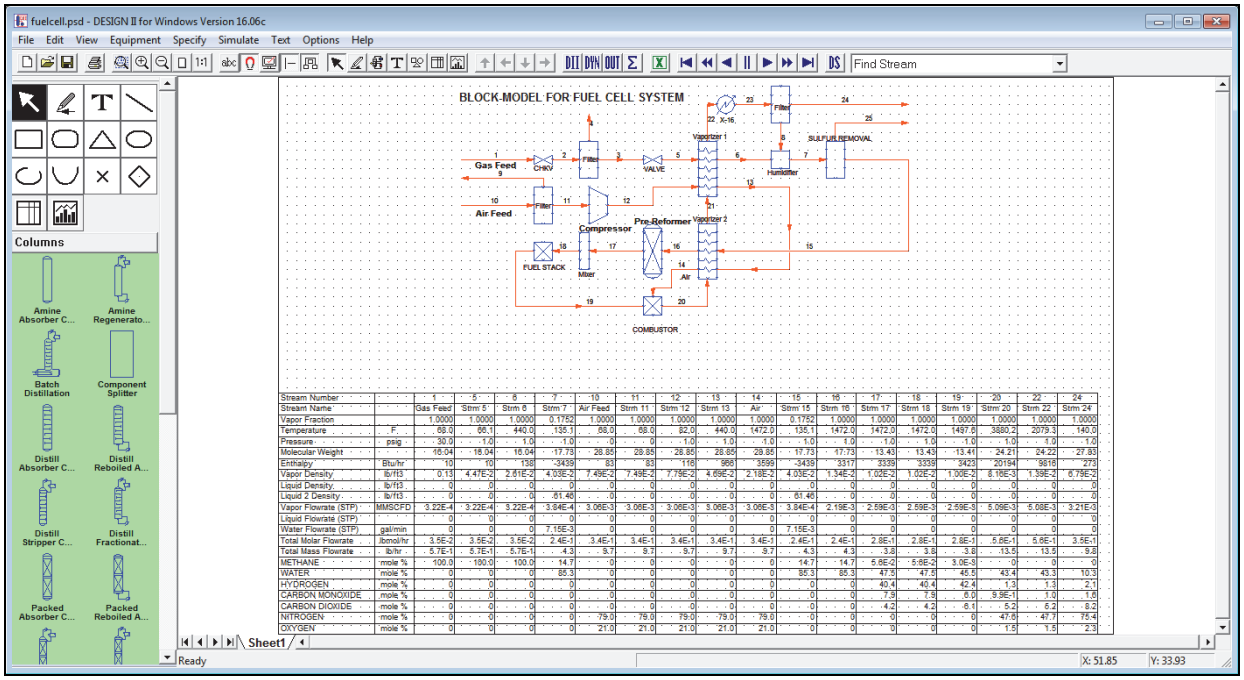
DESIGN II for Windows™

Tutorial and Samples

Version 16.0



WinSim® Inc.

Advanced Engineering Software

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TABLE OF CONTENTS

Chapter 1: Tutorial

The Flowsheet Example

In this chapter, we will walk you through the basic steps necessary to simulate process flowsheets such as the following:

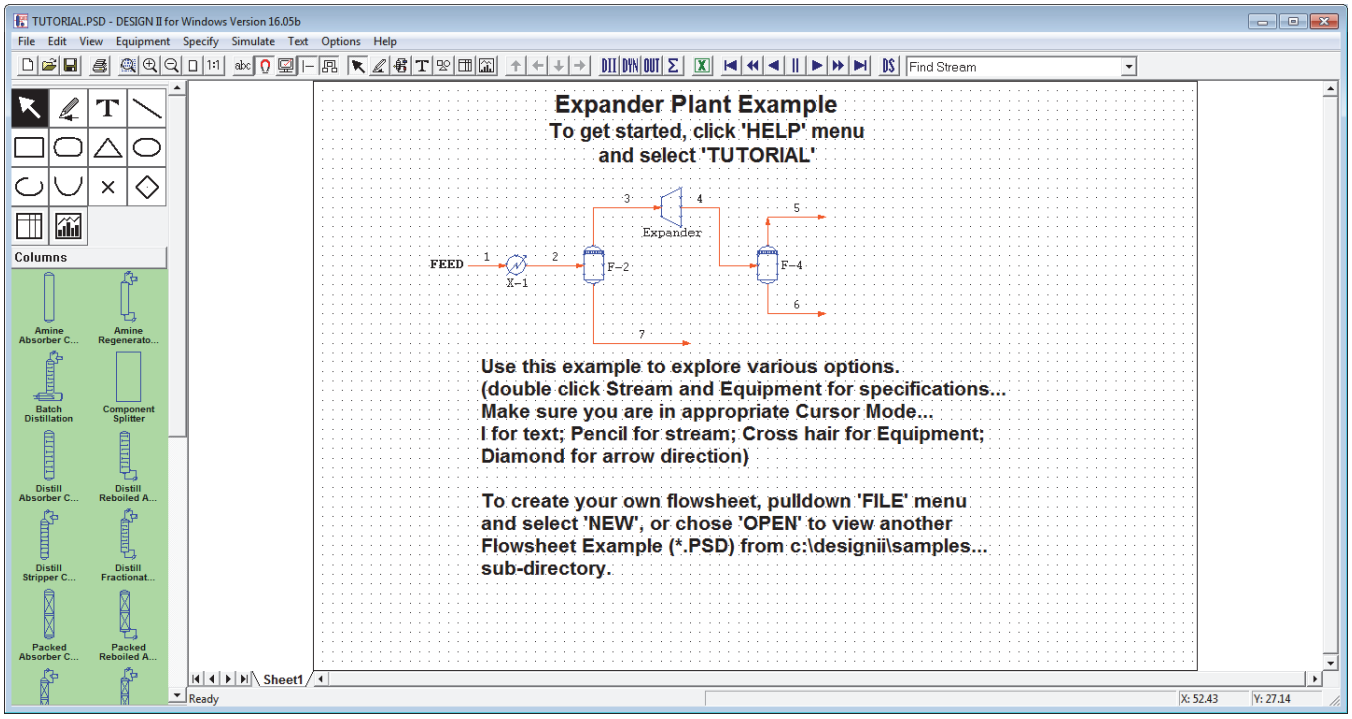


Figure 1 - 1: Sample Process Simulation (tutorial.psd)

What You'll Learn In This Tutorial

We show you how to accomplish these tasks in this chapter:

1. Start the DESIGN II for Windows program.
2. Draw the Process Simulation Diagram (PSD) for an Expander plant.
3. Specify Stream Components, Feed Stream Conditions, Equipment, and Thermodynamic Methods.
4. View the Input File has been created for DESIGN II.
5. Calculate the flowsheet.
6. View the results.

This flowsheet is a single stage expansion of a light hydrocarbon stream. First, the feed stream is cooled through a feed chiller (X-1) and any liquids are removed with a vertical separator (F-2). The vapor is expanded (E-4) and liquids separated using a second vertical separator (F-3). The expander process uses the following equipment:

Equipment	Tag	Specifications
Feed Chiller	X-1	Temp Out = -35° F, Delta Pressure = 10 psi Adiabatic Flash, Delta Pressure = 0
Two Vertical Separators	F-2 F-3	
Expander	E-4	Pressure Out = 275 psig Efficiency = .80

Feed Stream Specification:

Component	Lbmol/Hr	Specifications
Methane	6100	Temperature = 90° F
Ethane	500	Pressure = 980 psig
Propane	200	

N-Butane	100
N-Pentane	100
N-Hexane	70

Thermodynamics methods used
K-Value = API Soave
Enthalpy = API Soave
Density = Yen Woods (STDD)

Start Up DESIGN II for Windows

To start the program: Double click on the **DESIGN II for Windows** icon in the **Main** window of the Program Manager. The DESIGN II for Windows screen displays. The new drawing dialog will show up on the window. Select US mode and then select paper size C, then click OK.

