ENGINEERING STANDARD

FOR

PROCESS DESIGN OF LPG RECOVERY

AND SPLITTER UNITS

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0. INTRODUCTION

The purpose of LPG Unit is to process light hydrocarbons (C_1-C_5) into required component streams. The LPG Unit feed streams in a refinery are usually obtained from three refinery Units consisting of Atmospheric Crude Distillation, Platformer and Hydrocracker Units. However, any LPG stream which is economically justified for recovery of the components shall be routed to the LPG recovery Unit.

"Design of Non-Licensed Process Units" are broad and contain various subjects of paramount importance. Therefore, a group of process engineering standards are prepared to cover this subject.

This group includes the following:

STANDARD CODE	STANDARD TITLE
IPS-E-PR-360	"Process Design of Liquid and Gas Transfer and Storage"
IPS-E-PR-491	"Process Requirements of Refinery Non-Licensed Units"
IPS-E-PR-500	"Process Design of LPG Recovery & Splitter Units"
IPS-E-PR-551	"Process Design of Gas Treating Units"

This Engineering Standard Specification covers:

"PROCESS DESIGN OF LPG RECOVERY & SPLITTER UNITS"

1. SCOPE

This Engineering Standard Specification covers minimum process design requirements for LPG recovery & splitter Units. It should be expressed that only general process requirements are covered in this Standard and the Unit specific design conditions shall be determined based on the feed analysis and final product specifications during execution of the Unit conceptual design.

2. REFERENCES

Throughout this Standard the following standards and codes are referred to. The editions of these standards and codes that are in effect at the time of publication of this Standard shall, to the extent specified herein, form a part of this Standard. The applicability of changes in standards and codes that occur after the date of this Standard shall be mutually agreed upon by the Company and the Vendor/Consultant.

ASTM (AMERICAN SOCIETY FOR TESTING OF MATERIALS)

ASTM-D-2163
ASTM-D-2420

IPS (IRANIAN PETROLEUM STANDARDS)

IPS-G-ME-150	"Towers, Reactors, Pressure Vessels & Internals"
IPS-E-PR-360	"Process Design of Liquid and Gas Transfer and Storage"
IPS-E-PR-491	"Process Requirements of Refinery Non-Licenced Units"
IPS-E-PR-850	"Process Requirements of Vessels, Reactors and Separators"

GPSA (GAS PROCESSORS SUPPLIERS ASSOCIATION)

"Engineering Data Book", Vol. 2, Section 17-26, 1987

IP (INSTITUTE OF PETROLEUM, LONDON)

IP 10 (A) July, 1985 "Petroleum Measurement Manual", Part 10, "Meter Proving"

3. DEFINITIONS AND TERMINOLOGY

3.1 Absorption

Is a process which the liquid (absorbent) flows countercurrent to a gas stream for the purpose of removing one or more constituents (absorbate) from that gas.

3.2 Lean Oil (Absorbent)

Is usually those hydrocarbons which have molecular mass of about 180-200 and maximum final boiling point of 160°C (e.g., Naphtha).

3.3 Liquefied Petroleum Gas (LP-Gas or LPG)

Any material having a vapor pressure not exceeding that allowed for commercial propane composed predominantly of the following hydrocarbons, either by themselves or as a mixtures: propane, propylene, butane (normal butane or iso-butane) and butylene, as a by-product in petroleum refining or natural gasoline manufacture.

3.4 Reid Vapor Pressure (RVP)

Is the Pressure of the vapor in equilibrium with liquid at 37.8°C (100°F).

4. SYMBOLS AND ABBREVIATIONS

APC Advanced Process Control.
LPG Liquefied Petroleum Gas.
PFD Process Flow Diagram.
RVP Reid Vapor Pressure.

5. UNITS

This Standard is based on International System of Units (SI), except where otherwise specified.

6. BASIC DESIGN REQUIREMENTS

6.1 Process configuration of the LPG recovery Units shall be established based on the following factors and submitted for Company's approval:

- feed composition;
- upstream Unit process configurations;
- ultimate product consumption;
- product specifications;
- minimum C_3, C_4 and C_5 (if any) recovery.

