heat of absorbed component (Dean *et al.*, 2005). This effect

“changes physical properties, mass transfer coefficients and equilibrium concentrations” (Treybal, 1981) in the absorption column. The heat released increases the temperature within the column, therefore decreasing gas solubility (Smith, 2005). Therefore, providing cooling at intermediate location along the column reduces the temperature rise and improves absorption or product recovery (Smith, 2005). The temperatures of top and bottom tray in the primary absorber for the base case are 39°C and 50°C respectively, hence cooling is provided at two locations (Bumbac et al, 2007) which are chosen based on the column temperature profile in order to enable cooling at appropriate trays. Heat removal is carried out using two pumparounds with flow rates of $64\text{ m}^3/\text{h}$ (Haik, 2005) and the draw stage temperatures of 46°C (Tray 5) and 50°C (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°C , 26°C , 30°C and 34°C (Test 1, Test 2, Test 3 and Test 4 respectively).

RESULTS AND DISCUSSION

Simulation results from the case studies described in Section 2.1 are presented, analysed and discussed in this section. The results from the simulation of the gas concentration unit model, see **Figure 4**, shows that 691.9 kmol/h (67470 kg/h) and 1099 kmol/h (55830 kg/h) of gasoline and LPG is recovered from the process. 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). Thus, the C_3+ hydrocarbons in dry gas could be reduced by improving recovery of gasoline and LPG in the process, which in turns increase benefit.

Recall, the case studies involve investigating the impact of degrees of freedom such as solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling on product (gasoline and LPG) recovery, and heat recovery between the FCCU and the gas concentration unit. Therefore, Sections 3.1, 3.2, 3.3, 3.4, and 3.5 presents and discusses results on effect of solvent temperature, operating pressure, solvent (unstabilized naphtha) flow rate, solvent (unstabilized naphtha) composition, and inter-stage cooling respectively. Section 3.6 compares the best performance indicator obtained from each case study

Effect of solvent temperature to the primary absorber

The performance indicators from the investigation of solvent temperature on recovery of gasoline and LPG are presented in **Figure 5**. From the results, increase in solvent temperature by approximately 5°C from the base case temperature, decreases gasoline and LPG recovery by 1% and 0.4% respectively, while the C_3+ hydrocarbon (gasoline and LPG fraction) components in dry gas is increased by 21%. Conversely, decrease in solvent temperature by approximately 5°C from the base case temperature, increases gasoline and LPG recovery by 1% and 0.4% respectively, while the fraction of C_3+ hydrocarbon (gasoline and LPG fraction) in dry gas decreases by 26%. Similar trends are observed at 10°C increase or decrease in solvent temperature from the base case temperature. Haik (2005) observed similar trend by reducing lean oil temperature from 55°C to 37°C and therefore, recovery of LPG is increased slightly (amount of improvement not reported).

