CHAPTER 3

EXPERIMENTAL PROCEDURES

3.1 INTRODUCTION

The experimental procedures employed in the present work are aimed towards the evaluation of an industrial hydrotreating/hydrocracking catalyst system in a pilot plant trickle bed reactor unit under varied operating conditions. The pilot plant data are used to estimate kinetic parameters for various kinetic and modeling schemes proposed for different hydrocracking reactions and for product selectivity schemes. The experimental data are also used to validate the various kinetic reaction schemes developed in the present work and simulate the performance of pilot plant and industrial hydrocracking units based on the kinetic parameters obtained from pilot plant experiments over a wide range of operating conditions. The details of pilot plant hydrocracking experiments and the analysis of feed and product samples are discussed in this chapter.

The experimental procedures are also oriented to study the integration of hydrocracking unit with other important process units in a refinery especially fluid catalytic cracking unit and lube oil base stock manufacturing unit. Mild hydrocracking is increasingly being considered as an option for the pretreatment of FCCU feedstock. Integration of MHC unit with that of FCCU is beneficial in terms of improved product quality and yield pattern from the later unit. The details of pilot plant experimental studies on mild hydrocracking of vacuum gas oil and MAT studies using MHC
bottoms as feedstock are also discussed in this chapter. To explore the possibility of producing high quality lube base oils from hydrocracker bottoms, pilot plant studies have been conducted for isodewaxing of hydrocracker bottoms. The results of pilot plant isodewaxing studies have been compared with the solvent dewaxing data obtained from laboratory scale dewaxing unit. The details of these pilot plant and laboratory scale dewaxing studies are also discussed in this chapter.

3.2 PILOT PLANT EXPERIMENTS - HYDROCRACKING

3.2.1 Pilot Plant Reactor

The experiments were conducted in a continuous down flow isothermal trickle bed pilot plant reactor system with two reactors operated in series. The pilot plant reactors were designed to operate at a maximum temperature of 550°C and a pressure of 250 kg/cm². The reactors are made of stainless steel-316 H with an internal diameter of 1.7 cm and a length of 270 cm. The pilot plant reactors were operated in isothermal mode by independent temperature control of five zone electric furnace. The electric furnace has five heating zones and temperature of each of these zones can be controlled at ± 1°C. The schematic diagram of the hydrotreating pilot plant unit is shown in Figure 3.1. The photograph of the pilot plant reactors is shown in Figure 3.2.

3.2.2 Experimental Details

In the present work, a hydrotreating/hydrocracking catalyst samples procured from an industrial catalyst supplier was evaluated using vacuum gas oil feedstock collected from one of the crude distillation units operating at Chennai Petroleum Corporation Limited. To study the effect of feedstock on conversion, product yields, product selectivity and product quality, the
Figure 3.1  Schematic Diagram of Hydrotreating Pilot Plant

Figure 3.2  Photograph of Hydrotreating Pilot Plant
catalyst evaluation was also performed using a blend of 80% VGO and 20% Deasphalted Oil (DAO) at varied operating conditions. The physical characteristics of the catalyst samples and properties of feedstocks are discussed in chapter 5.

### 3.2.3 Operating Conditions

Operating conditions that are very important in the operation of hydrocracker unit are reactor temperature, liquid hourly space velocity, operating pressure and \( \text{H}_2/\text{oil} \) ratio. These operating parameters have strong influence on feedstock conversion, product yields and product quality. Out of these four operating parameters, reactor temperature and liquid hourly space velocity are the most important and normally these two parameters are varied to get desired conversion and product quality from an industrial reactor for a given catalyst and feed system. Operating pressure is not normally used as an operating variable since it is limited by design pressure of the unit. But operating pressure has strong influence on conversion, product yield and product quality during hydrocracking process. The \( \text{H}_2/\text{oil} \) ratio also is not used as an operating variable and is usually kept at an optimum level to achieve desirable cycle length of the catalyst. The details of the operating conditions employed in experimental studies are presented in chapter 5.

