

# CHAPTER I

## INTRODUCTION OBJECT AND SCOPE OF THE STUDY

## 1. INTRODUCTION :

The necessity of lubricants have been felt since the invention of the wheel. Mostly castor oil had been used to lubricate the axle of the bullock cart wheels. This is still being practised in rural India. To cope with the thrust of modern technology which introduced more efficient engines designed for high speed and high load, a need for better quality of lubricants is being observed. The substantial rise of automobiles in India over the years has created a staggering demand for petroleum products and lubricating oils. The consumption pattern of different petroleum products in recent years is shown in TABLE 1.1.

TABLE 1.1 CONSUMPTION OF PETROLEUM PRODUCTS IN INDIA<sup>a</sup>

Petroleum fractions	1983-84	84-85	85-86	86-87	87-88	88-89	89-90
	(In million tonnes)						
Light distillates	5.6	6.28	6.91	7.39	7.54	8.56	9.4
Middle distillates	20.49	22.30	24.31	25.26	28.0	29.74	26.8
Heavy ends	8.93	9.24	9.65	10.45	10.86	11.44	11.8
Lube oils	0.47	0.41	0.50	0.49	0.48	0.49	0.54

<sup>a</sup> Indian Petroleum & Natural Gas Statistics (April-Dec) 1990, Ministry of petroleum & chemicals, Govt. of India.

TABLE 12 LUBE- BASE STOCK PRODUCTION CAPACITY IN INDIA

Oil Refineries	Production capacity million tonnes per year
1. Indian Oil Corporation Ltd (Haldia)	0.200
2. Madras Refineries Ltd (Madras)	0.140
3. Hindusthan Petroleum Corporation Ltd. (Bombay)	0.152
Total :-	0.492

The current overall national demand for lube stock is over 8 lakhs tonnes per year which is much higher than the indigenous production capacity of the lube producing refineries (TABLE 1.2.). A deficit of lube base stock around 0.32 million tonnes was met by import (Rs.183 crores) for the year 1989-90 (a). The projected deficit of lube base stock is around 0.75 million tonnes in the year 2000 taking into account of the proposed refining capacity of Indian refineries in 8th Five Year Plan. The alarming increase in import bills for petroleum crudes and its products over the years had a shattering effect on the economy of many developing countries including India. Crude output by ONGC had stagnated around 30 million tonnes per annum for 3 years, whereas demand is rising @ 8% per annum.

In the present scenario of supply and demand of petroleum products and its effect on the economy, a need arises to look into the application of new technology and utilisation of non conventional feed stocks for the production of different grades of lubricating oil. Therefore, the present status of different

technologies available for production of lubricating oils is reviewed keeping in view the feed stock characteristics, the process development and its limitations.

### 1.1 LUBE OIL MANUFACTURING PROCESS :

Lubricating oils are produced from heavier petroleum fractions within the boiling range of 350-550°C containing dissolved petroleum waxes as much as 40 % and other components such as isoparaffins, monocycloparaffins and aromatics (Edwards\*et at ,1957). The potential of different crudes to provide lube base stock varies due to the variation in there composition of its components as shown in TABLE-1.3.

Petroleum waxes are classified according to their physical forms. n.Paraffinic waxes form large crystals and they are obtained in light lubricating oil fraction while small wax crystals (amourphous), called microcrystalline waxes are generated by non-normal components which predominate in heavier cuts. Paraffin waxes are undesirable in motor oils because they crystallize out at lower temperatures. Hence removal of certain quantity of paraffin wax i.e; dewaxing is needed to maintain the flow properties of the oil at lower tempratures and it is measured empirically in terms of the physical property, FOUR POINT and it is defined as the highest temperature at which oil ceases to flow.

Separation of petroleum waxes from lube fraction by distillation is not feasible as it boils in the same temprature range. The necessity for production of lubricating oils suitable to flow at low temperature led to early dewaxing processes.

\*P.S. References cited in this chapter are listed in page No.178 onwards.

TABLE 13 LUBE-BASE STOCK POTENTIAL OF FEW CRUDES

Viscosity grades	wt % on crudes			
	North Gujarat	Bombay high	Gulf	Assam
Light viscosity	9.0	8.8	11.0	Unsuitable
Medium viscosity	12.37	5.7	10.5	"
Heavy viscosity	14.4	2.45	7.6	"

## 1.2. OLD DEWAXING METHODS :

### 1.2.1. CHILLING PROCESS :

The basic principles for dewaxing in the early 19th century were as given below :

- (1) Pressing and sweating in case of low viscosity oils.
- (2) Cold settling and centrifuging in case of high viscosity oils.

