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(54) **Recovery for solvent from a hydrocarbon extract.**

(57) An energy saving process for recovery of solvent from a solvent-oil mixture in a staged series of vaporization zones at progressively increasing pressure in which external heat is supplied to the vaporization stage having the highest pressure and heat for each preceding vaporization stage is supplied by heat exchange with vapors from succeeding vaporization stages with vaporization at the lowest pressure stage carried out in a plurality of vaporization zones at substantially equal pressure.

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RECOVERY OF SOLVENT FROM A HYDROCARBON EXTRACT

The invention relates to an improved process for the recovery of solvent employed in processing a petroleum oil fraction containing constituents having different physical and chemical properties. In one of its more specific aspects, the invention relates to a method for recovering solvent from hydrocarbon extract in a lubricating oil solvent refining process utilizing N-methyl-2-pyrrolidone as a solvent.

The process of this invention is related to the processes disclosed in U.S. Patent Applications, Serial Nos. 377,293 and 377,294, both filed on 12th May, 1982, and co-pending Patent Application No. , (Ref. D.77,220-PBS-1-FB), which claims the priority of those U.S. Applications.

It is well known that aromatic and unsaturated components of a hydrocarbon oil charge stock may be separated from the more saturated hydrocarbon components by various processes involving solvent extraction of the aromatic and unsaturated hydrocarbons. Suitable solvents have an affinity for at least one component of the hydrocarbon oil charge stock and are partially immiscible with the charge stock under the temperature and pressure conditions employed in the solvent extraction step. Two liquid phases are present in the extraction zone; the two

liquid phases generally consist essentially of an extract phase containing the major amount of the solvent together with dissolved aromatic components of the charge stock and a raffinate phase containing non-aromatic components of the charge stock together with a minor amount of solvent. Among the solvents which are known to be useful for solvent extraction processing of petroleum base lubricating oil stocks are furfural, N-methyl-2-pyrrolidone, phenols, and other various well known organic and inorganic solvents. The removal of aromatics and other undesirable constituents from lubricating oil base stocks improves the viscosity index, color, oxidative stability, thermal stability, and inhibition response of the base oils and the lubricating oil products produced from hydrocarbon feedstocks.

Most recently N-methyl-2-pyrrolidone has displaced furfural and phenol in importance as a preferred solvent for extracting aromatic hydrocarbons from mixtures of aromatic and non-aromatic hydrocarbons. Some of the advantages of N-methyl-2-pyrrolidone as solvent are referred to, for example, in U.S. Patent 4,057,491. N-methyl-2-pyrrolidone is effective for the solvent extraction of aromatic components from lubricating oil charge stocks at relatively lower temperatures and lower solvent-to-oil dosages than most other known solvents. N-methyl-2-pyrrolidone is generally the most preferred solvent because of its chemical stability, low toxicity, and its ability to produce refined oils of improved quality. Some of the prior art processes employing N-methyl-2-pyrrolidone as solvent and illustrating conventional solvent recovery operations are disclosed in U.S. Patents 3,461,066 and 3,470,089.

The process of this invention is useful for upgrading existing N-methyl-2-pyrrolidone refining installations employing a single or multiple stage solvent recovery system and steam or inert gas stripping of the solvent from the products. The process of this invention is also particularly suited to the conversion of furfural and phenol process installations to N-methyl-2-pyrrolidone solvent systems with substantial savings in the energy requirements of the solvent refining process.

In recovering N-methyl-2-pyrrolidone from oil-solvent mixtures, e.g., the extract phase and the raffinate phase of a solvent refining system wherein solvent is separated from oil-solvent mixtures by a combination of distillation and stripping. Stripping with an inert gas rather than with steam for solvent purification often reduces the energy requirements of the process, as compared with conventional steam stripping. Inert gas stripping has been disclosed, for example, in U.S. 2,923,680; 4,013,549 and 4,057,491.

In conventional lubricating oil solvent refining processes, the solvent extraction step is usually carried out in a countercurrent extraction tower under conditions effective for recovery of about 30 to 90 volume percent of the lubricating oil charge as raffinate or refined oil and to extraction of about 10 to 70 volume percent of the charge as an aromatic extract. The lubricating oil stock is contacted with a solvent, e.g., N-methyl-2-pyrrolidone, at a temperature at least 5°C, preferably at least 50°C, below the temperature of complete miscibility of said lubricating oil stock in the solvent.