### 3.2.4 Catalyst Loading Pattern

The hydrocracking experiments were carried out in the pilot plant unit using similar catalyst loading pattern as that of industrial hydrocracker unit. The industrial hydrocracker unit has two reactors operated in series mode. The first reactor of the hydrotreating pilot plant system was loaded with 83 ml of hydrotreating catalyst at the top and 30 ml of dual functional hydrocracking catalyst at the bottom of the catalyst bed. The top 237 ml of the
reactor volume was filled with inert Al$_2$O$_3$ spheres to ensure homogeneous flow distribution of the feed streams, to heat the reactants to the required temperature as a preheat zone and to avoid end effects. The bottom 150 ml of the reactor volume was filled with inert Al$_2$O$_3$ spheres to serve as disengaging section.

The second reactor was loaded with 172 ml of dual functional hydrocracking catalyst in the middle of the reactor. The second reactor was loaded with maximum amount of hydrocracking catalyst to achieve moderate conversion of feedstock under pure catalyst environment. The top 178 ml and the bottom 150 ml of the reactor volumes were filled with inert Al$_2$O$_3$ spheres to work as preheating and disengaging sections for the second pilot plant reactor.

After loading the catalyst in both the reactors, the pilot plant unit was pressure tested to check for any leaks in the system. It was pressurized with nitrogen to 10 kg/cm$^2$ higher than the required operating pressure and then isolated from the cylinder. If the pressure loss was less than 10% of the system pressure for specific time duration, it was assumed that there were no leaks. Otherwise, all the tube fittings and valves were thoroughly tested with liquid soap and leaks were arrested.

### 3.2.5 Catalyst Presulfiding

The hydrotreating and hydrocracking catalysts contain metals such as Ni, Mo and W that are generally supplied in the form of metal oxides on γ-Al$_2$O$_3$ by catalyst suppliers. The catalysts are to be presulfided to convert these metal oxides into active metal sulphide phase which can catalyze hydrogenation/hydrotreating reactions.
The hydrotreating and hydrocracking catalyst samples were dried and presulfided prior to evaluation using n-heptane spiked with dimethyl disulfide (DMDS) as per a standard procedure obtained from the catalyst supplier. The following catalyst presulfiding procedure was employed in the present work for hydrocracking catalyst system.

Before starting the presulfiding procedure, two sulfiding solutions were prepared with 1.7 LV% DMDS in n-heptane (Solution A) and 6.8 LV% DMDS in n-heptane (Solution B). Initially nitrogen flow rate of 550 liters/hr was established through the catalyst system at atmospheric pressure. The temperature of the catalyst bed in both the reactors was increased to 232°C at the rate of 56°C per hour and held for one hour at the same temperature. Hydrogen flow was then established in place of nitrogen with the same flow rate at atmospheric pressure. The pressure of the unit was increased to 7 bars while maintaining other conditions and held for 2 hours at these conditions.

The Solution A was introduced at the rate of 300 ml/hr with all other conditions remain the same for 3 hours. Then the temperature was increased to 288°C at the rate of 60°C per hour and held for 3 hours. The pressure was increased to 55 kg/cm² and temperature to 315°C at the rate of 60°C per hour. Then the feed was changed to Solution B at the rate of 300 ml/hr and same conditions were continued for 3 hours. The temperature was increased to 343°C at the rate of 60°C per hour and held for 3 hours. The temperature was decreased to 232°C and the pressure and hydrogen flow rate were changed to the test conditions. The feed was changed to test feed and the conditions were maintained for one hour at 232°C. The temperature was increased to the required run temperature at the rate of 30°C per hour before starting pilot plant experiments.
3.2.6 Experimental Procedure

After presulfiding both the hydrotreating/hydrocracking catalyst system loaded in two pilot plant reactors operated in series, the pilot plant experimental runs were carried out. The experimental conditions were chosen such that severemost operating conditions were employed first to avoid experimental errors arising due to product contamination with line content. When the feedstock was changed during the course of pilot plant experiments, the plant was continuously run with the changed feedstock for one day at the typical operating conditions to replace all the line content of the previous feedstock. Similarly, when the operating conditions of the experiments were changed, the pilot plant was stabilized continuously for eight hours before actual samples collected for TBP distillation and product analysis.