In 1880, the dewaxing of petroleum distillates was conducted by chilling to crystallize the wax and it was separated in plate and frame filter press. But the separation was rather inefficient, the wax cake still contained as much as 50-60% oil (Miller and Osborn, 1938). Heavier residual lubricating oil did not form large wax crystals on cooling and could not be handled in such process.

A new 'Extractive Crystallization' process was developed by Sharples Corporation in 1920 to handle the residual lube

oil. Naphtha was used as solvent in this process and the crystallized wax was separated from chilled solution by centrifugation. Naphtha was recycled after distilling off from the solution. The wax obtained from this process also contained 30-40% oil which was recovered by sweating (fractional melting) process producing the paraffin wax of commercial grade containing 0.5% oil (Jones, 1938). Labour cost and energy consumption in this process were high and it had the additional disadvantage of processing of lube fraction containing waxes of various nature.

A commercial unit capable of dewaxing the entire range of distillate and residual stocks, known as 'Weir process' was commissioned in 1928. Diatomaceous earth was used in this process as filter aid to the naphtha solvent in achieving the filterable cake. Continuous rotary vacuum filters were used to remove the slurry containing wax. Filter aid was recovered from the wax cake and recycled. This process was not accepted widely in industry due to its high operating cost.

### 1.2.2 SOLVENT DEWAXING

The rapid growth of the chemical industry resulted in the production of numerous organic solvents which exhibit low wax solubility at dewaxing temperature as well as high solvent power for the oil had been applied in dewaxing process. The older dewaxing methods were replaced with solvent dewaxing process. The criteria for a good commercial dewaxing solvent are as following :

- a) Sufficiently low boiling point to facilitate separation from the wax and dewaxed oil.

- b) Low sensible and latent heat for economy in chilling and recovery.
- c) Completely miscible with the oil at the temperature of filtration and incapable to dissolve wax.
- d) Non corrosive to equipment and non toxic to avoid health hazards.

The first commercial dewaxing unit employing organic solvents was installed in 1927 by Indiana Refining Company, Texas. A solvent mixture of acetone and benzene was used in which benzene was used as non polar solvent for solubilising the oil part and acetone used to facilitate the precipitation of wax crystals. Later this process was improved by substituting methyl ethyl ketone (MEK) for acetone and benzene was replaced with toluene. Another ketonic solvent, methyl iso butyl ketone (MIBK) was also applied with limited operation flexibility (Teidje, 1955).

The 'Barisol Process' (1930) which employed 78% ethylene dichloride and 28% benzene or chloroform was developed for centrifugal separation of oil solvent slurry on account of high specific gravity of the solvent used. However the refrigeration requirements were much higher resulting in increased cost of chilling.

Dewaxing by sulfur dioxide - benzene in 'Edeleanu solvent extraction process' (Nelson, 1958) was commercialised to eliminate double recovery of solvent as well as operational advantage of internal refrigeration with  $SO_2$  gas. A mixture of  $SO_2$  and benzene was used for aromatic extraction and wax separation in a single operation. The necessity for removal of solvent from waxy

raffinate after extraction stage could be avoided. The process was more suitable for extraction of aromatics from sharply defined cut than a broad fraction. The process became unpopular because of extensive corrosion and safety risk experienced with  $SO_2$  in presence of moisture.

Propane dewaxing process was developed and the first commercial unit was installed in 1932 by Standard Oil Co., Illinois (Gross and Gee, 1951). Selective solubility of petroleum wax in low molecular weight petroleum hydrocarbons e.g; liquid propane and its self refrigeration property made its suitable choice as dewaxing solvent.

### 1.3 MODERN DEWAXING PROCESSES :

A modern lube oil complex consists of following process units .

- A. Vacuum distillation unit
- B. Propane deasphalting unit
- C. Furfural / Phenol extraction unit
- D. Solvent dewaxing unit
- E. Hydrofinishing unit

A schematic flow diagram of the modern lube oil complex is given in Fig.1.1.

There are various processes developed for dewaxing of different petroleum fractions but solvent dewaxing process is commercially accepted by the refining industry for the manufacture of lube oil till now.