In the extraction step, operating conditions are selected to produce a primary raffinate having a dewaxed viscosity index of about 75 to 100, and preferably about 85 to 96. Solvent extraction temperatures within the range of 43 to 100°C (110 to 212°F), preferably within the range of 54 to 95°C (130 to 205°F), and solvent dosages within the range of 50 to 500 volume percent, basis hydrocarbon feedstock, and preferably within the range of 100 to 300 volume percent, are suitable. Extraction pressure at the solvent to raffinate interface is preferably 1.4 bar to 2 bar. Water or wet solvent may be injected into the extractor near the point of withdrawal of the extract and solvent mixture to control solvent power and selectivity.

To produce a finished lubricating oil base stock, the primary raffinate is dewaxed to the desired pour point. If desired, the refined or dewaxed oil may be subjected to a finishing treatment for color and stability improvement, for example, mild hydrogenation. Dewaxing processes employing various solvents are also well known in the prior art. Solvent recovery in accordance with the process of this invention may be applied to such solvent dewaxing processes.

The operation of the extraction tower involves counterflow of the two immiscible liquid phases. Therefore, the mechanical feasibility of the solvent refining process depends on a significant density difference between the solvent-rich phase, or extract phase, and the oil-rich phase, or raffinate phase. Within the solvent dosage range of 100 to 500 volume percent, i.e., 100 to 500 volumes of solvent to each 100 volumes of lubricating oil feedstock, the density difference increases with increased solvent dosage. At very low solvent dosages, for

example, less than 100 percent, the density difference can become so low as to severely limit the throughput of feed to the solvent extraction tower.

N-methyl-2-pyrrolidone is such an effective solvent for aromatics that in the case of some hydrocarbon charge stocks the solvent dosage needed to produce the desired raffinate quality is impractically low. When operating an extraction tower with dry N-methyl-2-pyrrolidone at the minimum practical dosage, i.e., about 100 percent, and temperature, i.e., about 60°C (140°F), the refined oil quality is higher than desired and in some cases the refined oil yield is lower than desired.

The extraction step may be operated with a dry solvent dosage effective for rapid separation of the two liquid phases within the extraction tower, and the extraction tower refluxed by the introduction of water or wet solvent into the tower near the point of withdrawal of the extract phase, i.e. between the point of introduction of the hydrocarbon feedstock to the separation system and the point of withdrawal of the extract phase, to obtain the desired quality raffinate product with a high yield of refined oil.

It has been proposed heretofore to add water to the N-methyl-2-pyrrolidone in the extraction tower either as such or in admixture with the solvent as a reflux to reduce the solubility of the aromatic hydrocarbons in the solvent. The present process employing N-methyl-2-pyrrolidone as solvent provides improvements in the methods of separating solvent from the extract and raffinate products, eliminating oil contamination in the solvent, and controlling the water content of the solvent in the solvent refining system. In a preferred solvent extraction process, dry solvent is used as the primary

solvent in the extraction tower and water or wet solvent is employed as a reflux whereby a high yield of refined oil of desired quality at a given solvent dosage is obtained.

The above mentioned U.S. Patent Applications, Serial Nos. 377,293 and 377,294, and co-pending Patent Application No. (Ref. D.77,220-PBS-1-FB) provide improved methods for recovery of solvent from mixtures including those comprising the extract phase obtained in solvent refining lubricating oil base stocks, wherein solvent is removed from a solvent and oil mixture by vaporization of solvent in a series of stages involving partial vaporization in a first low pressure vaporization zone and then vaporizing additional portions of the solvent from the mixture in a plurality of vaporization zones at progressively higher pressures with heat from an external source supplied only to the highest pressure vaporization zone and heat for each preceding vaporization zone supplied by heat exchange with vapors from each succeeding vaporization zone. Part of the vapors from the last or highest pressure vaporization zone may be mixed with vapors from a medium pressure vaporization zone as heat supply to the low pressure solvent/vaporization zone. In these systems, additional solvent is recovered from the extract and solvent mixture leaving the high pressure vaporization zone by vaporization in a subatmospheric pressure flash zone at a temperature higher than the temperature of the high pressure vaporization zone. Final traces of solvent are removed from the extract by stripping with inert gas.