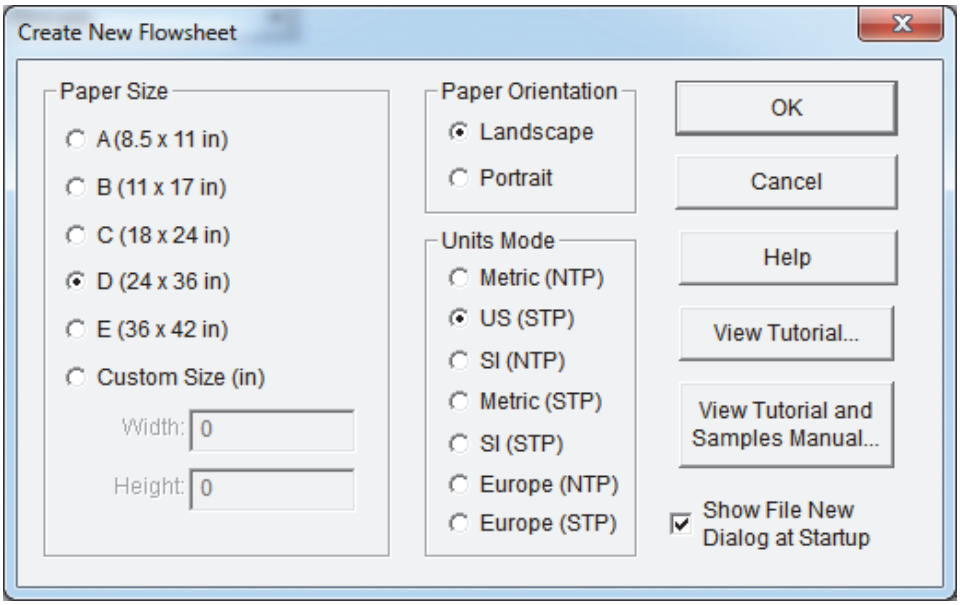


Figure 1 - 2: DESIGN II for Windows Screen

NOTE: To exit the program, open the File menu and select Exit. If you have a drawing open, you will be prompted to save it if you have modified the drawing.

You will now have a blank window on which to draw your flow sheet.

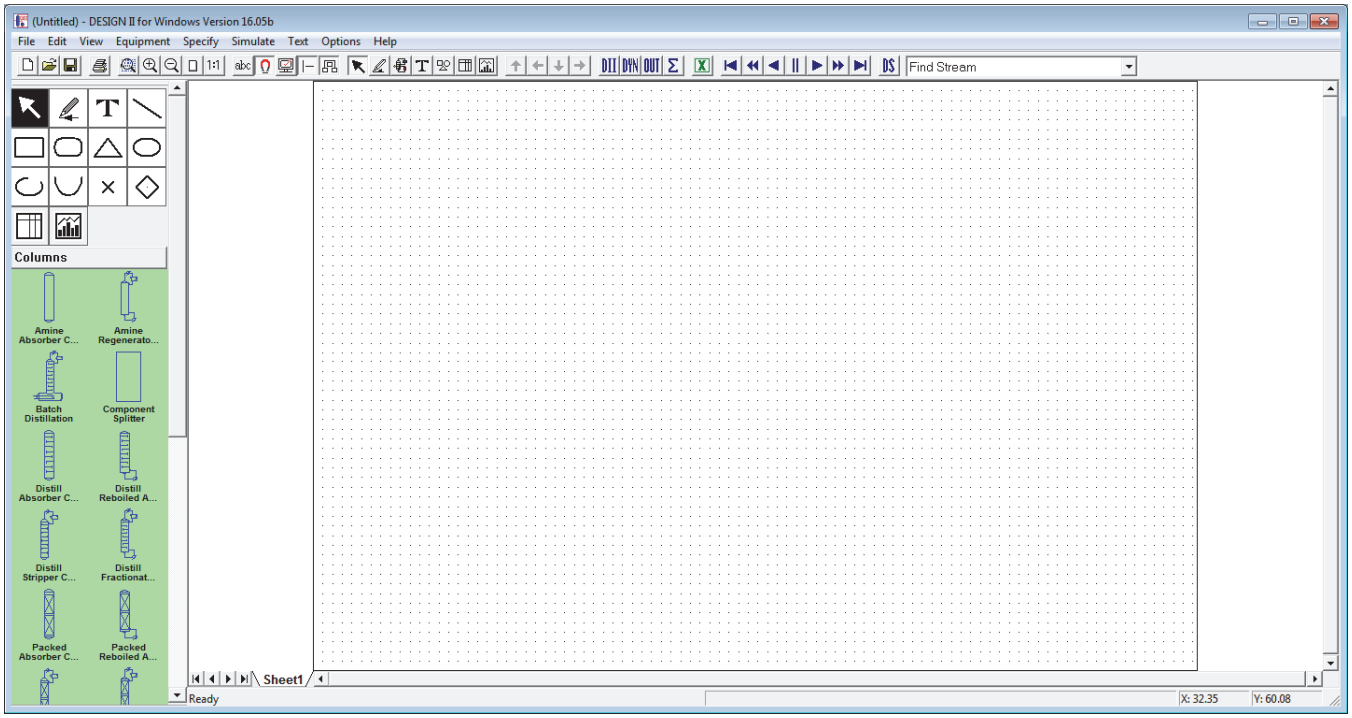


Figure 1 - 3: DESIGN II for Windows Screen

Let's take a look at the components of the DESIGN II for Windows screen:

Windows border/corners

The border/corners are the outside edge of a window; enlarge or reduce the size of a window by dragging on a border or corner

Title bar

The program and drawing names are displayed here. If you have not named your flowsheet, (**Untitled)** displays.

Control Menu box

Located in the upper left corner of the window; click on this box to display a menu listing Windows-related functions. You can move, resize, and close a window, or switch to another application. Refer to your ***Microsoft Windows User's Guide*** for details.

Minimize / Maximize buttons

Located in the upper right corner of the window; these buttons change the size of the window. The **Minimize** button (a box with an arrow pointing down) changes the window into an icon. The **Maximize** button (a box with an arrow pointing up) will enlarge a window to fill your screen. After you enlarge a window, the **Maximize** button changes into the **Restore** button (a box with arrows pointing up and down). Click the **Restore** button to return the window to its previous size.

Top Menu

A set of menus pop-up, or open, when you click on the menu name; these menus are File, Edit, Equipment, Specify, Simulate, Text, Options, Size, and Help. These menus list functions you will use to create a flowsheet and execute your simulation.

The **File** menu allows you to start a new flowsheet, open an existing one, save, rename, print, etc.

Use the **Edit** menu to cut/copy, paste, and delete objects.

Use the **View** menu to redraw the flowsheet or to change the appearance of the main window.

Select equipment symbols to place on your flowsheet using the **Equipment** menu.

Use the **Specify** menu to enter components, select thermodynamic methods, enter assay data, etc.

View the DESIGN II input and output files, and run the DESIGN II program using the **Simulate** menu.

Change text attributes such as font and size with the **Text** menu.

With the **Options** menu, select to display rulers, a grid, and snap points, or set other options.

For information about a task or function while using the program, select Help. Most dialogs have a Help button; help messages are contextual.

Another way to open a pop-up menu is to hold down the **Alt** key and type the letter in the menu that is underlined. For example, to open the Edit menu, hold down the **Alt** key and press the **E** key.

To close a menu, click anywhere outside the menu.

To select a function from a pop-up menu, click on the menu to open it, and then click on the desired function. The function is performed. An option is to drag the highlight down the menu to the desired function, then release the mouse button.