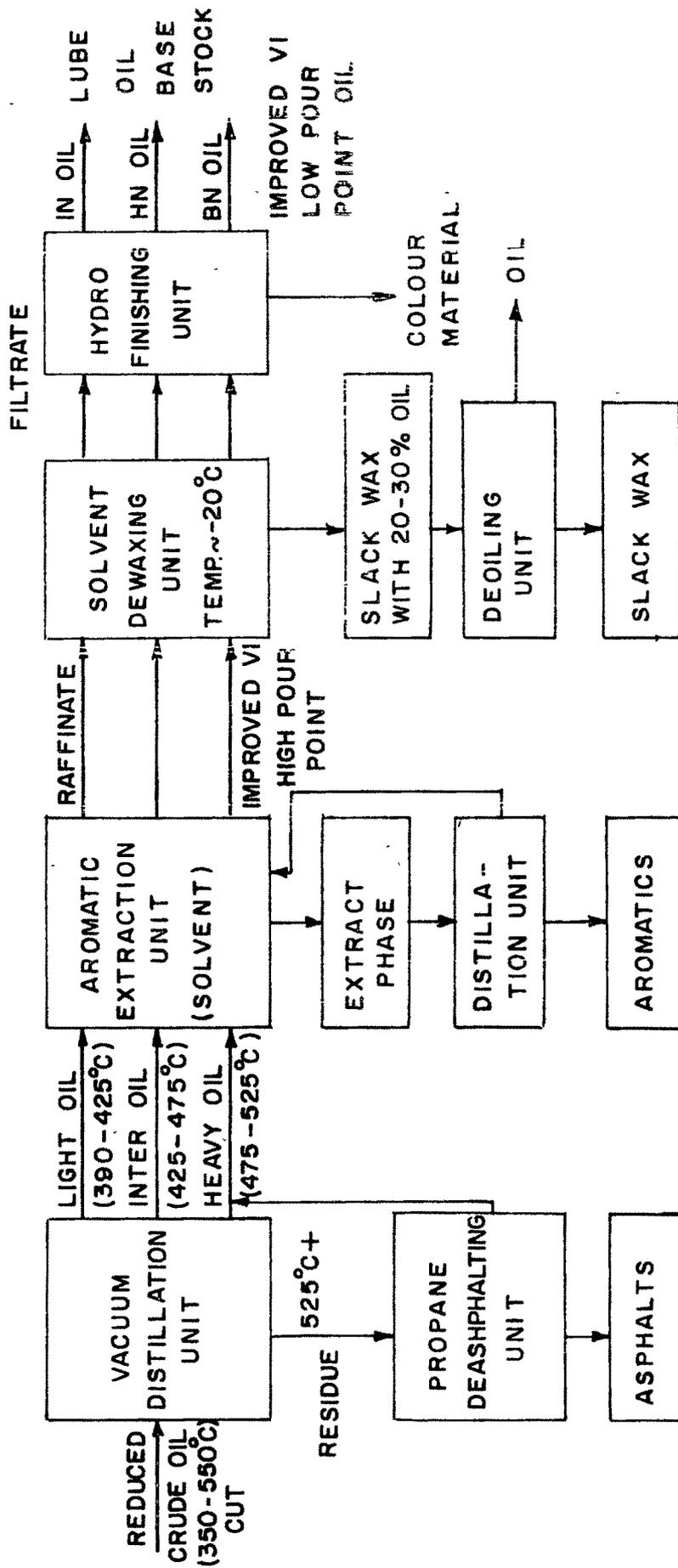


FIG. 1.1 - FLOW DIAGRAM OF LUBE OIL PROCESSING UNITS

Restricting the aspect of lube oil processing towards the dewaxing step, a modern solvent based dewaxing unit use either propane or methyl ethyl ketone and toluene as dewaxing solvent. However methyl isobutyl ketone and N-methyl pyrrolidine (NMP) are also found suitable in laboratory scale. The process parameters, the relative advantages and disadvantages of using different solvents in solvent dewaxing process had been discussed in detail by Nelson, 1959; Marpel and Landry, 1965; Berridge, 1973; Sukhanov, 1982.

Nearly 50% of the dewaxing units in U.S (Refining handbook 1984) and two out of three dewaxing units in India, at present producing lube oil are based on M E K dewaxing process while rest of the units are following either propane dewaxing process or catalytic dewaxing process.

Operating parameters of both the processes and their relative advantages and disadvantages are discussed in brief in the following section.

### 1.3.1 PROPANE DEWAXING

Propane dewaxing process has been generally accepted as a versatile and economical technique due to the following advantages.

- 1) The usage of common solvent for deasphalting as well as dewaxing is economically beneficial in solvent procurement and recovery system.
- 2) The solvent is cheap, non corrosive and readily available in refineries.

- 3) The direct and internal evaporative chilling eliminates the problem of wax adhesion to heat exchanger surface during chilling operation.
- 4) Propane is miscible in all proportion with oil.
- 5) High filtration rate.

The brief description of the process includes the following steps as shown in fig.1.2.

The feed and propane mixture is preheated to 45-50°C under pressure for complete dissolution of wax. Direct chilling of the solution by evaporation of propane to crystallise wax @ .4kg/kg of wax crystallised. Chilling is controlled @ 3 °C/min . Additional liquid propane is added during chilling to control slurry viscosity. The final temperature was maintained at -30 to -40 °C . Pressure filtration at 2-15 psi is generally carried out with rotary filters. Washing of cake is done with fresh propane. Evaporation of propane is carried out in two stages ,initially by flash evaporation and finally by steam stripping for complete removal of propane from the oil.