In accordance with the present invention, a plurality of low pressure flash zones are employed for solvent recovery prior to flash vaporization in the medium and high pressure flash vaporization zones common to these systems.

The process of the invention will be more readily understood by reference to the accompanying drawing and the following detailed description of a preferred embodiment of the process.

The figure is a simplified flow diagram of a preferred embodiment of the process of the invention as applied to a lubricating oil solvent refining process employing N-methyl-2-pyrrolidone as solvent.

With reference to the figure, a petroleum base lubricating oil feedstock is supplied to the solvent refining process illustrated through line 1 and split into two streams. Part of the feedstock passes through line 2, heater 3 and line 4 to the upper part of absorber tower 5 wherein the lubricating oil feedstock is brought into intimate countercurrent contact with an inert stripping gas, e.g. nitrogen, containing solvent vapors entering the lower part of the absorber tower 5 through line 6. Absorber tower 5 comprises a countercurrent vapor-liquid contacting tower wherein liquid flowing down the tower is intimately contacted with gases and vapors passing upwardly through the tower. Means for ensuring intimate contact between vapor and liquid, e.g. bubble cap trays, perforated plates, packing material, or the like, are provided within the tower. A preferred embodiment of the process is illustrated and described as a specific example; in this example, the lubricating oil feedstock from line 2 is heated in heater 3 to a temperature of 66°C and absorber 5 is operated at 1.7 bar. In the absorber 5, solvent vapors are absorbed by the lubricating oil feedstock and the recovered solvent returned with the feedstock to the process. Stripping medium, from which solvent has been removed, is discharged through line 7 and heater 8 for reuse in the process.

A second portion of the lubricating oil feedstock from line 1 is passed through line 12, heater 13 and line 14 into the upper part of an absorber tower 15 wherein the lubricating oil feedstock is brought into intimate countercurrent contact with a mixture of steam and solvent vapors entering the lower part of absorber 15 through line 16. Absorber 15 comprises a countercurrent contacting tower similar to absorber 5 described above and, as a specific example, may be operated at a pressure of 1.1 bar and a temperature of 102 to 104°C. Steam from which solvent has been removed is discharged through line 17 to condenser 18 wherein the steam is condensed and the condensate accumulated in "rate" drums 19 where it is stored until tested for solvent content and, if sufficiently low, released to the sewer system.

The lubricating oil feedstock streams discharged from the lower part of absorbers 5 and 15 are combined and passed through line 22, cooler 23, and line 24 to the lower part of extraction tower 25 wherein the lubricating oil feedstock is intimately countercurrently contacted with dry N-methyl-2-pyrrolidone solvent introduced into the upper part of extraction tower 25 through line 26. As used herein, "dry" N-methyl-2-pyrrolidone means N-methyl-2-pyrrolidone containing 0.3 weight percent water or less. Extract tower 25 is preferably operated at an interface pressure of 1.4 to 2 bar; in a specific example, the interface pressure is 1.4 bar with a raffinate outlet temperature of 63°C and an extract outlet temperature of 46°C.

The raffinate mixture, comprising typically 85 percent hydrocarbon oil admixed with solvent is discharged from the extraction tower 25 through line 28 and processed for

the recovery of raffinate from the solvent. The raffinate, after separation of solvent, is a solvent refined lubricating oil base stock, the desired product of the process. The recovery of solvent from the raffinate is described hereinafter.