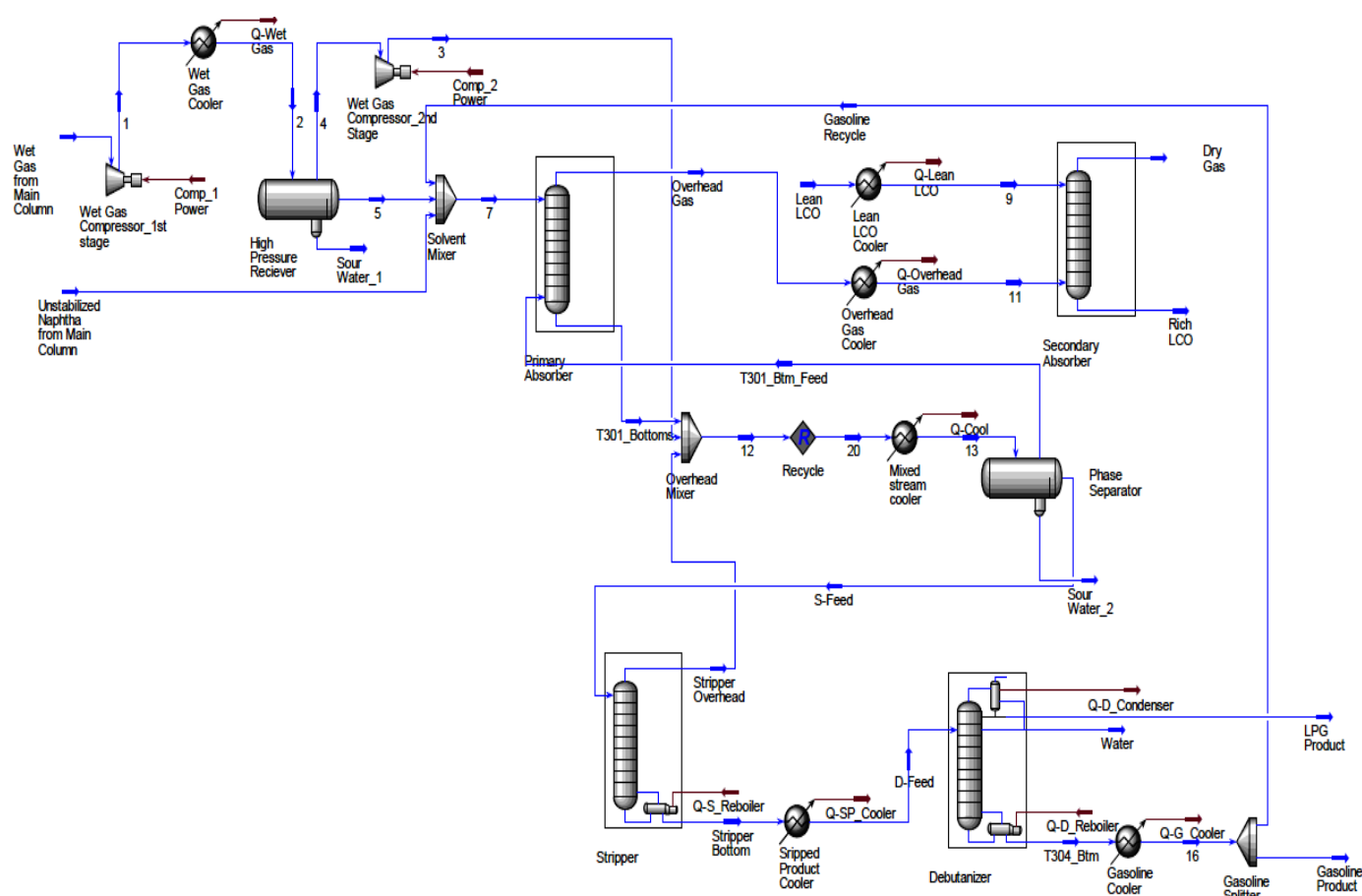


Figure 4. Simulation model of the base case gas concentration unit of a fluidised catalytic cracking unit

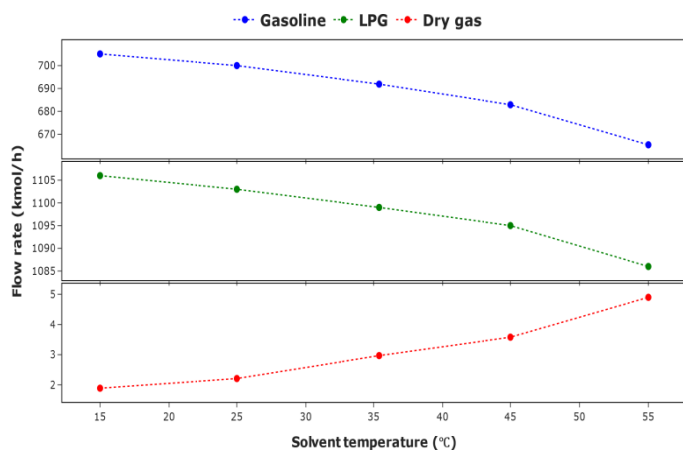


Figure 5. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with solvent temperature