Initially the hydrogen flow was set at required level calculated from the liquid hourly space velocity and H₂/oil ratio of the experimental run. The hydrogen flow to the system was controlled by a Bronkhorst electronic mass flow controller. The pressure of the reactor was controlled by a pressure control valve located at the gas outlet of high pressure separator. All the reactor furnace and line heating were switched on and required temperature was set for each reactor zone and line heaters. The shell (outer wall of the reactor) and catalyst bed temperatures of the reactor were monitored. The catalyst bed temperature was monitored by five multi-point insertion thermocouples located in a thermowell mounted co-axially at the center of the reactor. These thermocouples measure the temperature of the catalyst bed at five different points along the catalyst bed length.

The feedstock was introduced into the system after the system attained required pressure for the experimental conditions and when all the line heaters and reactor zones were heated sufficiently to avoid solidification
of feedstock. The feed rate was regulated at required level by a variable stroke length high pressure pump.

The total reactor products were sent to a high pressure separator where the gas and liquid phases got separated. The liquid product was collected from high pressure separator through a level control valve by maintaining a preset liquid level. The liquid product from high pressure separator was sent to a low pressure separator in which some of the remaining gases were separated from the liquid fraction. The liquid from low pressure separator was sent to a product stripper before being collected in a product tank. The product gases from the separators were washed with caustic solution in a caustic scrubber to remove acid gases such as hydrogen sulphide and ammonia. The gases from the scrubber were metered through a wet gas meter before venting out to the atmosphere. A gas sample was also collected for GC analysis.

Each experimental run was continued for more than 40 hours under steady state conditions to collect sufficient liquid products for distillation in a true boiling point distillation apparatus to separate various product fractions such as naphtha, kerosene, jet fuel, diesel and unconverted oil. The gas products from the separators were analyzed in a Refinery Gas Analyzer (RGA) to get their composition. The amount of hydrogen sulfide and ammonia formed during the experimental run was estimated from material balance.

During each run, material balance was performed for one hour every 6 hours to check for any deviation from steady state. This was done to ensure that the data obtained from pilot plant studies was fairly accurate. A material balance of ±2 wt% was obtained during the experimental runs.
3.2.7 TBP Distillation of Hydrocracked Products

Hydrocracked product samples at varied operating conditions were collected continuously at steady state conditions in the pilot plant for subsequent True Boiling Point (TBP) distillation. For each experimental run about 10-20 litres of hydrocracked product sample was collected for further distillation into various product fractions such as naphtha, jet fuel, kerosene, diesel and unconverted oil.

The TBP distillation of hydrocracked products was done according to ASTM D-2892 in a state of the art, processor controlled TBP distillation pilot plant procured from Fischer, Germany. The TBP distillation unit is a fully automated distillation system from the start of the distillation up to the end of the pre-selected or detected end point, without interruption or operator intervention between the individual distillation stages, for a final temperature up to 400°C AET. The photograph of the TBP apparatus is shown in Figure 3.3.

Figure 3.3 Photograph of TBP Apparatus
There are many operating steps or distillation sequence involved during TBP distillation. In the debutanization step, LPG was collected up to a maximum vapor temperature of 15°C. After debutanization, distillation at atmospheric pressure was done to collect distillates up to a maximum vapor temperature of 190°C at a reflux ratio of 5:1 and distillation rate of 4-6 gms/min. During this step, naphtha and part of kerosene fractions were collected. After distillation at atmospheric pressure, distillation was performed using a vacuum of 100 mm Hg and called as first stage vacuum. In the first stage of vacuum, remaining kerosene fraction boiling up to a maximum vapor temperature of 270°C was collected at a reflux ratio of 5:1 and distillation rate of 6-8 gms/min. Distillation at second stage vacuum was performed at 10 mm Hg. During this stage, part of the diesel fraction was collected up to a maximum vapor temperature of 300°C at a reflux ratio of 5:1 and distillation rate of 8-10 gms/min. During the third stage of vacuum, distillation was conducted under a vacuum of 2 mm Hg at a reflux ratio of 2:1 and distillation rate of 10-12 gms/min. During third stage of vacuum, remaining diesel fraction was collected up to a maximum vapor temperature of 370°C. The product fractions obtained from TBP distillation unit were then subjected to detailed characterization for their important properties and reported for data analysis.