For typical process flow diagrams of LPG recovery unit in a refinery, reference can be made to the arrangements shown in Drawing No. E-05-4-001 of Esfahan Refinery (Typical Process Flow Diagram for LPG Unit), and Drawing No. 05-0001 of Arak Refinery (Typical Process Flow Diagram for LPG Unit).

6.2 Minimum recovery of LPG components shall be as follows. However, economical study shall be practiced to justify provision of the absorption/stripping lean oil system if required by the minimum product recovery specification as instructed by the project scope of the work. The final process configuration shall be approved by the Company.

- Minimum C_3 recovery: 92 (vol. %).
- Minimum C_4 recovery: 98 (vol. %).
- Minimum C_5 recovery: 99.5 (vol. %).

6.3 The Unit product minimum specifications shall be as follows, unless otherwise specified in the project specification. The specifications of finished LPG product shall be as outlined in Appendix A of this Standard.

	C_3	C_4	C_5
- C_2 (vol. %)	0.3 (max.)	—	—
- C_3 (vol. %)	Balance	0.2 (max.)	—
- C_4 (vol. %)	3.5 (max.)	Balance	1 (max.)
- C_5 (vol. %)	—	1 (max.)	Balance
H_2S (ASTM D-2420)	Negative	Negative	

6.4 If required by the feed compositions, and product specifications, treating facilities shall be provided.

6.5 The Unit design throughput shall be based on the sum of maximum flowrates of various feed streams to the Unit when the upstream Units are operating at their design capacities.

6.6 C_3 and C_4 products must be manufactured separately and each stream must be suitable for LPG blending as per LPG specification shown in Appendix A.

6.7 Drying facilities for C_3 product shall be provided.

6.8 Feed surge drum shall be provided to receive all feed streams into the LPG recovery Unit.

6.9 Unless otherwise specified, the Unit turndown capacity shall be 60% of design throughput, without loss of efficiency in fractionation while meeting the product specifications.

6.10 Where the fin fan coolers are utilized, spare for addition of one bay for future expansion shall be provided.

6.11 The Unit design capacity shall be determined based on the Licenser's information on the upstream Licensed Process Units and shall take into consideration the variations resulted in the relevant process Units.

6.12 One cooler shall be supplied to cool the feed gas from the crude distillation overhead compressor (if any).

6.13 General equipment design criteria shall be according to the specification setforth in Article 6.2 of IPS-E-PR-491, "Process Requirements of Refinery Non-Licensed Units".

6.14 Reference shall be made to IPS-E-PR-360, "Process Design of Liquid and Gas Transfer and Storage", for LPG storage and handling.

6.15 Process design requirements and equipment design oversizing factors shall be according to Article 6.3 of IPS-E-PR-491.

6.16 The following design notes shall be taken into consideration if LPG caustic treating section to be supplied as per feed and product specifications.

- Caustic dissolving facilities shall be included if supply of the caustic outside of the Unit battery limit is not feasible.

- Caustic regeneration facilities shall be provided, if economically are justified.

- Spent caustic degassing and storage shall be provided.

6.17 Safety considerations shall be fully complied.

6.18 Special attention shall be made to the flexibility and ease of operations, equipment interchangeability and optimization.

6.19 Maximum energy conservation shall be applied.

6.20 Kettle type reboilers shall be provided to maintain the bottom temperature of the following towers:

- deethanizer;
- depropanizer;
- propane dryer;
- debutanizer.