The major portion of the solvent is contained in the extract mixture withdrawn from the bottom of extraction tower 25. In this example, an extract mixture comprising about 85 percent solvent is withdrawn from tower 25 through line 31 and passed through heat exchangers 32, 33 and 34, which serve to preheat the mixture, into the first of a series of low pressure flash towers 35, 36, and 37 wherein water and part of the solvent are vaporized. Flash towers 35, 36 and 37 are provided with vapor-liquid contacting means, e.g. cascade trays, in their upper part to effect countercurrent contact between reflux liquid flowing down the tower and solvent vapors flowing up the tower. A part of the extract and solvent mixture from the bottom of tower 35, is cooled and introduced into the upper part of each of flash towers 35, 36 and 37 as reflux in known manner, not illustrated. Flash tower 35, 36 and 37 are preferably operated at approximately the same pressure, suitably in the range of 1.15 to 1.4 bar. In this specific example, the pressure in the flash towers are 1.15 bar, 1.15 bar, and 1.4 bar, respectively, and the flash tower temperatures are about 180°C, 186°C, and 211°C, respectively.

Solvent and extract mixture leaving flash tower 35 is further heated in heat exchanger 38 prior to introduction to flash tower 36.

Solvent vapors separated from the extract mixture in flash towers 35 and 36 contain considerable amounts of water vapor. The mixed solvent and water vapors from

flash towers 35 and 36 pass through line 39 to heat exchanger 32 where most of the solvent vapor and part of the water vapor are condensed, preheating the extract and solvent mixture from line 31. Condensate and uncondensed vapors pass through line 41 to accumulator 42 as part of the feed to drying tower 45 as described hereinafter.

The solvent and extract mixture leaving low pressure flash tower 36 is passed through heat exchanger 46 further heating the mixture prior to introduction to low pressure flash tower 37. The solvent vapor mixed with water vapor from flash tower 37 is passed through heat exchanger 34 to further preheat the extract and solvent mixture from line 31, condensing part of the water and solvent vapors. Condensate and uncondensed vapors are passed through line 41 to accumulator 42 as part of the feed to drying tower 45.

The remaining solvent and extract mixture from which part of the solvent has been removed by vaporization in flash towers 35, 36 and 37 is passed through heat exchangers 48 and 49 to medium pressure flash tower 50, similar in construction to flash towers 35, 36 and 37. The medium pressure flash tower 50 suitably is operated at a pressure in the range of 1.7 to 2.9 bar; in this specific example, the medium flash tower pressure is 2.07 bar and the flash tower temperature is 239°C. A minor portion of the extract solvent mixture from the bottom of flash tower 35 is introduced to the upper part of the flash tower 50 as reflux in known manner, not illustrated.

The solvent vapors leaving the top of medium pressure flash tower 50 are passed through line 51 to heat exchanger-condenser 46 in indirect heat exchange with the solvent and extract mixture from flash tower 36, and then to heat exchanger 33 to supply heat to the extract and solvent mixture in line 31. Condensed solvent vapors from heat exchanger 33 flow through line 52 to accumulator 92 as dry solvent for reuse as described hereinafter.

Extract solvent mixture from which a further part of the solvent has been removed by vaporization in medium pressure flash tower 50, is withdrawn from the lower part of flash tower 50 and passed through heater 54 where the mixture is heated to a temperature in the range of 288 to 310°C and introduced into high pressure flash tower 55 for the removal of most of the remaining solvent from the extract mixture. The high pressure flash tower 55 is similar in construction to flash towers 35, 36, 37 and 50 and suitably is operated at a pressure within the range of 2.9 to 3.14 bar, and in this specific example, at 2.9 bar. A minor portion of the extract and solvent mixture from the bottom of flash tower 35 is introduced to the upper part of the high pressure flash tower 55 as reflux in known manner, not illustrated.

Solvent vapors leaving the top of high pressure flash tower 55 pass through line 56 and heat exchanger 49 in indirect heat exchange with the extract and solvent mixture from low pressure flash tower 37, condensing the solvent vapors and supplying heat to the extract solvent mixture prior to its introduction to medium pressure flash tower 50. Solvent vapors are condensed in heat

exchanger 49 and the condensate passed through heat exchangers 38 and 33 to line 52 as part of the dry solvent supplied to extraction tower 25.