The disadvantages of propane dewaxing are as following :

1. High temperature differential is needed for a given pour point of the dewaxed oil, compared to dewaxing with ketonic solvents.
2. Batch chilling process is inconvenient to control uniformity of product except by electronic control systems .

3. High compression capacity is required for recompression of propane vapour to allow fast chilling at low propane vapour pressure.
4. The process is suitable for dewaxing of residual heavier petroleum fraction ( $525^{\circ}\text{C} +$ ) only.

### 1.3.2 MEK DEWAXING PROCESS

MEK dewaxing process is suitable for dewaxing the entire range of lube oil fraction for the removal of waxy paraffinic hydrocarbon from the raffinate obtained from solvent extraction unit. The dual solvent system, Methyl ethyl ketone (MEK) and toluene is used in the proportion of 2:3. Each component of the solvent has a specific function. The ketonic part (MEK) facilitates wax to crystallise while toluene increases capacity of the solvent for dissolving oil. (Gruse and Stevens, 1942; Kalichevsky and Stagner, 1942). The solvent has the following advantages over other type of solvents.

1. Low freezing point of MEK ( $-86.9^{\circ}\text{C}$ ) and toluene ( $-9.5^{\circ}\text{C}$ ) compared to acetone ( $-95^{\circ}\text{C}$ ) and benzene ( $0.5^{\circ}\text{C}$ ).
2. Very low solubility for waxes and oil i.e high selectivity in MEK and high solubility for oil and partly for waxes i.e; high solvent power in toluene.
3. Completely miscible with each other at all working temperatures and compositions.
4. Solvent power and selectivity can be adjusted by changing the ratio between the solvents.

Process description :

A typical flow diagram of the process is shown in Fig.1.3. involving the following steps.

- a) Chilling section - Waxy oil feed is mixed with solvent and heated to a temperature 35-45°C for complete dissolution of the wax. The solvent and oil mixture is then chilled in stages to crystallize the wax. Chilling is accomplished in scraped surface double-pipe heat exchanger first by exchange with cold product and then with a refrigerant @ 1.5 °C per minute. The final temperature after chilling is maintained around -20°C .
  
- b) Filtration section . The chilled mixture is then filtered continuously on rotary vacuum filter to separate crystallised wax from the oil solvent solution. Wax separated on the filter is in the form of thin cake ,washed with chilled solvent to displace retained oil. The filter cake is transferred by screw conveyors to wax boots for slack wax recovery. Filters require periodic washing with hot solvent to remove materials that plug the cloths . Sometimes, the plant is instrumented to take a filter off stream and to wash it automatically whenever the charge rate to it drops below a preset value.
  
- c) Solvent recovery section - The dewaxed oil/solvent mixture is heated and sent to four flash towers in series, the last one is used as steam stripper. The evaporated solvent is condensed and recycled.

D) Inert gas section - The inert gas is required for filter operation and as an inert gas blanket on the solvent vessels to avoid formation of explosive mixture as well as to protect the solvent against oxidation.

A typical operating conditions of MEK dewaxing process are given in TABLE 1.4.

TABLE 1.4 . OPERATING PARAMETERS OF MEK DEWAXING PROCESS

Operating parameters		Values
1. Solvent composition	MEK-TOLUENE	40-60
2. Solvent to oil ratio	wt%	
a) Before treatment		0.5 : 1
b) Adding solvent in three batches		
at 20°C -		0.5 : 1
at 10°C -		2.0 : 1
at 0°C -		2.0 : 1
Total dilution -		5.0 : 1
3. Wash solvent		1.0 : 1
4. Thermal treatment temperature	°C	45
5. Dewaxing temperature	°C	-15 to -20
6. Cake thickness	m m	4-5
7. Chilling rate	°C/h	
from 20 to 10 °C		80
from 10 to -20 °C		120
8. Rate of filtration on charge	Kg/ m <sup>2</sup> /h	120