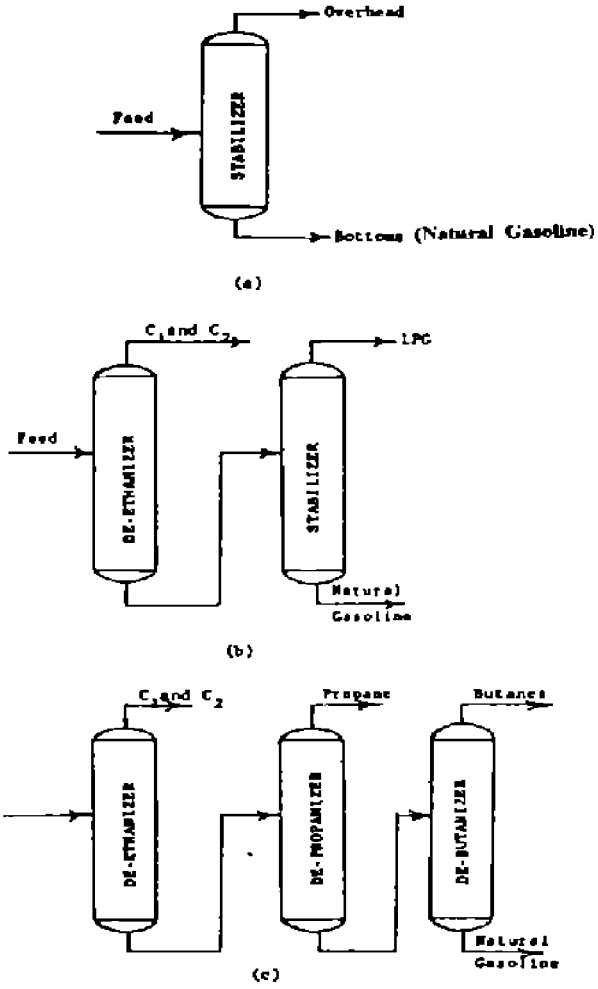
7. FRACTIONATION AND SYSTEM CONFIGURATION

7.1 General

7.1.1 Stabilization tower shall be used where a natural gasoline or stable liquid to be produced [see Fig. 1(a)].

7.1.2 The two tower system shown in Fig. 1(b) is most commonly used to produce an LPG mixture in the overhead and a natural gasoline product as the bottoms. In this system, the deethanizer must remove all methane, ethane and other constituents not suitable in the two product streams from the second tower. Any material that enters the second tower must necessarily leave in one of the product streams.

7.1.3 The three tower system shown in Fig. 1(c) most commonly produces commercial propane, commercial butane and natural gasoline as products. In this system also, the deethanizer must work properly to remove all constituents that cannot be sold in one of the three products. The sequence of fractionation following the deethanizer may be varied. In the second tower, an LPG mixture could be produced overhead with natural gasoline produced as bottoms. The third tower would then split the LPG into commercial propane overhead and commercial butane as bottoms. This sequence is favored sometimes where the market situation is variable and a market for LPG only exists during a portion of the year. During this period, the third tower would be shut down and not operated.



TYPICAL LPG FRACTIONATION SYSTEMS
Fig. 1

7.1.4 Regardless of how the fluids are removed from natural gas and/or gasoline, fractionation is necessary if products that meet any kind of rigid specification are to be made. The number of fractionating columns required depends on the number of products to be made and the character of the liquid which serves as feed. The single tower system shown in Fig. 1 (a) ordinarily produces one specification product from the bottom stream, with all other components in the feed passing overhead.

7.1.5 The operating pressure of a fractionation tower is ordinarily fixed by the condensing temperature of the overhead product. Temperature in the condenser is ordinarily controlled by the cooling medium. Allowing for sufficient temperature difference between the cooling medium and the overhead product, condenser temperature is fixed by the designer. In the case of a liquid distillates, the bubble point pressure is then calculated; for a vapor distillate product the dew point

pressure would be calculated. This pressure is the minimum pressure at which the tower can operate at the chosen condenser temperature. For vapor pressure chart, reference could be made to Appendix B (for the source, reference is made to GPSA in Section 2).

7.1.6 Economic investigation shall be made for selection of a total and partial condenser for a tower. At a given pressure, the dew point is always a higher temperature than the bubble point, and this tends to minimize cooling costs, where all other elements are equal.

7.2 Fractionation Design Considerations

7.2.1 If the tower involved is the first one in a fractionation system, the conditions of the feed to the column will be fixed by the separation process. A surge vessel prior to this tower might change the analysis of the feed if some vapor is withdrawn at that point. (An example of this is a rich oil flash drum situated between the absorber and the deethanizer). For a system containing several towers, the split desired in each column should be made before completing analysis of any one. This assures that the splits set up for the different towers produce satisfactory products in all streams. A perfect separation between adjacent components cannot be specified. This will lead to a situation impossible to achieve in an actual column. In producing propane, for example, it must be allowed that a small amount of ethane and butanes be present. In this case, however, the propane must still meet the purity specifications demanded.