The hydrocarbon oil extract and solvent mixture withdrawn from the bottom of high pressure flash tower 55 is passed through heat exchanger 48, to supply heat to the solvent and extract mixture from flash tower 37 and then through expansion valve 58 to heater 60 and vacuum flash tower 65 for further recovery of solvent from the extract. The vacuum flash tower may operate at a pressure within the range of 0.25 to 0.55 bar, and at a temperature in the range of 293°C to 315°C; in this specific example the vacuum flash tower pressure is 0.45 bar and the operating temperature is 293°C. A minor portion of the extract solvent mixture from the bottom of flash tower 35 is supplied to the top of vacuum flash tower 60 as reflux in known manner, not illustrated.

In the vacuum flash tower 65, additional separation of extract from solvent takes place. Solvent vapors are withdrawn from the top of vacuum flash tower 65 through line 66 to a condenser 67 and solvent accumulator 68. Uncondensed gases are withdrawn from accumulator 68 through line 69 to a suitable vacuum source, not illustrated, and may be discharged from the system.

The hydrocarbon oil extract withdrawn from the bottom of the vacuum flash tower 65 still contains some solvent, for example, 7 volume percent solvent and 93 volume percent hydrocarbon extract. This extract mixture is introduced into the upper portion of extract stripping tower 71.

Extract stripping tower 71 is typically a countercurrent vapor-liquid contact column provided with bubble cap trays in which the liquid extract flowing downwardly through the column is contacted with inert stripping gas introduced into the lower portion of tower 71 through line 72. A part of the extract mixture from the bottom of stripping tower 71 is cooled and returned to the upper portion of the tower as reflux through line 73.

Extract oil containing less than about 50 parts per million solvent, and typically comprising 80 weight percent unsaturated hydrocarbons and about 20 percent saturated hydrocarbons, is withdrawn from the lower end of stripping tower 71, passed through heat exchanger 74 where it is cooled, and discharged from the system through line 75 as a product of the process.

Inert stripping gas, e.g. nitrogen, and stripped solvent vapors are discharged from the upper part of stripping tower 71 through line 76 to condenser 77 where solvent vapors are condensed. Solvent condensate is collected in condensate accumulator 78 and returned through line 79 to dry solvent storage 92 for recycle to extraction tower 25. Inert gas separated from the condensate solvent in separator 78 is recirculated by compressor 80 to line 6 and absorber 5 for the recovery of trace amounts of solvent contained in the recirculated stripping gas. In this example, extract stripping tower 71 is operated at a pressure just above atmospheric pressure, e.g., 1.1 bar to 1.3 bar and a temperature of 299°C. Condenser 77 cools the stripping gas and solvent to a temperature of the order of 60°C effecting condensation of the major part of the solvent from the nitrogen or other stripping gas prior to recycle to absorber 5. Absorber 5 recovers substantially all of the residual solvent from the recycle nitrogen stream.

Raffinate mixture taken overhead from extraction tower 25 through line 28 typically comprises about 15 volume percent solvent and 85 volume percent hydrocarbons. In this particular example, the extraction tower is operated with a dry solvent dosage of 100 volume percent, i.e. one volume of solvent for each volume of oil charge stock. In the specific example, raffinate mixture is discharged from the extraction tower at a temperature of 63°C. The raffinate mixture from line 28 is collected in run tank 82, and heated in heat exchanger 83 and in a fired heater 85 prior to introduction into vacuum flash tower 86 wherein solvent is separated from the raffinate mixture. In one preferred embodiment, raffinate vacuum flash tower 86 is operated at a pressure of 0.7 bar and a temperature of the order of 298°C. Reflux from a suitable source, e.g. dry N-methyl-2-pyrrolidone, is supplied to the top of vacuum flash tower 86 through line 87 as reflux.

In raffinate vacuum flash tower 86, separation of the major portion of the solvent from the raffinate takes place. Solvent vapors are withdrawn from the top of flash tower 86 through line 88; heat exchanger 83, and cooler 89 to solvent accumulator 90. Condensate solvent from accumulators 90 and 68 flow through line 79 to run tank 92 from which dry solvent is withdrawn through line 26 to extraction tower 25. Uncondensed gases are withdrawn from solvent accumulator 90 through line 93 to a suitable vacuum source, not illustrated, and may be discarded or further processed for the recovery of solvent vapors therefrom.