High gasoline and LPG recovery and less C_3+ hydrocarbon in dry gas are observed at lower solvent temperature. For ΔT_{min} of $10^\circ C$, operating at $15^\circ C$ or $25^\circ C$ requires chilled water at $5^\circ C$ or $15^\circ C$ respectively, which will require a refrigeration system, therefore huge capital investment is required. On the other hand, operating at $45^\circ C$ and $55^\circ C$ requires ambient cooling medium, but decreases product recovery and leads to high C_3+ hydrocarbon in dry gas. Thus, operating at low solvent temperature requires expensive utility, while high temperature reduces revenue (low product recovery); hence the best solvent temperature for the current study is $35.4^\circ C$, which is the base case value. The recoveries of gasoline and LPG obtained at $35.4^\circ C$ are selected for heat integration analysis and profit evaluation for comparison with other cases.

Effect of operating pressure of primary absorber

Figure 6 presents the case results of effect of operating pressure of primary absorber on gasoline and LPG recovery. Increasing operating pressure by 400 kPa, increases gasoline and LPG recovery by 1 mol% and 0.4 mol% respectively, while C_3+ hydrocarbons in dry gas is reduced by 26%. Conversely, decreasing operating pressure by 400 kPa decreases gasoline and LPG recovery by 2 mol% and 0.7 mol% respectively, while C_3+ hydrocarbons in dry gas is increased by 40%. Similar trends are observed at 200 kPa increase or decrease of operating pressure.

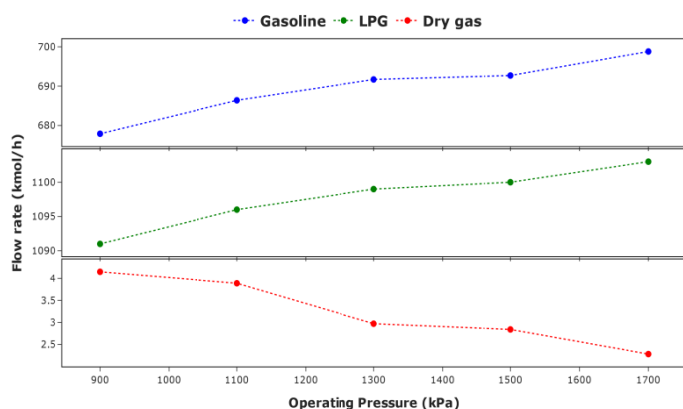


Figure 6. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with operating pressure

High operating pressure reduces lost C_3+ hydrocarbons in dry gas which in turns increases gasoline and LPG recovery. Haik

(2005) reported similar findings, obtained by increasing the operating in primary absorber from 1806 kPa to 2082 kPa to achieve 3% reduction of LPG in dry gas. Operating at high pressure improves recovery of gasoline and LPG (increase revenue) and reduces lost LPG in dry gas. Hence, the best operating pressure for the current study is 1700 kPa, which is well below the design operating pressure of the wet gas compressor (1802 kPa). Operating pressure above 1802 kPa requires replacement of the existing compressor, which leads huge capital investment.

Effect of solvent (unstabilized naphtha) flow rate to the primary absorber

Figures 7 and 8 present the effects of solvent flow on gasoline and LPG recovery by varying unstabilized naphtha flow rate and recycle gasoline flow rate. An increase in unstabilized naphtha flow rate by 50 kmol/h increases gasoline and LPG recovery by 7% and 0.6%, and decreases C_3+ hydrocarbon lost to dry gas by 6%. Comparatively, decreasing the flow by 50 kmol/h decreases gasoline and LPG recovery by 12 mol% and 0.83 mol%, and increases lost C_3+ in dry gas by 0.5%. Similar observations were made for increase or decrease of unstabilized naphtha flow rate by 100 kmol/h.