7.2.2 There are three ways in which the desired operation of a fractionation column is ordinarily specified:

- A specified percentage recovery of one component in the distillate or one component in the bottom.
- A specified composition of one component in the distillate or bottom stream.
- A specified vapor pressure for either the distillate or bottom product.

7.2.3 The design of a fractionator shall incorporate the following considerations:

- feed composition, quantity, temperature and pressure;
- desired products and their specifications;
- a reasonable condensing temperature for the overhead stream;
- degree of separation between products.

7.2.4 Overall tray efficiency for the most fractionators is normally about 50-80%. (see Table 1).

TABLE 1 - TYPICAL FRACTIONATOR PARAMETERS*

DESCRIPTION	OPERATING PRESSURE kPa (ga)	NUMBER OF ACTUAL TRAYS	REFLUX [1] RATIO	REFLUX [2]	TRAY EFFICIENCY (%)
Demethanizer	1400 - 2800	18 - 26	Top feed	Top feed	45 - 60
Deethanizer	2600 - 3100	25 - 35	0.9 - 2.0	0.6 - 1.0	50 - 70
Depropanizer	1700 - 1900	30 - 40	1.8 - 3.5	0.9 - 1.1	80 - 90
Debutanizer	500 - 620	25 - 35	1.2 - 1.5	0.8 - 0.9	85 - 95
Butane splitter	550 - 700	60 - 80	6.0 - 14.0	3.0 - 3.5	90 - 110
Rich oil fractionator (still)	900 - 1100	20 - 30	1.75 - 2.0	0.35 - 0.40	Top 67, bottom 50
Rich oil deethanizer	1400 - 1750	40	—	—	Top 25 - 40, bottom 40 - 60
Condensate stabilizer	700 - 2800	16 - 24	Top feed	Top feed	40 - 60

1 reflux ratio relative to overhead product, (mol/mol)

2 reflux ratio relative to feed, dm^3/dm^3 (gal/gal)

***Note:**

For source of these figures see GPSA in Clause 2.

8. ABSORPTION/STRIPPING

8.1 Basic Requirements

8.1.1 The Absorption/Stripping Process may be used in LPG recovery Unit if needed by the minimum recovery specification of the product streams as requested in the project scope of the work.

8.1.2 The design should incorporate both stripper and absorber towers with all associated facilities. Typical Absorption/Stripping Process is referred to drawing no. E-05-4-001 of Esfahan Refinery.

8.1.3 The resultant rich oil shall be stripped or denuded of the absorbed materials in the stripper tower. The stripped oil shall be recirculated to the Absorber tower as lean oil for Absorption of the LPG components.

8.2 Design Considerations

8.2.1 A temperature rise of 6°C - 16°C is usually designed into initial condition for the Absorption Process. The rise above this must be handled by intercoolers.

8.2.2 The pressure range for the design of Deethanizer/Absorber tower is normally in the range of 800 - 14500 kPa (ga) or 8 - 14.5 bar (ga) (see Table 2).

8.2.3 Overall tray efficiency for the most absorbing systems is normally about 25 - 40% as shown in Table 2 below:

TABLE 2 - ABSORPTION STRIPPING APPROXIMATE TRAY EFFICIENCIES

TYPE	PRESSURE RANGE, [kPa (ga)]	TEMP. (°C)	RANGE EFFICIENCY (%)
Absorption			
Hydrocarbon oils & vapors	0 - 5600	-1 to 54	35 - 50
Propane-key,	800 - 14500	—	30 (37*) - 38
Butane-key	800 - 14500		28 (33*) - 36
Stripping			
Hydrocarbon oils with steam	0 - 1000	150 - 300	50 - 80
Unsaturates in oil with closed reboiler	0 - 450		25 - 35

*** Average Value**

9. CONTROL AND OPTIMIZATION

9.1 Control Scheme

Control scheme design shall generally be according to IPS-E-PR-491, Appendix A.