Raffinate, still containing some solvent, is withdrawn from the lower part of vacuum flash tower 86 through line 95 to the upper part of stripping tower 96, wherein

residual solvent is removed from the raffinate by stripping with inert gas. Inert gas from absorber 5 is introduced into the lower part of stripping tower 96 via lines 7 and 97. A minor portion of the raffinate from the raffinate cooler 98 is reintroduced to the upper part of the raffinate stripping tower 96 as reflux in known manner, not illustrated. In a preferred embodiment, raffinate stripping tower 96 is operated at a pressure just above atmospheric pressure, e.g. 1.1 bar to 1.3 bar and at a temperature of 288°C. Nitrogen containing solvent from stripper 96 is combined with nitrogen containing solvent from stripper 71 and cooled in condenser 77 for condensation of solvent from the stripping gas recirculated to absorber 3.

Raffinate, substantially free from solvent, is withdrawn as a product of the process from the lower portion of stripper 96 through heat exchanger 98 where it is cooled and discharged to line 100 as the refined lubricating oil stock, the principal product of the process.

The solvent purification system of this process comprises drying tower 45 where water vapor or steam mixed with solvent vapors from low pressure flash tower 35 and from medium pressure flash tower 48 are processed for the recovery of dry solvent for reuse in extraction tower 25. Solvent vapors containing water vapor or steam are passed from low pressure flash tower 35 through line 39 to heat exchanger 33 wherein the vapors are cooled and partially condensed by heat exchange with the extract mixture leaving the bottom of extraction tower 25 through line 31. The resulting vapor-liquid mixture comprising wet solvent, solvent vapors, and water vapor pass through line 41 to accumulator drum 42 wherein wet solvent (liquid) is separated from solvent vapors and steam.

From accumulator drum 42, wet solvent is introduced into drying tower 45 through line 101 and steam containing solvent vapors is introduced into drying tower 45 through line 102 wherein dry solvent is separated from steam and solvent vapors. Solvent vapors from medium pressure separator 48 containing water vapor are passed through line 49 to heat exchanger-condenser 34 wherein they are cooled and partially condensed by indirect heat exchange with extract mixture from line 31. In heat exchanger-condenser 34 the extract mixture is preheated prior to its introduction to low pressure flash tower 35 condensing a portion of the solvent vapors from line 49. The condensed solvent is essentially free from water vapor and is withdrawn from heat exchanger-condenser 34 through line 50 to line 49 and passed through line 106 to dry solvent accumulator 92. Uncondensed vapor from heat exchanger-condenser 34 is passed through line 51 to drying tower 45 for the recovery of solvent therefrom.

Drying tower 45 comprises a fractionating column provided with suitable means, for example, perforated plates or bubble cap trays, for ensuring intimate countercurrent contact between vapors rising upwardly through the column and liquid flowing downwardly therethrough. Drying tower 45 is provided with a reboiler 103 at the bottom of the fractionating column to vaporize all of the water and part of the solvent entering the drying tower with the various feed streams. Dry N-methyl-2-pyrrolidone is withdrawn from the bottom of drying tower 45 through line 104 and passed through heat exchanger 33, and line 52 to dry solvent accumulator 92 as dry solvent for extraction tower 25. In this specific example, drying tower 45 is operated at a pressure of 1.08 bar with a bottom temperature, i.e. reboiler temperature, of (216°C) and a tower top temperature of (104°C to 132°C).

Part of the steam and accompanying solvent vapors taken overhead from drying tower 45 pass through line 108 and is cooled and condensed in condenser 109. Condensate water containing a small amount of solvent is accumulated in water drum 110 from which part of the water is returned through line 111 to the top of drying tower 45 as reflux and part is passed through line 27 to extraction tower 25 as a solvent modifier or reflux for the extraction tower. The remaining part of the overhead vapor from drying tower 45 comprising steam containing a minor amount of N-methyl-2-pyrrolidone is passed through line 16 to absorber tower 15 where it is brought into intimate countercurrent contact with a portion of the feed from line 14 recovering the solvent from the steam.