Flowsheet area

This is the workspace where you will draw a flowsheet. You can display horizontal and vertical rulers and a grid to help you draw the flowsheet. Units of measurements for the rulers and grid are either inches or centimeters.

Scroll bars

Horizontal and vertical bar are used for moving the view of the drawing to the left/right or up/down, respectively.

A scroll bar is made up of arrows on each end, the bar itself, and a scroll box. You can scroll by clicking on an arrow button once or continuously, dragging the scroll box to a new position on the scroll bar, or clicking the scroll bar to either side of the scroll box.

Sheet Tabs

The names of the sheets appear on tabs at the bottom of the flowsheet window. To move from sheet to sheet, click the sheet tabs. You can add, delete or rename sheets by right mouse clicking on any of the sheet tabs.

The sheets are named “Sheet1” through “SheetN” where N is the number of the sheet. You can change the name of the sheets to any name that you want, up to 40 characters long.

Toolbox

To construct the flowsheet, you add equipment to the drawing area, connect the equipment with streams for material flow, and add annotations/text/primitive drawing shapes (optional). You can select tools from the Toolbox to perform these activities.

You can click on the tool you want to use, then move to the flowsheet drawing area and use the selected tool. Following are the available tools:

Selection (arrow): Choose this symbol to use the standard tool for selecting and working with objects on the flowsheet, either singly or in groups by region.

Stream (pencil): Select this tool to add streams to your flowsheet.

Text (T): Select this tool to annotate or add text to your flowsheet.

Drawing elements/primitives (arc, line, etc.): Select these tools to add shapes to your flowsheet.

Equipment (categories): Select a category of equipment by clicking on the header name; the related types of equipment symbols/names are shown. Click on the desired equipment symbol and hold down the left mouse button as you drag (move the equipment) onto the desired position on the flowsheet. Release the mouse button when the equipment symbol is in the desired location. To hide the equipment symbols, click the category header name again.

Browser

A moveable tool palette used for drawing streams, text, arrows, or equipment. The Browser is composed of six tools: Selection (a left leaning arrow), Drawing Elements (a collection of rectangle, circle and triangle), Stream (a pencil leaning right), Text (a T), Flow Arrows, and Equipment (symbols).

Use the **Selection** tool to select an item or regions of items for cut, copy, paste, redraw, zoom, etc…

Use the **Stream** tool to draw streams and the **Text** tool to type information on the flowsheet.

Place flow direction arrows using the **Flow Arrow** tool and equipment symbols using the **Equipment** tool.

Later in this manual, we will refer to using different modes- stream, equipment, etc. What this means is some functions are enabled only when a certain tool on the Browser is selected. For example, click on the **Stream** tool to enter the stream mode.

Below the tools on the Browser is the Preview box; it shows what you can create with the selected tool. For the Flow Arrow or Equipment tool previews, you can change the orientation of the object by clicking on one of the four direction arrows.

Under the Preview box is a list of choices relevant to the selected tool. Select a line type for the Stream tool (solid, dashed, etc.). Choose a type size (tiny, large, etc.) for the Text tool. For the Flow Arrow tool, select a direction (an

option instead of clicking on a direction arrow in the Preview box). A list of equipment types is listed for the Equipment tool; scroll through the list using the scroll bar.

Specify Flowsheet Title

Click on the Specify menu and select Account and Title.

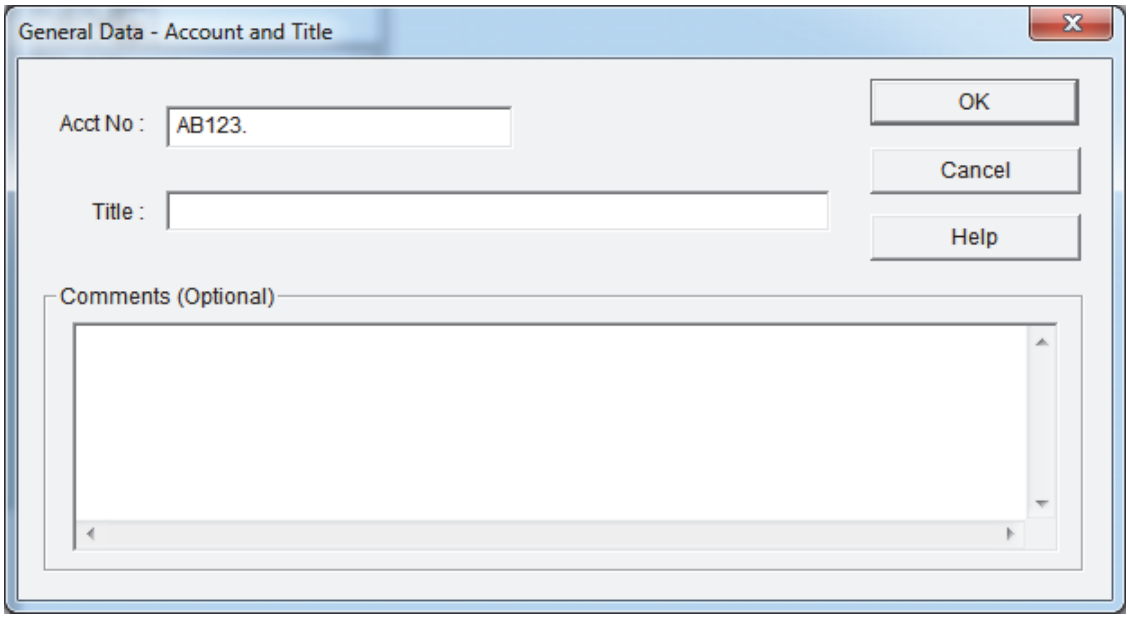


Figure 1 - 4: General Data- Account and Title Dialog

The General Data- Account and Title dialog box appears. Move the cursor to the **Title** box and click on it. Type *EXPANDER PLANT* and click **OK** to close the dialog box.

View a Help File

Let's look at the Help information on the Expander:

1. Open the **Help** menu and select **Contents**. The Help Index is displayed. Items displayed in blue are topics you can see more information on.