9.2 Reboiler Control

Control system of the reboilers shall be investigated for proper functioning based on the tower feed compositions. For typical reboiler control systems, reference can be made to the arrangements shown in Drawing No. 05-0001 of Arak Refinery (Typical Process Flow Diagram for LPG Recovery Unit). However, due to the characteristics of the deethanizer tower feed and possibility of drastic changes in the feed composition, the control system of the deethanizer tower

reboiler shall be in accordance with the arrangement shown in Drawing No. E-05-4-001 of Esfahan Refinery (Typical Process Flow Diagram for LPG Unit) for deethanizer tower (V-503).

9.3 Advanced Process Control and Optimization

9.3.1 Advanced Process Control (APC) and optimization shall be applied to upgrade plant safety, product quality and quantity and plant operation. The extent of application shall be as per Company's instructions.

9.3.2 The following APC loops shall be incorporated in the process design as minimum requirement. The APC loops shall preferably be functioned without implementation of on line process analyzer.

a) Deethanizer tower:

- Control system for matching specification of C_2 at bottom and LPG recovery at top.

b) Depropanizer and Debutanizer towers:

- C_3 quality control.
- C_4 quality control.

APPENDICES

APPENDIX A
LPG SPECIFICATION

	SPECIFICATION		TEST METHODS
C_2 Hydrocarbon	% vol.	NIL	ASTM D 2163
C_3 Hydrocarbon	% vol.	*	ASTM D 2163
C_4 Hydrocarbon	% vol.	*	ASTM D 2163
C_5 Hydrocarbon	% vol.	2 max.	ASTM D 2163
Hydrogen Sulphide		Negative	ASTM D 2420
Mercaptan Sulphur	mg/dm^3	0.0288 max. (1)	IP 10(A)
Odorizing Agent	g/m^3	12	

1) The limit applies to the product before addition of odorizing agent (Ethyl mercaptan).

*** Varies seasonally for refineries as follows:**

Name of Refineries		1st Khordad* to 1st Mehr	1st Mehr* to 1st Azar	1st Azar* to 1st Esfand	1st Esfand* to 1st Khordad
Abadan Refinery	C_3, vol.%	15 - 25	30 - 40	50 - 60	30 - 40
	C_4, vol.%	85 - 75	70 - 60	50 - 40	70 - 60

Name of Refineries		1st Khordad* to 1st Shahrivar	1st Shahrivar* to 1st Aban	1st Aban* to 1st Farvardin	1st Farvardin* to 1st Khordad
Tehran Mahshahr Shiraz Kermanshah Esfahan	C_3, vol.%	15 - 25	30 - 40	50 - 60	30 - 40
	C_4, vol.%	85 - 75	70 - 60	50 - 40	70 - 60