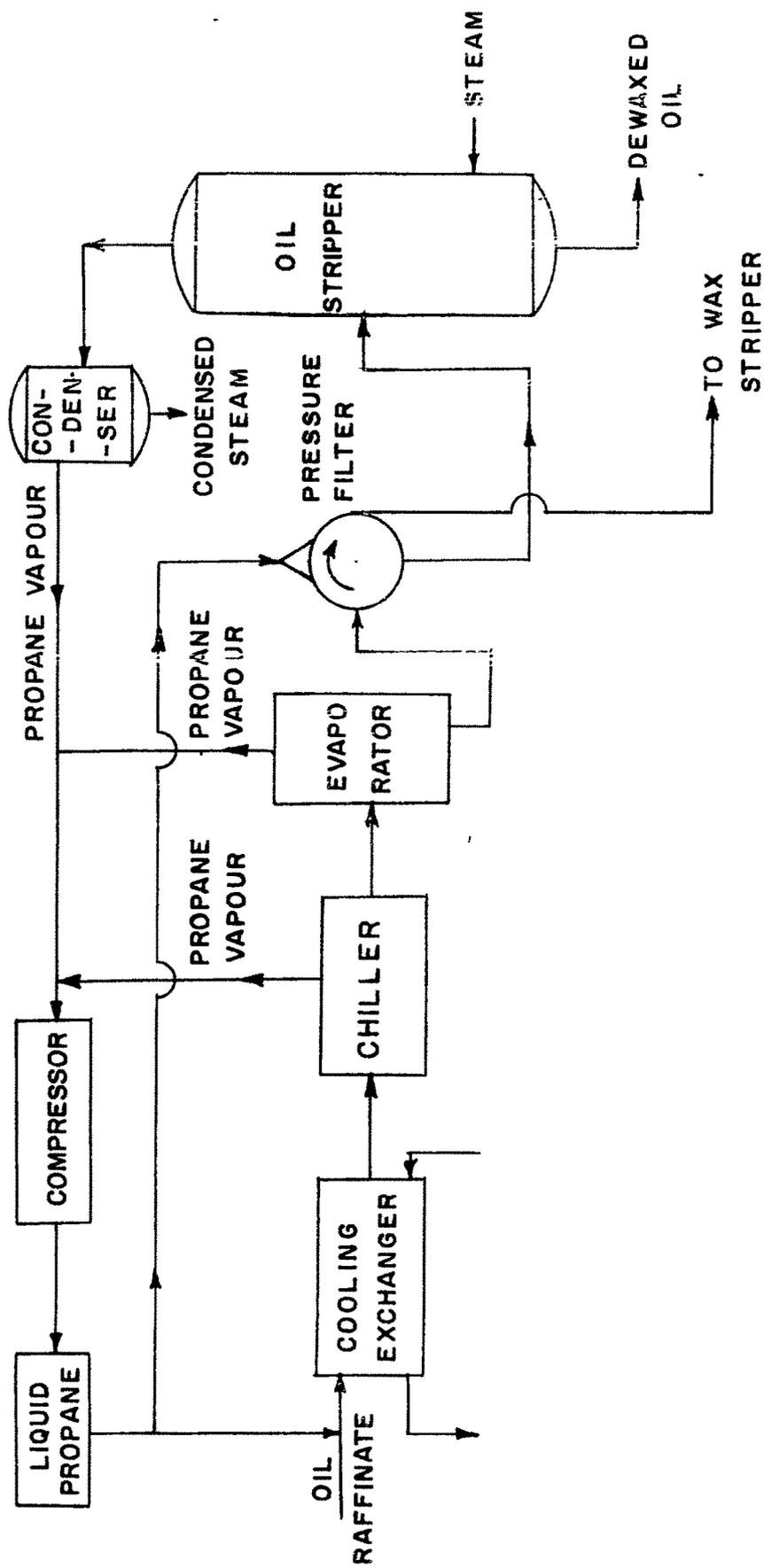


FIG. 1.2 - SCHEMATIC FLOW DIAGRAM OF PROPANE DEWAXING UNIT

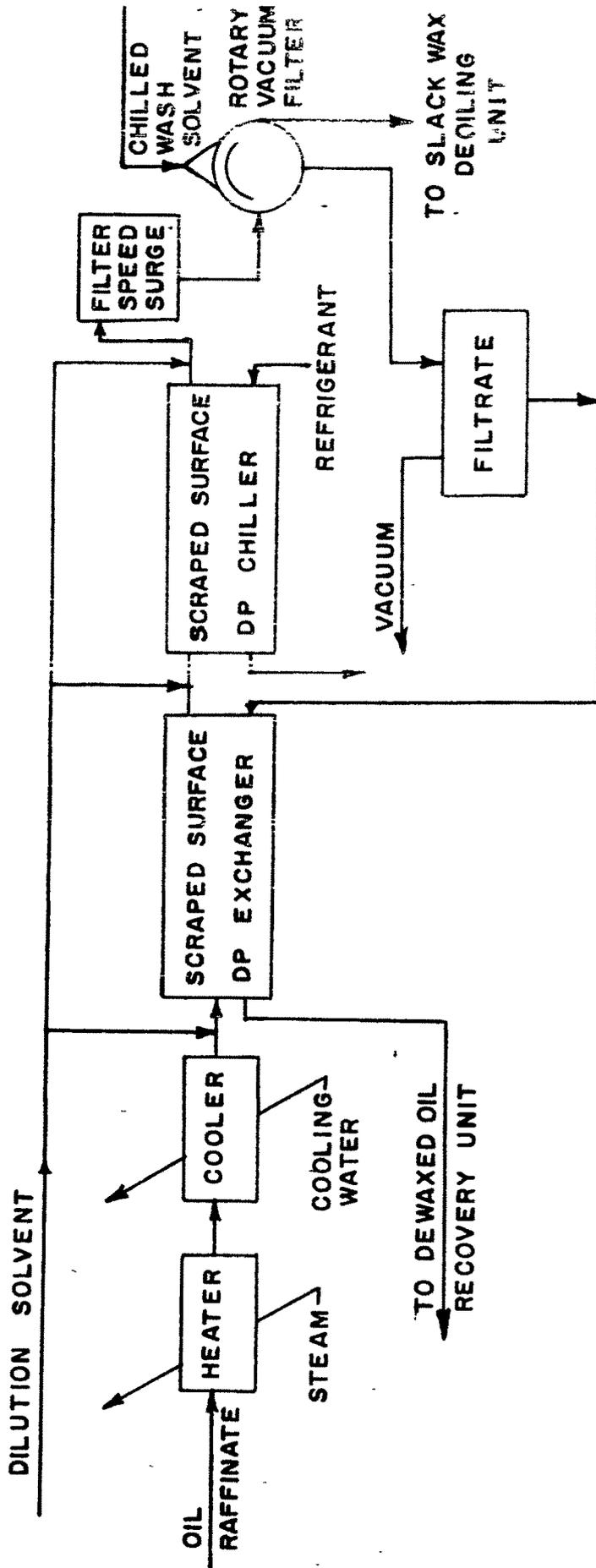


FIG. 1-3 - SCHEMATIC FLOW DIAGRAM OF MEK DEWAXING UNIT

#### LIMITATIONS OF OPERATING PARAMETERS OF THE MEK PROCESS :

1). Solvent composition - The desirable features of a dewaxing solvent is best achieved by the use of dual solvent consisting of aromatic component (Toluene) and a ketone (MEK). The solvent composition is needed to be adjusted to contain maximum MEK concentration so that highest filtration rate as well as lower dewaxing temperature differential can be obtained for saving the refrigeration cost (Reeves and Pattilo, 1948). On the contrary the high MEK concentration induces separation of oil phase with the wax at the filtration temperature. Such a condition produces high filtration rate but low yield of dewaxed oil. The immiscibility temperature for a given solvent composition at a given solvent /oil ratio is much higher for heavier cut compared to lighter distillate stock.

Hence a unit designed to process heavier stock would not be suitable to process lighter distillate unless provision is made to change the solvent composition as desired and vice versa.

2) Solvent dilution and its temperature - The technique of solvent addition to the oil feed affects crystal size which in turns control filtration rate (Table 1.5).

3) Size and shape of the crystal - The suitable size and shape of the wax crystals depend on the following process variables (Nelson, 1958 ; Marple and Landry, 1965 ; Aleksandrova, 1989 ).