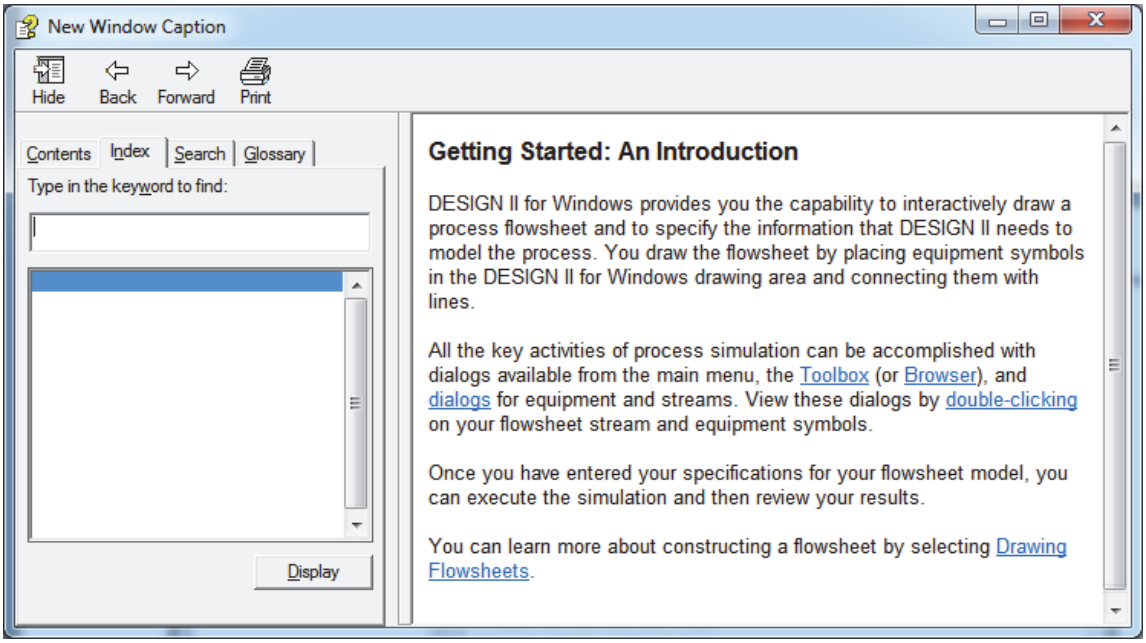


Figure 1 - 5: Help Index

2. Click on **Equipment Menu** under the **Menu** section. The list of equipment types is displayed.

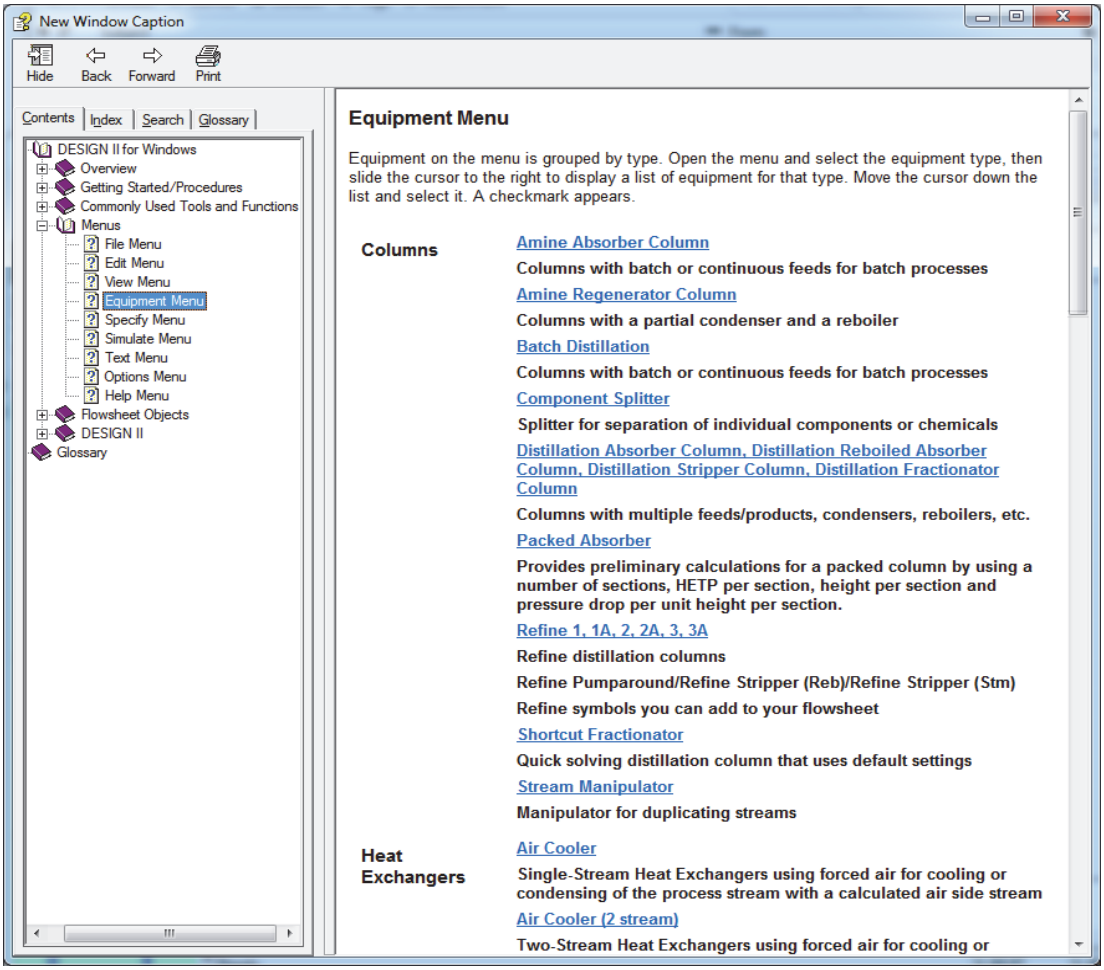


Figure 1 - 6: Equipment List

3. Click on **Expander**. Information on the expander is displayed. Read the file, and then click on other topics, if desired.

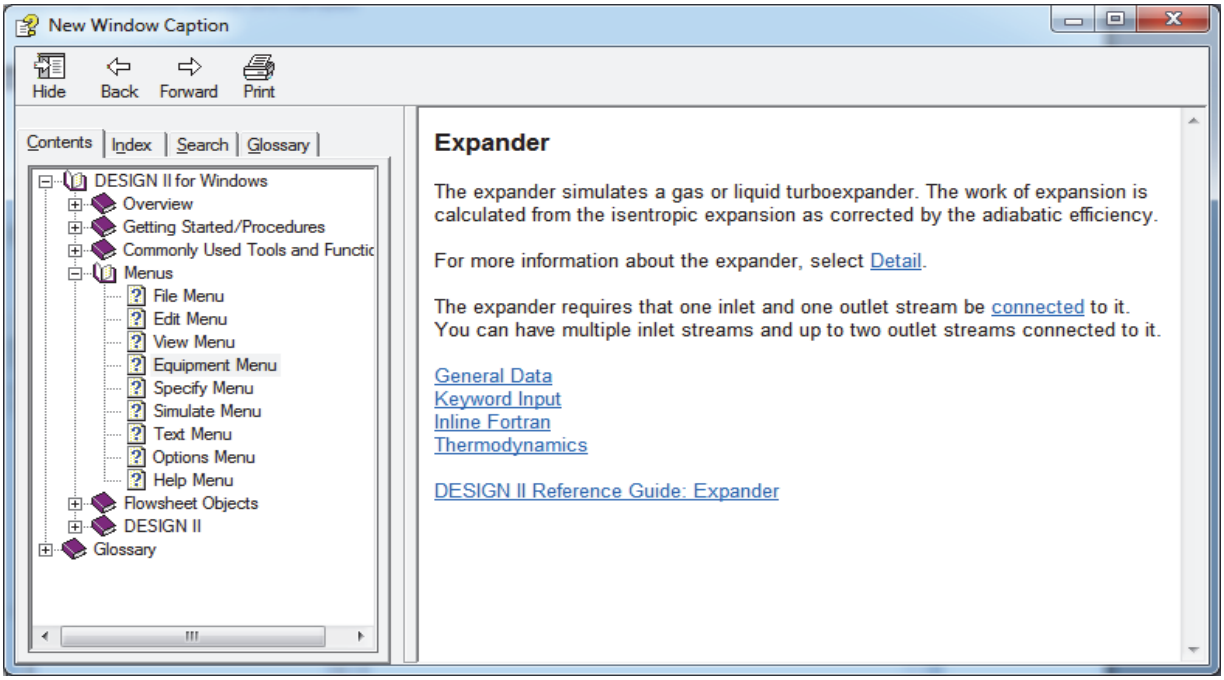


Figure 1 - 7: Expander Help

You can resize the DESIGN II for Windows and Help screens so that you can see both at the same time.

4. Open the **File** menu on the **Help** screen and select **Exit** when done.

Lay Out the Equipment

To select and place equipment:

1. Scroll down Toolbox to view **Equipment** categories.
2. Click on the desired equipment category to expand or collapse equipments of that type. Click and drag desired equipment symbol to flowsheet. For our example, drag **Expander** on to flowsheet as the first piece of equipment.