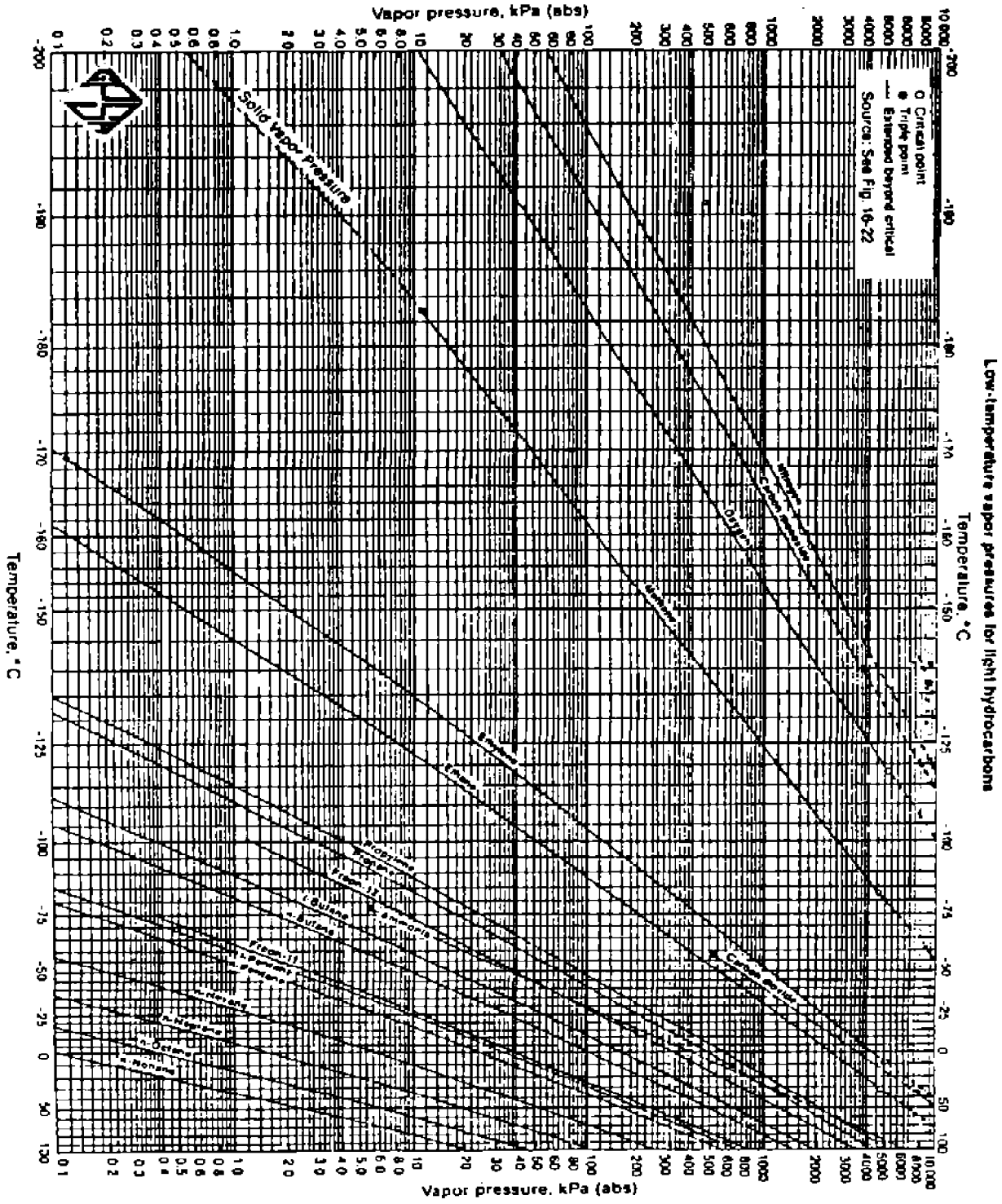
Name of Refineries		1st Tir* to 1st Shahrivar	1st Shahrivar* to 1st Aban	1st Aban* to 1st Ordibehesht	1st Ordibehesht* to 1st Tir
Tabriz Refinery	C_3, vol.%	15 - 25	30 - 40	50 - 60	30 - 40
	C_4, vol.%	85 - 75	70 - 60	50 - 40	70 - 60

The latest issues of the relevant test methods shall be used.

*** Note:**

1st Khordad to 1st Mehr corresponds to 21st May to 22nd September.
1st Mehr to 1st Azar corresponds to 22nd September to 21st November.
1st Azar to 1st Esfand corresponds to 21st November to 19th February.
1st Esfand to 1st Khordad corresponds to 19th February to 21st May.
1st Khordad to 1st Shahrivar corresponds to 21st May to 22nd August.
1st Shahrivar to 1st Aban corresponds to 22nd August to 22nd October.
1st Aban to 1st Farvardin corresponds to 22nd October to 20th March.
1st Farvardin to 1st Khordad corresponds to 20th March to 21st May.
1st Tir to 1st Shahrivar corresponds to 21st June to 22nd August.
1st Shahrivar to 1st Aban corresponds to 22nd August to 22nd October.
1st Aban to 1st Ordibehesht corresponds to 22nd October to 20th April.
1st Ordibehesht to 1st Tir corresponds to 20th April to 21st June.

APPENDIX B
VAPOR PRESSURE CHART



(to be continued)

APPENDIX B (continued)