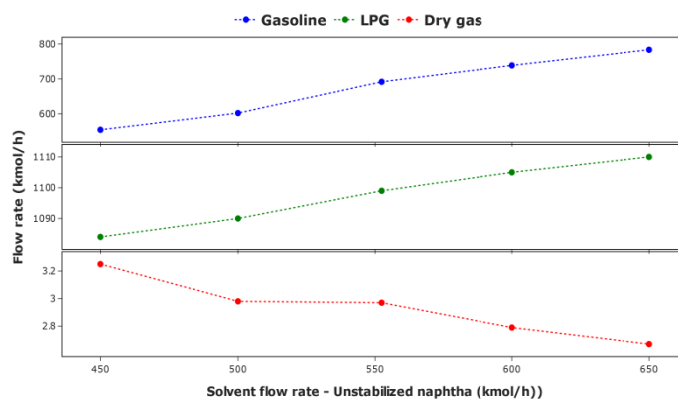


Figure 7. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with solvent flowrate (unstabilized naphtha)

The unstabilized naphtha flow has more influence on gasoline recovery than LPG recovery in the primary absorber, which is because of tendency of the predominant C_5+ hydrocarbons in the unstabilized naphtha to absorb chemically similar hydrocarbons in the feed gas (Mehra, 1986). Therefore, high unstabilized naphtha flow should be selected for optimum operation, but it should be within the capacity limit of the primary absorber. Thus for the current study, 650 kmol/h is the best unstabilized naphtha flow rate.

Figure 8 shows the effect of changing recycled gasoline flow rate: increasing the flow of recycled gasoline by 9628 kmol/h (20% of initial flow) increases the recovery of gasoline and LPG by 0.1 mol% and 0.3 mol%, and decreases lost C_3+ in dry gas by 17%. Conversely, decreasing the recycled gasoline flow rate by the same amount decreases gasoline and LPG recovery by 0.4 mol% and 0.1 mol%, and increases lost C_3+ in dry gas by 5%. Change in recycled gasoline flow rate impacts more on LPG recovery than gasoline recovery, which results from the fact that recycled gasoline has high absorption capacity for C_4- hydrocarbon than C_5+ hydrocarbons. Similarly, Haik (2005) achieved 5% decrease of lost LPG in dry gas by increasing recycled gasoline flow by 62%. Hence, recycled gasoline flow rate of 57768 kmol/h is the best choice for the current study.

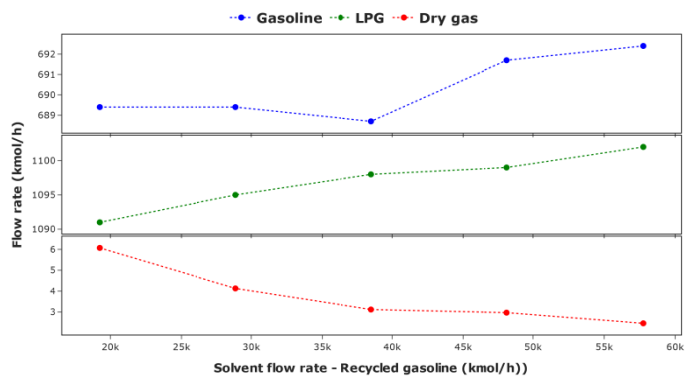


Figure 8. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with solvent flowrate (recycled gasoline)

Quantitative comparison of the best flow rate in **Figures 7 and 8** show that unstabilized naphtha recovers more gasoline and LPG than debutanized gasoline. Therefore, increasing unstabilized naphtha flow results to higher gasoline and LPG recovery than recycled gasoline in the process.