3.3 PILOT PLANT EXPERIMENTS - INTEGRATION OF HYDROCRACKER WITH FCCU

3.3.1 Pilot Plant Reactor

The pilot plant studies on mild hydrocracking of vacuum gas oil were carried out in a continuous flow high pressure trickle bed reactor system with a single reactor and was designed to operate at a maximum temperature of 600°C and a maximum operating pressure of 300 kg/cm². The reactor tube
was made of stainless steel-316 with an internal diameter of 2.54 cm and a length of 80 cm. The pilot plant reactor was operated in isothermal mode by independent temperature control of four zone electric furnace. The electric furnace has four heating zones and temperature of each of these zones can be controlled at ±1°C. The reactor temperature was monitored by four multi-point insertion thermocouples located in a thermowell mounted co-axially at the center of the reactor. The feed rate was regulated at required level by a variable stroke length high pressure pump.

The hydrogen flow to the system was controlled by an electronic mass flow controller. The pressure of the reactor was maintained by an electronic pressure sensor and solenoid valve system at the gas outlet of high pressure separator. The liquid product from high pressure separator was collected by maintaining preset liquid level with the help of a level control valve. The product gases were metered through a wet test meter before venting out. The experimental error for this study was within ±2 wt% of material balance. The schematic diagram of the high pressure pilot plant reactor system is shown in Figure 3.4. The photograph of the pilot plant reactor is shown in Figure 3.5.

Micro Activity Testing experiments were conducted using a computer controlled MAT unit designed to operate at a maximum pressure of 0.35 bar gauge. The MAT reactor was loaded with 4 grams of FCC equilibrium catalyst collected from industrial FCC unit. The photograph of MAT unit is shown in Figure 3.6.
Figure 3.4 Schematic Diagram of High Pressure Reactor System

TIC - Temperature Indication and Control
PIC - Pressure Indication and Control
FIC - Flow Indication and Control
LIC - Level Indication and Control
WI - Weight Indication
Figure 3.5 Photograph of High Pressure Reactor System

Figure 3.6 Photograph of MAT Unit
3.3.2 Experimental Details

A VGO feedstock originated from Escravos crude source and collected from industrial FCC unit of Chennai Petroleum Corporation Limited was used in the pilot plant studies. For mild hydrocracking experiments, there are two catalysts loaded in the pilot plant reactor as stacked bed arrangement in two beds. The first bed of the reactor was loaded with a hydrotreating catalyst that was used for pretreating VGO feedstock prior to MHC step. The second bed of the reactor was loaded with a MHC catalyst. For hydrotreating of VGO fraction, a hydrotreating catalyst was used in the pilot plant. All these catalysts samples were procured from an industrial catalyst supplier with well proven in industrial performance and used successfully in many refineries around the world for VGO hydrotreating and mild hydrocracking application. For MAT studies, CPCL FCC unit equilibrium catalyst was used for all the feedstocks. The physical characteristics of the catalysts are discussed in Chapter 5.

3.3.3 Operating Conditions

The experimental conditions for hydrotreating and mild hydrocracking were chosen based on typical operating conditions of industrial vacuum gas oil hydrodesulphurization units. MAT experiments were also conducted using untreated VGO, hydrotreated VGO and MHC VGO to compare conversion and yields from FCC. The details of operating conditions employed in experimental studies are presented in chapter 5 in detail.