In solvent refining systems, such as the one described herein, water almost inevitably enters the system with the lubricating oil feedstock so that even in a dry solvent extraction system, means must be provided for the removal of extraneous water from the system. Other sources of water contamination in a system such as the one described herein occur from leaks in heaters or heat exchangers employing steam or water as a heat exchange medium. Excess water is eliminated in the process of this invention by passing the excess water in the form of steam through line 16 to absorber tower 15 for trace solvent removal before condensation in condenser 18 and collection of the reject water in rate drum 19.

It will be evident to one skilled in the art that the process of this invention permits substantially complete recovery of solvent from solvent and hydrocarbon mixtures encountered in dewaxing, solvent refining and similar hydrocarbon processing operations and effects a savings in energy as compared with conventional solvent recovery methods. It is to be understood that, while the

process is described in detail herein with reference to a lubricating oil solvent refining process employing N-methyl-2-pyrrolidone as solvent, the process is also applicable to other solvent refining processes, solvent dewaxing processes, and other processes in which a lower boiling solvent or mixture of solvents is to be separated from a higher boiling liquid.

CLAIMS

1. A process for recovery of volatilizable material from a mixture comprising said volatilizable material and a less volatile material wherein said volatilizable material is removed from said mixture by vaporization in a plurality of flash vaporization zones serially at progressively increasing pressure levels and wherein heat from an external source is supplied to said last high pressure vaporization zone and heat for each preceding vaporization zone is supplied by heat exchange with vapors from succeeding vaporization zones, characterized by subjecting said mixture to a plurality of stages of flash vaporization at said first pressure level effecting removal of a portion of said volatilizable material from said mixture and thereafter subjecting the resulting mixture to vaporization in a plurality of flash vaporization zones at successively higher pressure levels.

2. A process according to Claim 1 characterized in that said mixture comprises the extract phase from a lubricating oil solvent refining process employing N-methyl-2-pyrrolidone as solvent.

3. A process for the recovery of solvent from a solvent and hydrocarbon mixture wherein said solvent is lower boiling than said hydrocarbon and wherein solvent is separated from said hydrocarbon in a plurality of flash vaporization zones at progressively increasing pressure levels, characterized by subjecting said solvent and hydrocarbon mixture to a plurality of stages of flash vaporization at said first pressure level affecting removal of a portion of said solvent from said mixture and thereafter subjecting the resulting solvent and extract mixture

to flash vaporization in a plurality of flash vaporization zones pressure at progressively higher levels.

4. A process according to Claim 3, characterized in that said mixture comprises paraffinic hydrocarbons.

5. A process according to Claim 3, characterized in that said mixture comprises aromatic hydrocarbons.

6. A process according to Claim 3, characterized in that said mixture comprises naphthenic hydrocarbons.

7. A process for solvent refining a lubricating oil feedstock wherein said lubricating oil feedstock is contacted under pressure with N-methyl-2-pyrrolidone as a selective solvent for aromatic constituents of said feedstock in an extraction zone under solvent refining conditions thereby forming a raffinate phase comprising a minor amount of said solvent and an extract phase comprising a major amount of said solvent, said raffinate phase is separated from said extract phase, and said solvent is removed from the solvent and extract mixture comprising said extract phase by flash vaporization serially in a plurality of flash vaporization zones at progressively increasing pressure levels the first of which is at a pressure lower than that of said extraction zone, and wherein heat from an external source is supplied to said high pressure vaporization zone and heat for each preceding vaporization zone is supplied by heat exchange with vapors from a succeeding vaporization zone, characterized by subjecting said solvent and extract mixture to a plurality of stages of flash vaporization at said first pressure level effecting removal of a portion of said solvent from said mixture and thereafter subjecting the resulting solvent and extract mixture to flash vaporization in a plurality of flash vaporization zones at progressively higher pressure levels.

8. A process according to Claim 7, characterized in that said flash vaporization zones at said first pressure level comprises at least two flash vaporization zones operating at substantially equal pressure.

9. A process according to Claim 7 or 8, characterized in that the pressure of said first pressure level flash vaporization zones is in the range of 1.15 to 1.4 bar.

10. A process according to Claim 9, characterized in that the pressures of said succeeding pressure levels are in the range of 1.7 to 2.9 bar and 2 to 7 bar, respectively.

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