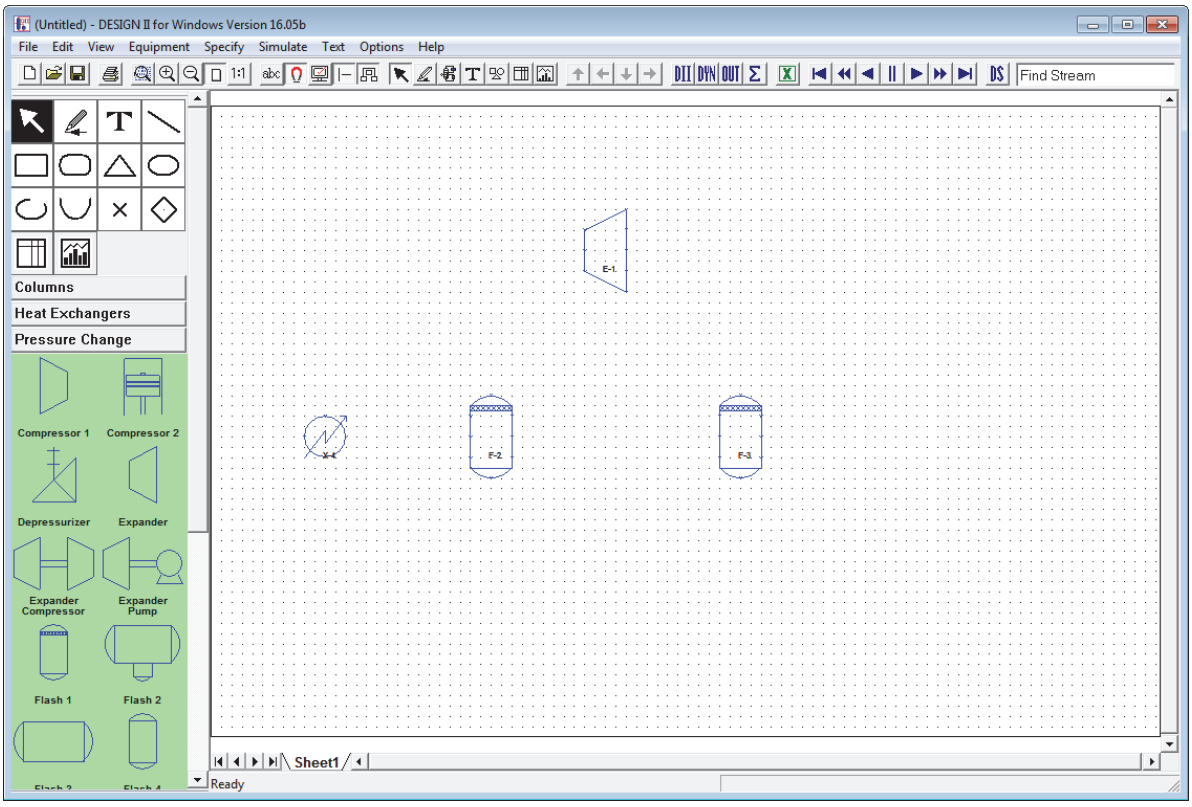


Figure 1 - 8: Layout of Expander Plant Equipment

3. Select a location on the flowsheet to place the equipment item, and then release mouse button to place it. If the initial placement of an equipment item is not satisfactory, you can move it.

   To move the item, click on it (the outline becomes dashed instead of solid), drag the equipment to a new location, and release the mouse button. You may click anywhere inside the item; you do not have to click on the symbol outline to select the item.

   To delete the item, click on it to select it and use delete key. Alternatively, you can use the right mouse button, and choose the delete option from the drop-down menu.

   Repeat steps 2-3 to place additional equipment.

   For our example, select **Flash 1** for the separators.

   NOTE: The equipment items are numbered in the sequence you place them.

Add Streams and Labels

To draw a straight line stream:

1. Click on the **pencil** icon on the Toolbox. The cursor changes to a pencil, angled to the left.
2. Click on the **Options** menu and select **Show Snap Points**. This will show where streams may connect to the various equipment symbols.
3. To begin drawing a stream for our example, position the pencil to the far left of the heat exchanger. Then draw a line to the left side snap point on the heat exchanger as illustrated on the figure below.

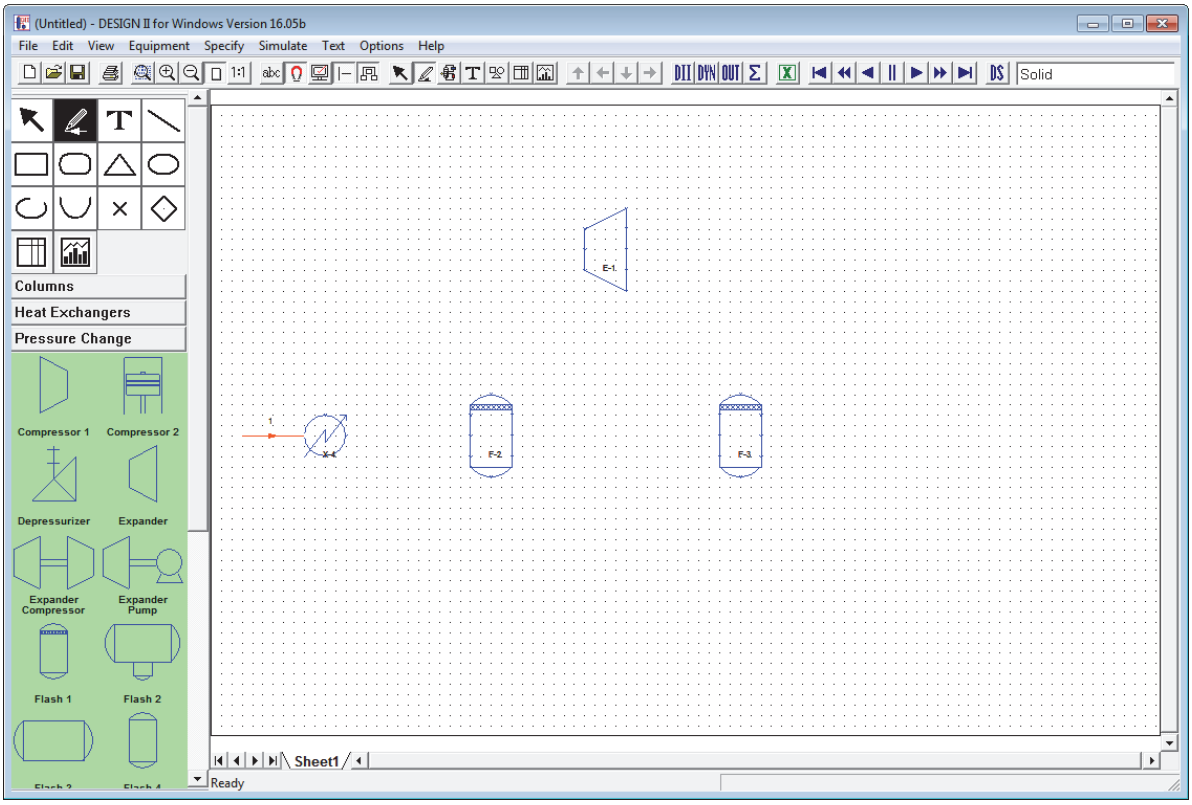


Figure 1 - 9: Start the Stream

4. Click a second time at the exchanger to complete the line. The dashed lines disappear and a stream label appears near the stream. Note that the line arrow will be automatically added to the stream in the direction that the stream was drawn. You can place additional arrows on the streams or remove any arrows that you do not want to see on the drawing.

   Streams are numbered in the sequence you draw them. These can be changed, if desired. Position the pencil on the stream and double-click. The Stream dialog will appear. Type in a new number; then click on **OK**. A warning message will appear if the number is already in use.

   NOTE: The program will use the stream and equipment connections to set up the topology for flowsheet simulation.

   NOTE: The program will use the direction of flow arrows to determine flowsheet topology and calculation order in the simulation.