3.3.4 Catalyst Loading Pattern

The high pressure pilot plant trickle bed reactor was loaded with 100 ml of industrial hydrotreating catalyst in the top section and 150 ml of
industrial mild hydrocracking catalyst in the bottom section. The hydrotreating catalyst loaded in the top section of the reactor removes nitrogen and sulphur prior to mild hydrocracking and thus protects zeolite based hydrocracking catalyst from poisoning. A stacked bed hydrocracker reactor configuration composed of conventional hydrotreating and zeolite based catalysts is shown to offer marked improvements in hydrocracker activity and catalyst stability compared to single bed systems (Esener and Maxwell 1989). The top 150 ml and bottom 100 ml of the reactor was filled with inert $\alpha$-$\text{Al}_2\text{O}_3$ spheres to serve as preheating and disengaging sections respectively. After catalyst loading in the high pressure reactor, the pilot plant unit was pressure tested to check for any leaks in the system.

3.3.5 Catalyst Presulphiding

Generally hydrotreating and MHC catalysts are supplied in the form of metal oxides by catalyst suppliers. These catalysts have to be presulphided to convert these metal oxides into metal sulphide phase, which can catalyze hydrotreating/mild hydrocracking reactions. The following catalyst presulphiding procedure was employed for HDT/MHC and HDT catalyst systems used in the present work.

Prior to experimental runs, the catalyst was dried and presulphided with atmospheric gas oil spiked with Di-methyl Di-sulfide so as to get 2.5 wt% sulphur in presulphiding medium. Initially the catalyst was purged with nitrogen at the rate of 100 litre/h for 30 minutes at atmospheric pressure and temperature. The temperature of the catalyst bed was increased to 100$^\circ$C and pressurized with $\text{H}_2$ to 10 kg/cm$^2$. The presulphiding medium was introduced at a liquid hourly space velocity of 2.0 h$^{-1}$ with a $\text{H}_2$/oil ratio of 120 m$^3$/m$^3$ and maintained for 3 hours. Then the liquid hourly space velocity was reduced to 1.3 h$^{-1}$ and the temperature was increased to 320$^\circ$C at a rate of
30°C/h. These conditions were then maintained for 3 hours to complete catalyst presulphiding. The catalyst bed was gradually cooled to ambient temperature and kept under H₂ atmosphere for the experiments.

### 3.3.6 Experimental Procedure

After presulphiding both the hydrotreating/ mild hydrocracking catalyst system loaded in the pilot plant reactor, the experimental runs were carried out. Initially the hydrogen flow was set at required level calculated from the liquid hourly space velocity and H₂/oil ratio of the experimental run. The hydrogen flow to the system was controlled by a Brooks electronic mass flow controller. The pressure of the reactor was controlled by a pressure control valve located at the gas outlet of high pressure separator. All the reactor furnace and line heating were switched on and required temperature was set for each reactor zone and line heaters. The catalyst bed temperature was monitored by four multi-point insertion thermocouples located in a thermowell mounted co-axially at the center of the reactor.

The liquid and gas product samples were collected under steady state conditions. The total reactor products were sent to a high pressure separator where the gas and liquid phases were separated. The liquid product was collected from high pressure separator through a level control valve in which a preset level automatically opens a solenoid valve to a product tank. The liquid from the product tank was sent to a stripping set up where the entrained hydrogen sulfide, ammonia and other light gases are stripped off from the liquid using nitrogen. The product gases were metered through a wet gas meter before venting out to the atmosphere. A gas sample was also collected for GC analysis.
After a stabilization period of 8 hours, each experiment was continued for around 4 hours under steady state conditions to collect sufficient liquid products for distillation in a Simulated Distillation (Simdist) apparatus. The gas products from the separator were analyzed in a Refinery Gas Analyzer to get their composition. The experimental runs at different conditions were repeated till a material balance of ±2 wt% was obtained.