TABLE 15. EFFECT OF SOLVENT DILUTION ON FILTRATION RATE

Solvent /oil	Crystal size	Effect on filtration
2	small crystals	poor <u>filtration</u> .
4	medium crystals	high <u>filtration</u> rate and oil yield.
6	fluffy crystals	high <u>filtration</u> rate but poor oil yield.

- 1) Feed composition.
- 2) Boiling range of feed.
- 3) Rate of cooling.
- 4) Rate of agitation.
- 5) Viscosity of crystallizing medium.
- 6) Dependence of solvent / oil dilution on feed viscosity for a given solvent viscosity . High viscous feed requires higher dilution ratio than lighter feed stocks

4) Rate of chilling - The crystal formation and subsequent filtration rate are influenced by the rate of cooling. A slow chilling rate produces large crystals which are easier to filter.

5) Filtration operation - The important operating variables of rotary filters are listed below :

- 1) Filter area immersed
- 2) Drum speed
- 3) Pressure differential for filtration and wash
- 4) Cake thickness.

In order to achieve the desired yield and quality of the dewaxed oil the above operating conditions are essentially maintained in its optimum value which are designed for a particular type of feed stock. It becomes necessary to redesign the operating parameters for any compositional change of feed stock.

5) Operation cost - Solvent dewaxing is the most expensive step in lube oil manufacturing process because the present technology is manpower intensive and consumes large amount of energy for refrigeration requirement to achieve the desired pour point (Farrel, 1986).