5. Draw a stream from the exchanger to flash F-2. To bend a stream:

   a. Click at the top of flash F-2 and drag the line upwards.

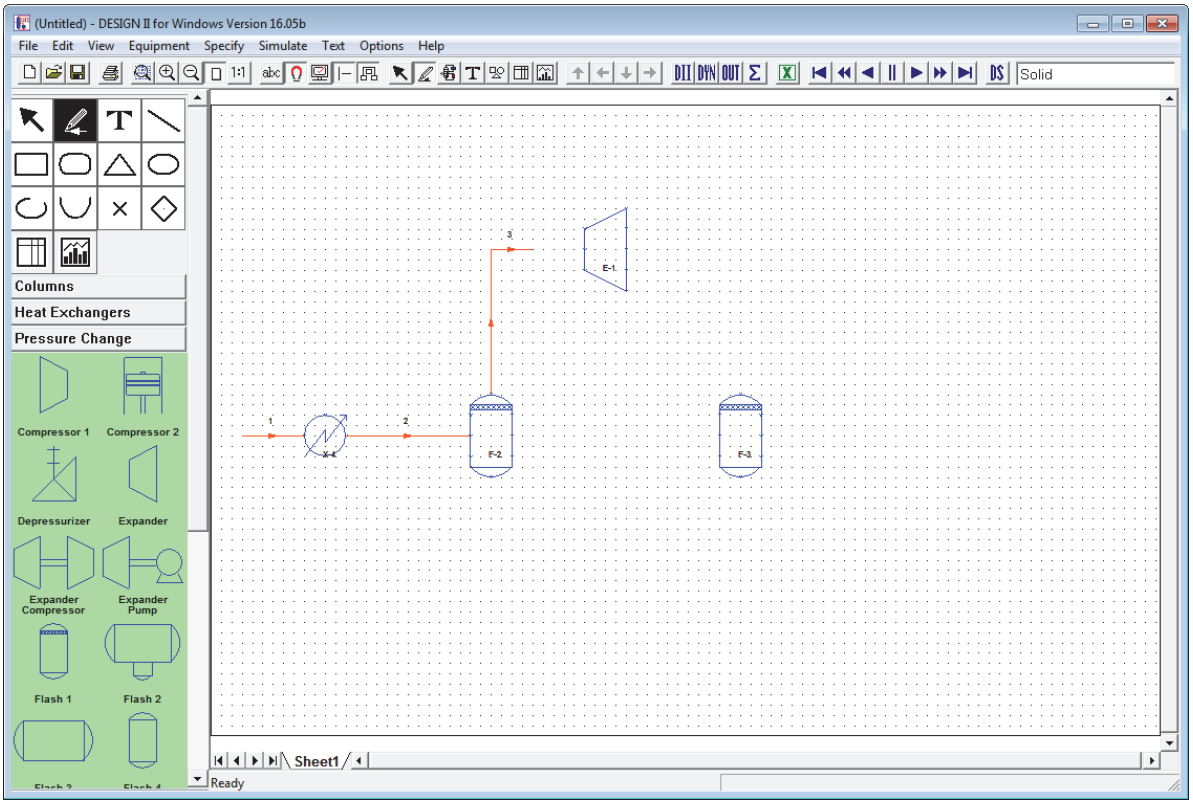


Figure 1 - 10: Click to Start a Bend

a. Click again when the pencil is at the place of the desired bend.

b. Drag the pencil in the new direction.

c. Click at the expander snap point, and then click again to complete the line.