3.3.7 Micro Activity Testing of Untreated, HDT and MHC VGO

Micro Activity Testing unit is a computer controlled system for performing MAT test to evaluate and compare the activity and selectivity of various feedstocks and catalyst system. The system is designed to operate at a maximum pressure of 0.35 bar g. Catalyst loading of about 4-5 gms is possible in the MAT reactor.

Generally the following operating modes are employed during MAT experiments:

- Purging - 15 minutes
- Reaction - 75 seconds
- Regeneration - 15 minutes
- Stripping - 15 minutes
- Cooling - 25 minutes

Similar operating conditions were used in MAT unit for the evaluation of MAT activity and product yields for all the three feedstocks, namely MHC VGO, HDT VGO and untreated VGO. The operating conditions employed for MAT evaluation is presented as follows:
Feedstock: MHC VGO/ HDT VGO/ Untreated VGO
Catalyst: FCC Equilibrium Catalyst
Catalyst Weight: 4 gms
Catalyst/Oil Ratio: 3
Residence Time: 75 secs
Feed Weight: 1.33 gms
Reactor Temperature: 482 °C
Regenerator Temperature: 650 °C

3.4 INTEGRATION OF HYDROCRACKER WITH LUBE BASE OIL UNIT

Studies were conducted in lab scale dewaxing unit and a high pressure trickle bed reactor system using hydrocracker bottom sample obtained from a fuels refinery hydrocracker unit to study various options for producing Group II base oils from hydrocracker bottoms.

3.4.1 Pilot Plant Reactor

The isodewaxing experiments were conducted in a continuous flow, high pressure reactor, same as the one used for conducting mild hydrocracking experiments (Figure 3.4). The details of the pilot plant system have already been discussed under section 3.3.

3.4.2 Experimental Details

In the present work, a hydrocracker bottom sample obtained from a fuels refinery hydrocracker unit was used at various operating conditions over indigenously developed Pt SAPO-11 isodewaxing catalyst. The isodewaxing
catalyst used for the experiments contained 0.5 wt% of platinum, around 60 wt% SAPO-11 and remaining being binder alumina.

### 3.4.3 Operating Conditions

As first technological option, studies were conducted to explore the manufacture of Group II base oils through isodewaxing process. The isodewaxing pressure was changed in the range of 60 to 140 kg/cm$^2$ at a constant liquid hourly space velocity of 1.0 h$^{-1}$, H$_2$/oil ratio of 400 m$^3$/m$^3$ and a reactor temperature of 340°C. The isodewaxed product was distilled in a pot still distillation unit to separate 370°C−, 370-400°C, 400-440°C, 440-500°C cuts.

As second option, studies were conducted to explore the possibility of producing Group II base oils from hydrocracker bottom using existing solvent dewaxing facilities. The hydrocracker bottom was subjected to vacuum distillation in a pot still distillation unit procured from Fisher, Germany and various fractions were collected and analyzed. Dewaxing experiments were conducted in lab-scale dewaxing unit using commercial solvent (MEK and Toluene with a solvent mix ratio of 50:50). The feed and solvent (solvent/feed ratio is 4:1) were heated up to 75°C and then gradually cooled to −17°C to crystallize the wax and then filtered through NS-3 filter cloth.

### 3.4.4 Catalyst Loading Pattern

The isodewaxing experiments were carried out using 200 ml of catalyst in 40 cm bed. The top 20 cm of the reactor was filled with 100 ml of inert Al$_2$O$_3$ spheres to ensure homogeneous flow distribution of the feed streams, to heat the reactants to the required temperature as preheat zone and
to avoid end effects. The bottom 20 cm of the reactor was filled with 100 ml of inert Al$_2$O$_3$ spheres to serve as disengaging section.