6) Removal of desirable components - Composition of waxes present in the vacuum distillate is the function of average carbon number of the petroleum fraction (Ramage, 1986). Hence some desirable branched and cyclic paraffins get separated by chilling operation, which otherwise impart higher viscosity index to the lube oil (Table 1.6)

TABLE 1.6. ROLE OF HYDROCARBON TYPE IN FLOW PROPERTIES OF LUBE OIL

Hydrocarbon type	Melting point °C	Viscosity index	Remarks
<u>n-paraffins</u>			
n- C <sub>26</sub> H <sub>54</sub>	56	194	Unsuitable
<u>Branched paraffins</u>			
$  \begin{array}{c}  \text{C}_8 - \text{C} - \text{C}_8 \\    \\  \text{C}_8  \end{array}  $	-13.8	116	suitable
<u>Cyclic paraffins</u>			
	-45	135	suitable
<u>Aromatics</u>			
	217	-60	unsuitable

Recent development of the solvent dewaxing process which is aimed at minimising the cost of operation and to increasing the yield of dewaxed oil, has been summarised in Refining process handbook , 1984.

a) DILCHILL dewaxing process- The process was developed by Exxon Research and Engg. Co, USA by employing a unique method of wax crystallisation by direct injection of cold dilution solvent into the warm waxy feed in a highly agitated crystallizer. The typical solvent to feed ratio is maintained between 2.5 to 6.0 depending

upon the viscosity level and other feed characteristics. A high filtration rate and dewaxed oil yield is obtained due to formation of highly discrete and compact wax crystal agglomerates.

2) M P LUBE OIL REFINING - The process was developed by Texaco Development Corporation USA for upgradation of raw paraffinic lubricating oil by using the solvent, N-Methyl-2-pyrrolidone (NMP) having high selectivity for wax. Low solvent / oil ratio is used to achieve a given yield of high quality dewaxed oil.

### 1.3.3. OTHER DEWAXING METHODS :

A) Urea dewaxing process : The process is based on the principle that n-paraffinic hydrocarbons form solid complex adduct with urea when a mixture of hydrocarbons is treated in presence of a suitable activator such as acetone or methanol . The n-paraffinic hydrocarbons of more than 5 carbon atoms that have little or no branching selectively get occluded inside the hexagonal channel (4-7<sup>o</sup>A dia ) formed by crystallizing 0.7 mole of urea for each carbon atom (Fetterley, 1957; Rogers et al, 1957; ).

Petroleum fraction, urea and the activator were mixed in the ratio 1:1.2:3 (vol :wt:vol ) and the mixture was stirred thoroughly for one hour at a temperature of 30 <sup>o</sup>C. The urea adduct was then filtered and the residual hydrocarbons were extracted with solvents e.g; toluene, benzene or naphtha .The urea aduct was decomposed in presence of the solvent or water at temperature 60-80<sup>o</sup>C. The vapours evolved were condensed and sent back to the vessel. Paraffinic hydrocarbons were recovered after distilling the solvent from the filtrate.

The process was used to dewax (a) gas oil, (b) light spindle oil to produce low pour point winter grade diesel oil and transformer oil (Hermann, 1965 ; Kuliev et al. 1989; Abdellaev et al , 1990).

The urea adducting process is suitable for dewaxing lighter stocks (250-400 °C) because it contains a higher percentage of straight chain paraffins. Dewaxing of lube fractions is not suitable by this process because adduct formation with the long chain paraffins becomes unstable and branched chain paraffins of high melting point are not accessible for adduct formation. So the decrease of pour point of the fraction are sometimes not up to the desired level (Gopalan ,1955).

However , several processes based on urea adduction have been developed for the separation of n-paraffinic hydrocarbons from kerosene and diesel fraction (Sista,1989) and industrial plants have been set up based on these processes. (Branken and Richter, 1979). Consumption of utilities by urea adduction process for the dewaxing of spindle oil and gas oil was reported by Hermann ,1965.

B) Molecular sieve adsorption process : The process is based on the principle of preferential adsorption of n-paraffinic hydrocarbons on molecular sieves i.e; zeolites . Petroleum fractions of boiling range 200-470°C can be dewaxed partially or fully to get the desired flow properties of the oil. A purge compound is used for desorbing the wax from molecular sieve under high vacuum. The process known as Molex process (Broughten ,1968) is under operation in commercial scale (IPCL, Baroda ). Union

Carbide has developed Isosieve process for separation of iso / n-paraffins from naphtha (Symoniak, 1980).

The process is not suitable for the petroleum fractions containing sulphur as the molecular sieves are contaminated by sulphur and their adsorption capacity is reduced. The process is also restricted to carbon numbers up to 20 of the hydrocarbons present in the fraction.

C).Catalytic dewaxing : Recent development in dewaxing technology is the catalytic conversion of paraffinic hydrocarbon to lower alkanes in presence of hydrogen and the process is being accepted as an efficient alternative to the solvent dewaxing process (Refining process handbook, 1984; Zakarian <sup>et al</sup> and farrell, 1987). Several petroleum processing industries e.g; Mobil, B.P, and Chevron have commercialized catalytic lube oil dewaxing process (Ramage et al, 1986; Wise et al, 1986 ;Bill and Marmin, 1979. ).