Effect of Solvent (unstabilized naphtha) Composition

The investigation results of the effect of solvent (unstabilized naphtha) composition on recovery of gasoline and LPG is presented in **Figures 9 and 10**. **Figure 9** presents the effect of hydrocarbon class (such as paraffin, olefin etc.) on product recovery, while **Figure 10** presents the effect of carbon number (which affects molecular weight of solvent) on product recovery.

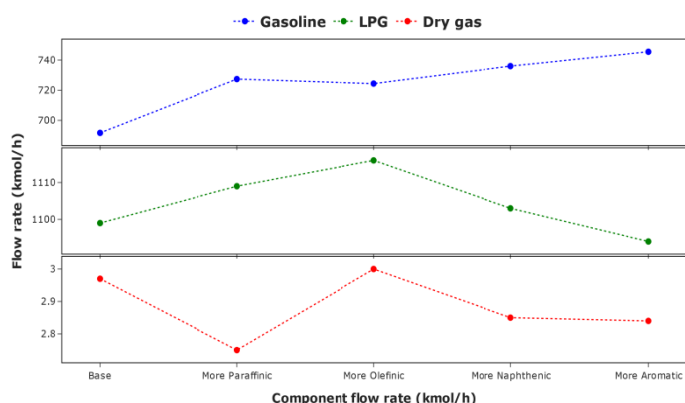


Figure 9. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with component flowrate

Based on the results in **Figure 9**, the following deduction can be drawn:

- Increasing the paraffinic content of the solvent by 50 kmol/h of n-pentane, increases gasoline and LPG recovery by 5% and 0.9% respectively, and decreases C_3+ hydrocarbons in dry gas by 7.4%.
- Increasing the olefinic content of the solvent by 50 kmol/h of 1-pentene, increases the recovery of gasoline and LPG by 5% and 1.5% respectively, and increases C_3+ hydrocarbons in dry gas by 1%.
- Increasing naphthenic content of the solvent by 50 kmol/h of cyclopentane, increases the recovery of gasoline and LPG by 6% and 0.4% respectively, and decreases C_3+ hydrocarbons in dry gas by 4%.
- Increasing the aromatic content of the solvent by 50 kmol/h of benzene, increases the recovery of gasoline by 8% and

decreases LPG recovery by 0.45 %, and decreases C_3+ hydrocarbons in dry gas by 4.4%.

Hence, there is a trade-off in selection a solvent composition that enhances recovery of both gasoline and LPG. High aromatic solvent is preferred for high gasoline recovery, but impact negatively on LPG recovery, while high olefinic solvent is preferred for high LPG recovery, but gasoline recovery is low compared with aromatic and naphthenic solvent. For the current study, gasoline is more valuable than LPG; therefore high aromatic solvent is selected as the best option.

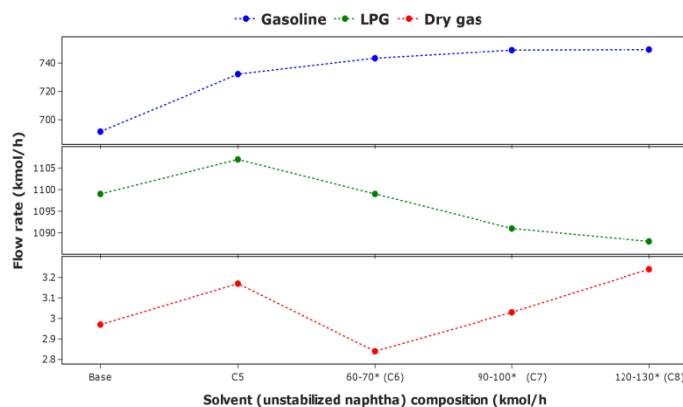


Figure 10. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with solvent (unstabilized naphtha) composition

A similar deduction is carried out for the results obtained from the effect of solvent (unstabilized naphtha) carbon number on gasoline and LPG:

- Increasing the C_5 content of the solvent by 50 kmol/h of n-pentane, increases gasoline and LPG recovery by 6% and 0.7% respectively, and increases C_3+ in dry gas by 6.7%.
- Increasing the C_6 content of the solvent by 50 kmol/h of 60-70* (pseudo component), increases the recovery of gasoline by 7.5% and has no effect on LPG recovery. Lost C_3+ in dry gas decreases by 4.4%
- Increasing C_7 content of the solvent by 50 kmol/h of 90-100* (pseudo component), increases the recovery of gasoline by 8.3% and decreases the recovery of LPG by 0.7%. Lost C_3+ in dry gas increases by 2%
- Increasing the C_8 content of the solvent by 50 kmol/h of 120-130* (pseudo component), increases the recovery of gasoline by 8.4% and decreases LPG recovery by 1%. Lost C_3+ in dry gas increases by 9.1%

Hence, solvent containing hydrocarbons with high carbon number (heavy solvent) improves the recovery of gasoline, but decreases LPG recovery compared with the base case, while solvent containing hydrocarbons with low carbon number (less heavy solvent) improves the recovery of LPG, but gasoline recovery is low compared with results obtained for high molecular weight solvent (high carbon number). For high recovery of gasoline and LPG, solvent containing high C_6 hydrocarbons is selected as the best option. Comparing the selected modifications in **Figures 9 and 10** shows that high aromatic solvent recovers high gasoline and slightly reduces LPG recovery, while C_6 (60-70*) enhanced solvent has more effect on gasoline recovery. Hence, selecting the best option depends on circumstance i.e., the demand of either gasoline or LPG. For the current study, gasoline has high market value

compared to LPG; therefore C₆ enhanced solvent is selected as the best option. In practice, stream rich in C₆ from other refinery units such as alkylation unit, polymerization unit etc. could be used in the light-end separation process.

Effect of inter-stage cooling in primary absorber

Figure 11 presents the case results for effect of inter stage cooling on recovery of gasoline and LPG. From the results, at pumparound return temperature of 22°C, recovery of gasoline and LPG is almost same as base case. At return temperature of 26°C, the recovery of gasoline decreases by 0.16%. Similarly, at 30°C and 34°C, gasoline recovery decreases further by 0.19%, while LPG recovery remains same as base case.

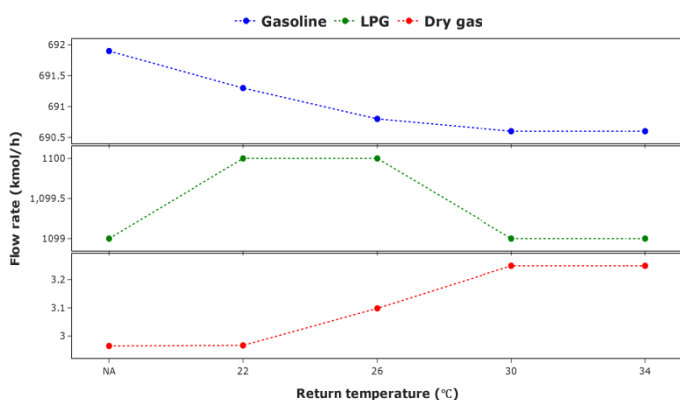


Figure 11. Amount increase/decrease in product flowrate (gasoline and LPG) or dry gas with pumparound return temperature.

Based on the present investigation, decreasing pumparound return temperature has more influence on LPG recovery compared with gasoline recovery, because, the partial pressure of LPG components is higher than gasoline components in the primary absorber. Test 1 has the highest gasoline and LPG recovery among the four tests, but requires sub ambient cooling, which is expensive; therefore, the base case is selected as the best option.

Comparison of performance indicator for case studies

The data in Table 2 presents the performance for selected modifications in each case study.