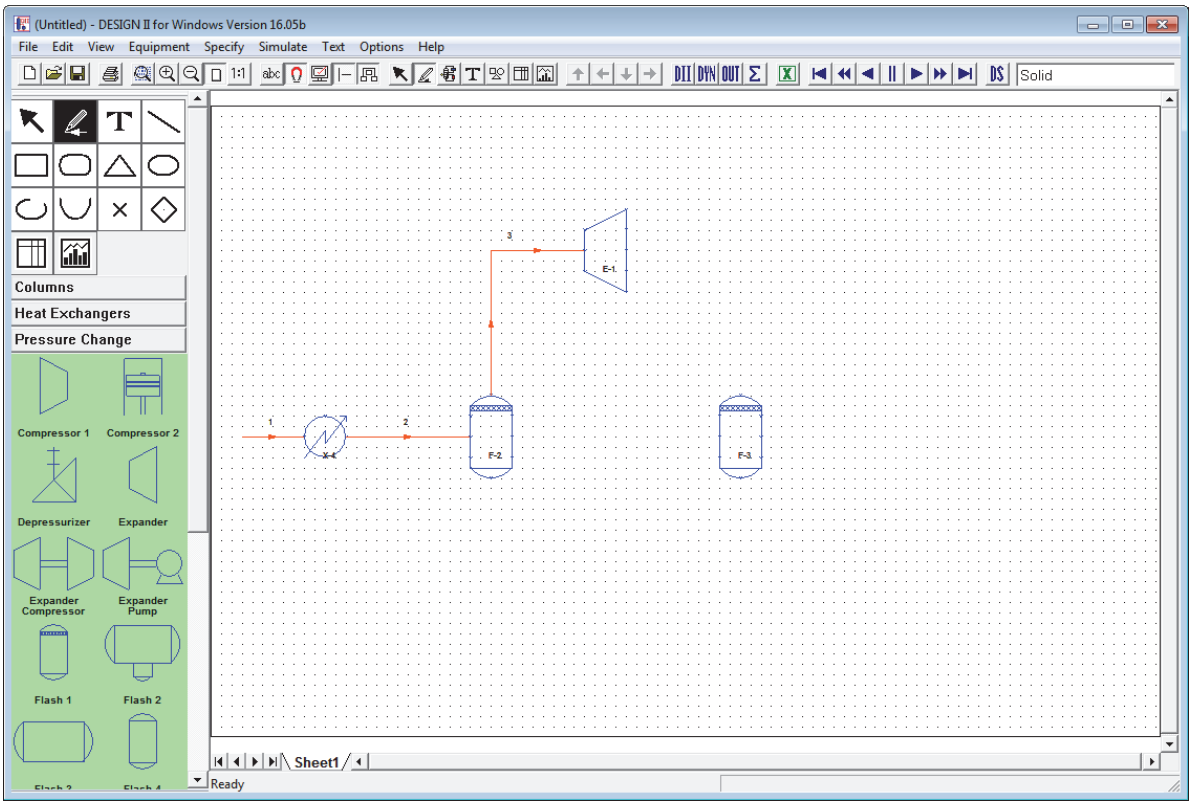


Figure 1 - 11: Click Twice to End the Line

6. Complete the flowsheet by drawing the remaining streams shown in the Figure below.

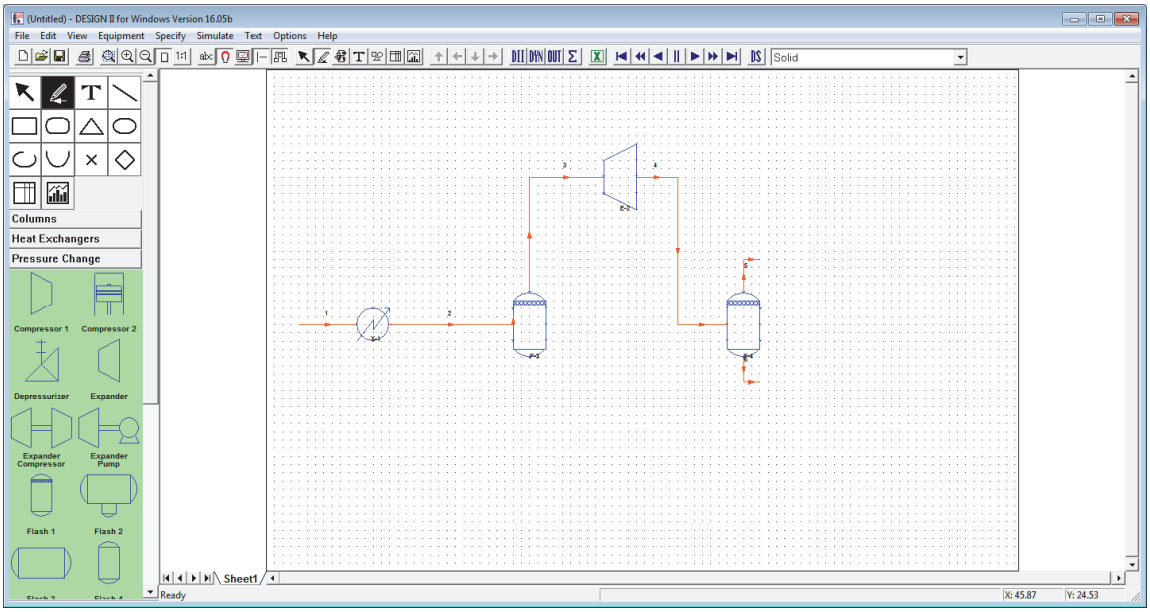


Figure 1 - 12: Draw the Remaining Streams

Add a Title to the Drawing

1. Click on the **T icon** on the Toolbox to enter text.
2. To set the font style and size for the type, click in an open area of the drawing to place the title. Right click and select font from the menu. A dialog will open allowing you to select font type and size. For this example click on **Arial** to set the font and **48** to set the size.
3. Type *DESIGN II for Windows*.

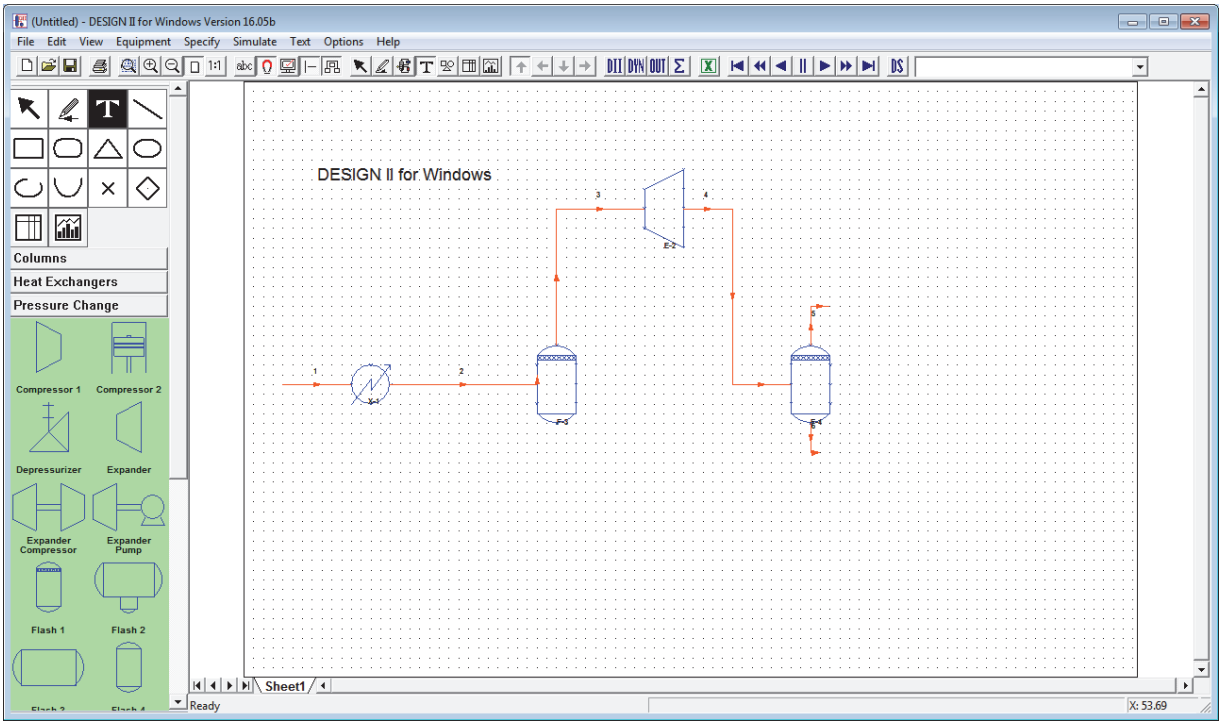


Figure 1 - 13: Completed Process Simulation Diagram (PSD)

This completes your Process Simulation Diagram.

Specify the Feed Stream

The feed stream is specified using the Stream Mode or the Selection mode.

1. Double click on the feed stream (Stream 1) and the **Stream** dialog box will pop up. You can use this dialog to edit the stream name and number. Please note stream numbers and equipment numbers must be unique. Check the **Display box** next to Name and/or Number to show the information on the flowsheet.

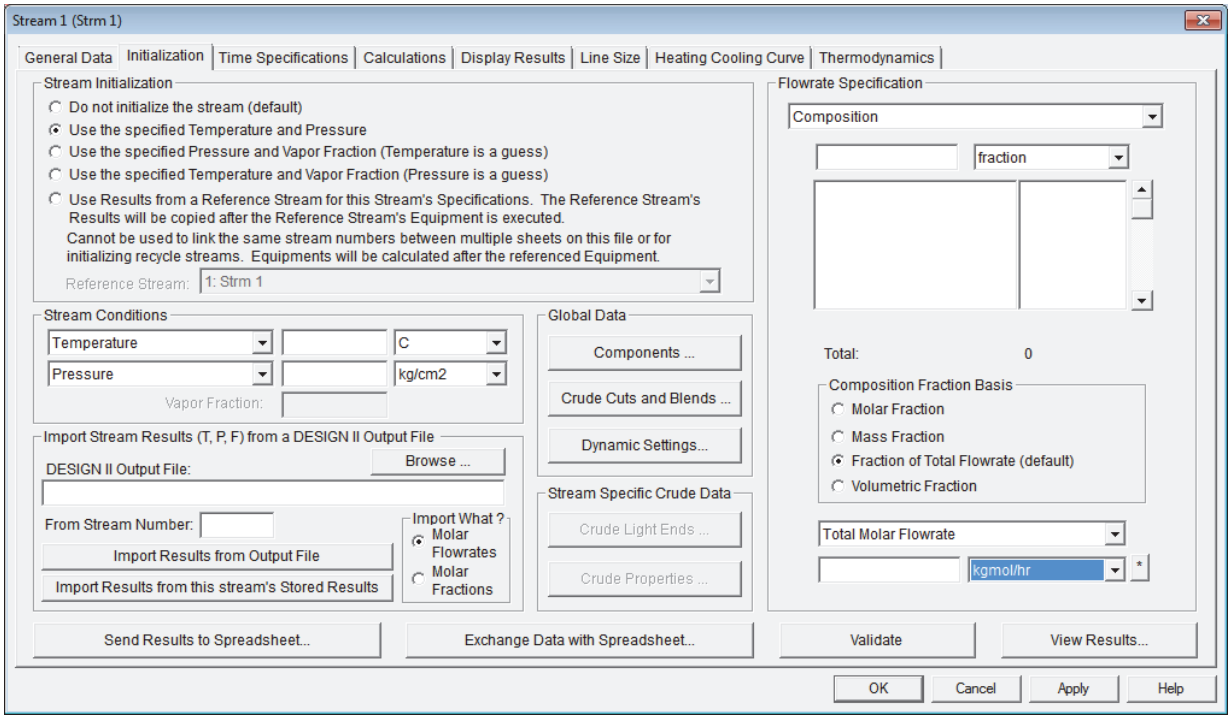


Figure 1 - 14: Stream Specifications Dialog Box

2. In this example select "Use the specified Temperature and Pressure" option under Stream Initialization.
3. Next you need to specify the components used for this flowsheet. Components are selected for the entire flowsheet and are the same for all the streams and equipment modules. Click the **Components…** button to bring up the Component Dialog. You can also pull down the **Specify** menu and select the **Components…** menu item to get the Components dialog.