### 3.4.5 Catalyst Activation

The isodewaxing catalyst samples are to be activated prior to evaluation using a standard procedure. The reactor was pressured initially to 5 kg/cm$^2$ with 100 SLPH of nitrogen gas flow. The catalyst bed temperature was increased to 300$^\circ$C at the rate of 60$^\circ$C per hour. The catalyst bed temperature was then raised to 480$^\circ$C and maintained for 6 hours. The reactor temperature was reduced to the required run temperature of 340$^\circ$C and depressurized to 1 atm. Hydrogen was introduced at the rate of 200 SLPH and continued for 2 hours. The system pressure was increased to the required run pressure of 60 kg/cm$^2$ and hydrogen flow rate was changed to the required level. The feedstock was then introduced at required flow rate.

### 3.4.6 Experimental Procedure

After activation of the isodewaxing catalyst, the pilot plant experiments were carried out. The experimental procedure is the same as that described under section 3.3. The product analysis was carried out by gas chromatographic methods. The gaseous and liquid fractions were analyzed in a capillary column (PONA column 0.2 mm x 50 mm; GC, HP 5880). The liquid fraction was analyzed by simulated distillation (G.C. Chrompak, capillary column, ASTM D2887).

### 3.5 ANALYSIS OF FEED AND PRODUCT SAMPLES

The feedstocks, hydrocracked product samples and fractions obtained from TBP distillation were analyzed to determine their physical
properties and measure sulphur, nitrogen, saturates, aromatics etc. The feedstock and products collected from industrial reactor were also analyzed for their important properties. The analysis of various feedstocks and products were done as per ASTM and IP standard test methods. Table 3.1 presents the details of analytical tests, test method and equipment employed for the analysis of feed and product samples.

**Table 3.1  Details of Analytical Tests and Equipment**

<table>
<thead>
<tr>
<th>Test</th>
<th>Test Method</th>
<th>Test Equipment</th>
</tr>
</thead>
<tbody>
<tr>
<td>Determination of Sulphur Content</td>
<td>ASTM D 4294</td>
<td>X-ray Fluorescence spectrophotometer (Oxford Lab-X 3500)</td>
</tr>
<tr>
<td>Determination of Nitrogen Content</td>
<td>ASTM D 4629</td>
<td>Antek 703C</td>
</tr>
<tr>
<td>Determination of Saturates and Aromatics</td>
<td>IP 368/90</td>
<td>Waters 410 instrument</td>
</tr>
<tr>
<td>Determination of Light Hydrocarbons in Gas Products</td>
<td>-</td>
<td>Refinery Gas Analyzer (Chromepack)</td>
</tr>
</tbody>
</table>

3.5.1 **Determination of Physical Properties**

The physical properties of feedstocks and various product fractions obtained from TBP distillation of hydrocracked products were measured as per ASTM standard methods. Table 3.2 shows the list of test methods used for the analysis of feed and product samples.
### Table 3.2 List of Standard Test Methods Applied

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Description</th>
<th>Method No.</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Practice for Density, Relative Density (Specific Gravity) or API Gravity of Crude petroleum and Liquid Petroleum Products by Hydrometer Method</td>
<td>ASTM D 1298</td>
</tr>
<tr>
<td>2</td>
<td>Test Method for Kinematic Viscosity of Transparent and Opaque Liquids (The Calculation of Dynamic Viscosity)</td>
<td>ASTM D 445</td>
</tr>
<tr>
<td>3</td>
<td>Test Method for Aniline Point and Mixed Aniline Point of Petroleum Products and Hydrocarbon Solvents</td>
<td>ASTM D 611</td>
</tr>
<tr>
<td>4</td>
<td>Test Method for Flash Point By Pensky-Martens (Closed Cup) Tester</td>
<td>ASTM D 93</td>
</tr>
<tr>
<td>5</td>
<td>Test Method for Pour Point of Petroleum Products</td>
<td>ASTM D 97</td>
</tr>
<tr>
<td>6</td>
<td>Test Method for Ramsbottom Carbon Residue of Petroleum Products</td>
<td>ASTM D 524</td>
</tr>
<tr>
<td>7</td>
<td>Test Method for Distillation of Petroleum Products at Atmospheric Pressure</td>
<td>ASTM D 86</td>
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</table>