Table 2. Result for best performance for case studied

Case	Degree of freedom	Gasoline flow rate (kg/h)	flow (kmol/h)	LPG flow rate (kg/h)	flow rate (kmol/h)	Dry gas flow rate 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

Summary of effect of degrees of freedom on recovery of gasoline and LPG in the gas concentration unit of a FCC plant:

i. Decreasing solvent temperature entering the primary absorber to below ambient condition improves gasoline and LPG recoveries, consequently, sub ambient cooling required, which is expensive. On the other hand, solvent

temperature above ambient decreases gasoline and LPG recovery. Therefore, solvent temperature that requires ambient cooling is the best for the current study.

- ii. Increasing the operation pressure of primary absorber in the gas concentration unit by 400 kPa increases recoveries of gasoline and LPG by 1% and 0.4%, and reduces C₃+ hydrocarbons in dry gas by 26%, while decrease in operating pressure by 400 kPa decreases gasoline and LPG recovery by 2% and 0.7%, and C₃+ hydrocarbons in dry gas increases by 40%
- iii. Increasing the flow rate of debutanized gasoline by 20%, increases recovery of gasoline and LPG by 0.1% and 0.3%, and decreases hydrocarbon lost in dry gas by 17%, while decreasing recycled gasoline flow decreases gasoline and LPG recovery by 0.4% and 0.1%, and increases lost C₃+ in dry gas by 5%. On the other hand, recoveries of gasoline and LPG are improved by 7% and 0.6%, C₃+ in dry gas decreases by 6% due to increase unstabilized naphtha flow by 50 kmol/h. Decreasing unstabilized naphtha flow by 50 kmol/h decreases gasoline and LPG recovery by 12% and 0.83%, and C₃+ in dry gas is increase by 0.5%.
- iv. Enhancing solvent composition by separately increasing flow of n-pentane, 1-pentene, cyclopentane and benzene by 50 kmol/h in the unstabilized naphtha stream, improves product recovery. Benzene produces the highest beneficial improvement of 8% increase in gasoline and 0.45% decrease in LPG recovery. Lost C₃+ hydrocarbons in dry gas decreases by 4.4%. On the other hand, increasing flow of C₅, 60-70* (pseudo component), 90-100* and 120-130* in the unstabilized naphtha stream improves product recovery. The high beneficial improvement is observed by increasing the amount of 60-70* hydrocarbons, which increases gasoline recovery by 7.5% but does not improve LPG recovery. lost C₃+ hydrocarbons in dry gas decreases by 4.4%
- v. Inter stage cooling with pumparound return temperature in the range of 22°C to 34°C does not improve gasoline recovery and LPG recovery is almost same as base case.

Conclusion

This research work investigates degrees of freedom that can impact gasoline and LPG recovery in a gas concentration unit of a FCC plant and proposes a methodology for improving gasoline and LPG recoveries. The effect of degrees of freedom (solvent temperature, operating pressure, solvent flow rate, solvent composition and inter-stage cooling) on gasoline and LPG recovery has been investigated by adjusting each degree of freedom independently and the corresponding improvement of gasoline and LPG recovery are monitored, thus providing insight into which degrees of freedom have high influence on gasoline and LPG recovery. In conclusion, 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. 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. 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. C₇, C₈) recovers more gasoline than LPG, while solvent dominated by low carbon number hydrocarbons (e.g. C₅, C₆) recovers more LPG than gasoline. Therefore, three carbon atoms heavier than C₃ is the best (i.e. C₆) for gasoline and LPG recovery. Increase of C₆ in solvent (unstabilized naphtha) has the highest impact on overall 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. Based on the current study, it has been noticed that the FCC light ends separation unit is an overall heat source, with high potentials for steam generation and heat recovery. Integrating the FCC light end separation unit with other units in the refinery that are overall heat sinks (e.g. FCC reactor) would enable transfer of considerable amount of heat or steam between the two plants. Future work could consider enhancement of gasoline and LPG recovery simultaneously with heat recovery between the FCC light end separation unit and the FCC reactor. Also, the design of the heat exchanger network for heat recovery in the process could be considered.

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