The principle of the process involves adsorption of n-paraffins in the precisely structured pores of synthetic zeolites ( ZSM-5) and hydrocracking of straight chain and slightly branched paraffins to lower fractions e.g.; gasoline and LPG as by product. The reactor effluent is cooled and flushed, the recycle gas being treated for removal of trace impurities. Separator liquid is flushed at low pressure and fractionated to give the dewaxed product. (Sivasanker et al, 1990).

The operating temperature is a function of pour point depression in the lube base stock and varies between 275-300°C at

start of run conditions. Reactor temperature also requires to be raised at regular intervals to compensate for catalytic deactivation. The hydrogen partial pressure at reactor outlet is about  $28 \text{ kg/cm}^2$ . The consumption of utilities was reported by Hargrove et al, 1979.

Although the catalytic dewaxing process is quite comparable, if not better than solvent dewaxing process with respect to yield and quality of lube oil, yet there are some limitations as given below :

- Deactivation of catalyst by cake formation
- High energy intensive process.

D) Microbial dewaxing : Several types of microorganism e.g.; bacteria, yeast and fungi are known to grow on petroleum fraction selectively assimilating n-paraffinic hydrocarbons (Shennan and Levi, 1974; Davis, 1975; Gutnick and Rosenberg, 1977). Literature indicated that considerable progress has been made towards the development of Single Cell Proteins (SCP) using middle distillate petroleum fractions as feed stock and also detailed study on biodegradation of hydrocarbons are in progress (Laine and du Chaffaut, 1975 ; Moo Young, 1976 ; Litchfield, 1977; Kappeli and Flechter, 1980;  Rhem and Reiff, 1981; Leahy and Colwell, 1990).

Assimilation of paraffinic hydrocarbons present in gas oil fraction by yeast was found advantageous for the following reasons:

- 1) Production of SCP
- 2) Dewaxing of gas oil

Based on this concept British Petroleum has developed a process for production of SCP and simultaneous dewaxing of gas oil (300-380 °C) using as feed stock for the 16,000 t/annum plant at Lavera refinery , France (Laine, 1974 )

Although the process contributed in producing dewaxed oil which was reused by the refinery ,much attention had been paid towards the process development ,yield and acceptability of SCP produced from gas oil. n Paraffins separated from the gas oil was substituted as feed stock for SCP production (Laine,1974).

Hence the choice of petroleum fraction as feed stock for either SCP production or dewaxing has been restricted to middle distillate (250-380 °C) and use of heavier petroleum fraction (350-500 °C ) for this purpose has not yet been reported .

The application of microbial dewaxing technique on lubricating oil fraction (b.p. 350-500°C) would be worth exploiting because of the simple operation and low cost involvment. The biomass which would be obtained as byproduct can be utilised as SINGLE CELL PROTEINS (SCP) in the cattlefeed.

## OBJECT OF THE STUDY

It has been established that industrial production of alcohol, organic acids, antibiotics, etc., by microbial processes have replaced the corresponding chemical processes due to their ambient operating conditions, lower energy input, high selectivity for the substrate present in the feed stock and desired product formation.

The progress made on process development for production of SCP from hydrocarbons as well as dewaxing of gas oil by yeast strains has encouraged us to study the aspects of microbial dewaxing of lube fraction and SCP production. The biotechnological process of dewaxing the heavier fraction is expected to be advantageous over solvent and catalytic dewaxing process for the production of low cost dewaxed oil due to the following reasons.

- 1) The growth temperature of hydrocarbon utilising yeast is near ambient ( $34 \pm 4$  °C) as compared to solvent dewaxing ( $-20$  °C) and catalytic dewaxing ( $300$  °C and  $27$  kg/cm<sup>2</sup> pressure).
- 2) Selective assimilation of paraffinic wax by growing yeast in the lube fraction may give better yield and flow properties.
- 3) Operational flexibility for the use of feed stocks obtained from different crudes.

The biomass (SCP) which would be obtained by the process can be used as protein supplement to the cattlefeed.

## THE SCOPE OF THE PRESENT STUDY

1. Selection and adaptation of a suitable yeast strain to grow on heavier petroleum fraction .
2. Studies on optimisation of growth parameters e.g.; temperature, pH, substrate concentration .
3. Batch and semicontinuous studies at optimised conditions to compare the the sp. growth rate, yield of biomass, yield of dewaxed oil and rate of pour point decrease.
4. Scale- up studies of the fermentation process with respect to oxygen transfer rate and power per unit volume . Design of a pilot scale fermentor (1 M<sup>3</sup> cap.) based on the scale up data.
5. Recovery of dewaxed oil from the fermented broth.
6. Analysis of dewaxed oil and SCP obtained by the process.