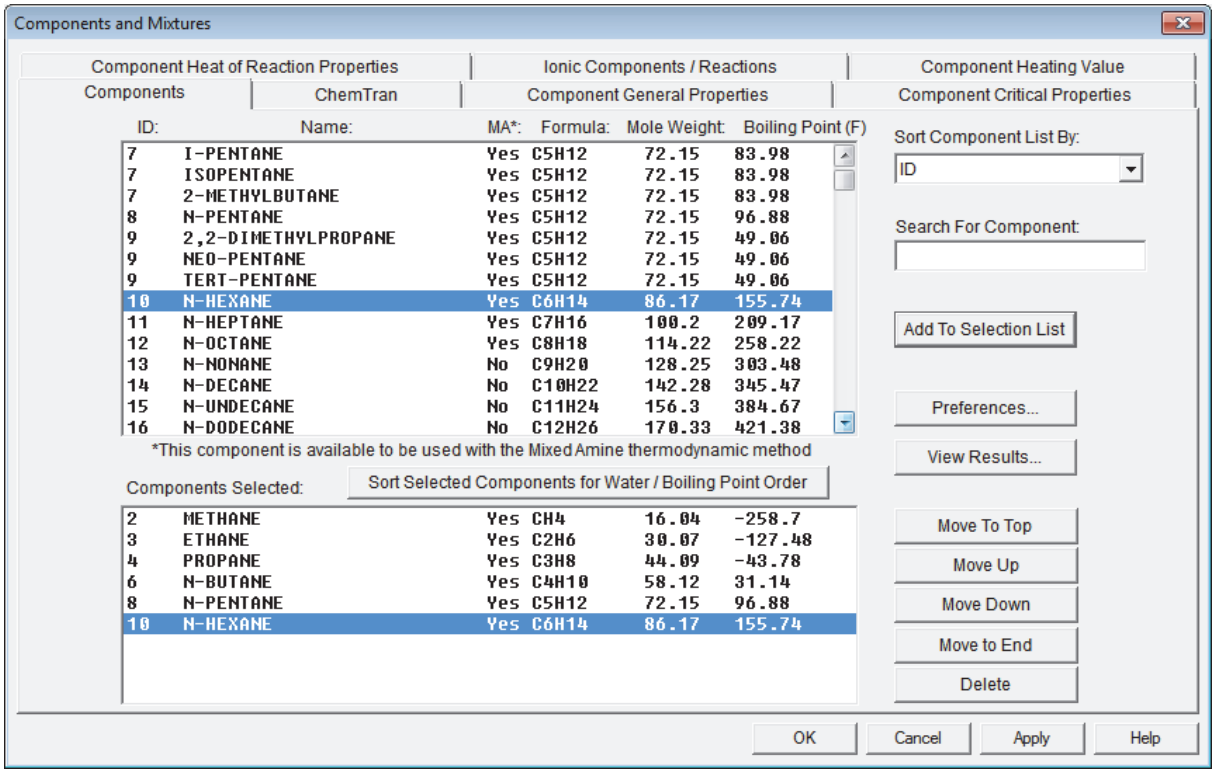


Figure 1 - 15: Components Selection Dialog Box

For our example, select the following components:

METHANE, ETHANE, PROPANE, N-BUTANE, N-PENTANE, N-HEXANE

Components are added to the end of the list in the **Components Selected** box. Use the **To Top** button or the **To End** button after selecting a component to move it to the top or bottom of the list, respectively. To delete a component from the list, click on it then click on the **Delete** button. Click **OK** when done to save the component selections you have made.

4. Change the dimensional units for the stream pressure from **psia** to **psig** by clicking on the **down arrow** button and selecting **psig**.
5. Enter the temperature **90 F** and the pressure **980 psig** for the stream initial conditions.
6. Select **Comp Molar Flow** from the Flowrate specification box; click on the **down arrow** button and select your choice from the pull-down box appears.
7. Click on the first component and type the flowrates from the following list into the edit box.

Component	Lbmol/Hr
Methane	6100
Ethane	500
Propane	200
N-Butane	100
N-Pentane	100
N-Hexane	70

8. Highlight the next component and type in the flowrate.

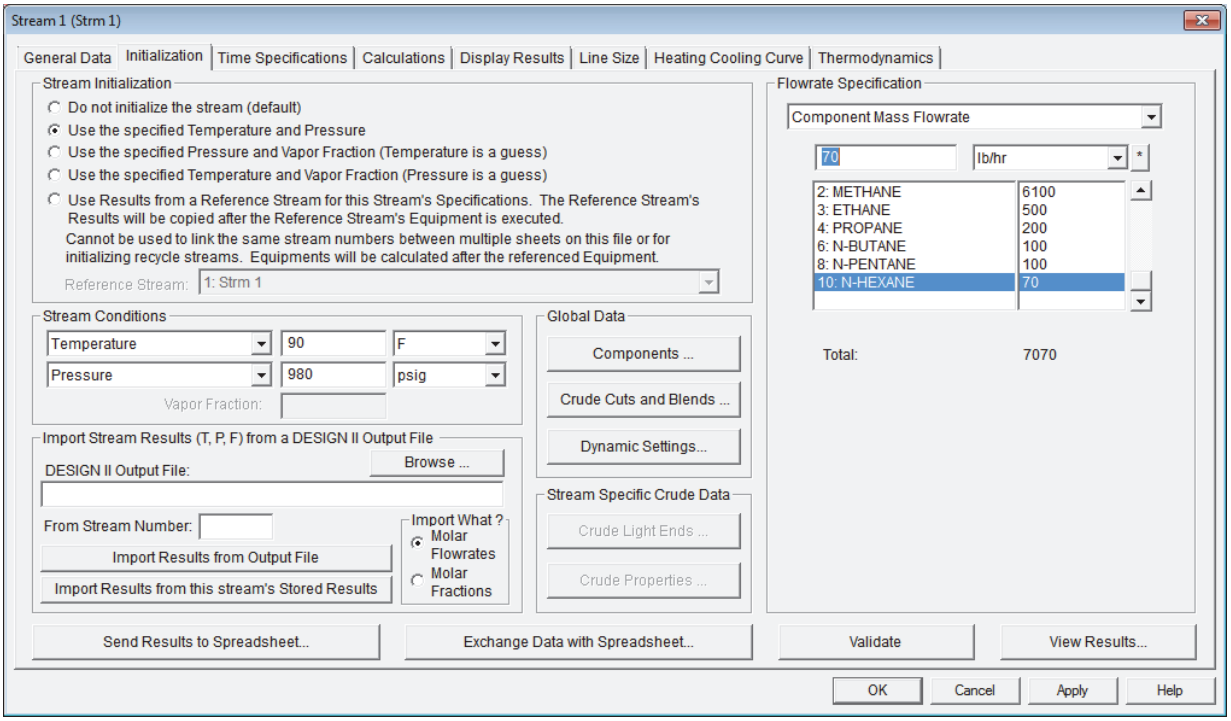


Figure 1 - 16: Stream Specifications Dialog Box

9. Click on **OK,** or press **Enter** when done, to save the data you entered on the Stream specifications dialog box.
10. Click on **OK** to close the Stream dialog box.

Enter Equipment Specifications

Enter equipment specifications.

1. Double click on the heat exchanger (X-1) to bring up the Heat Exchanger dialog box. You can use this dialog box to edit the name and number of the equipment, if desired. You can also select to display the equipment name on the drawing.

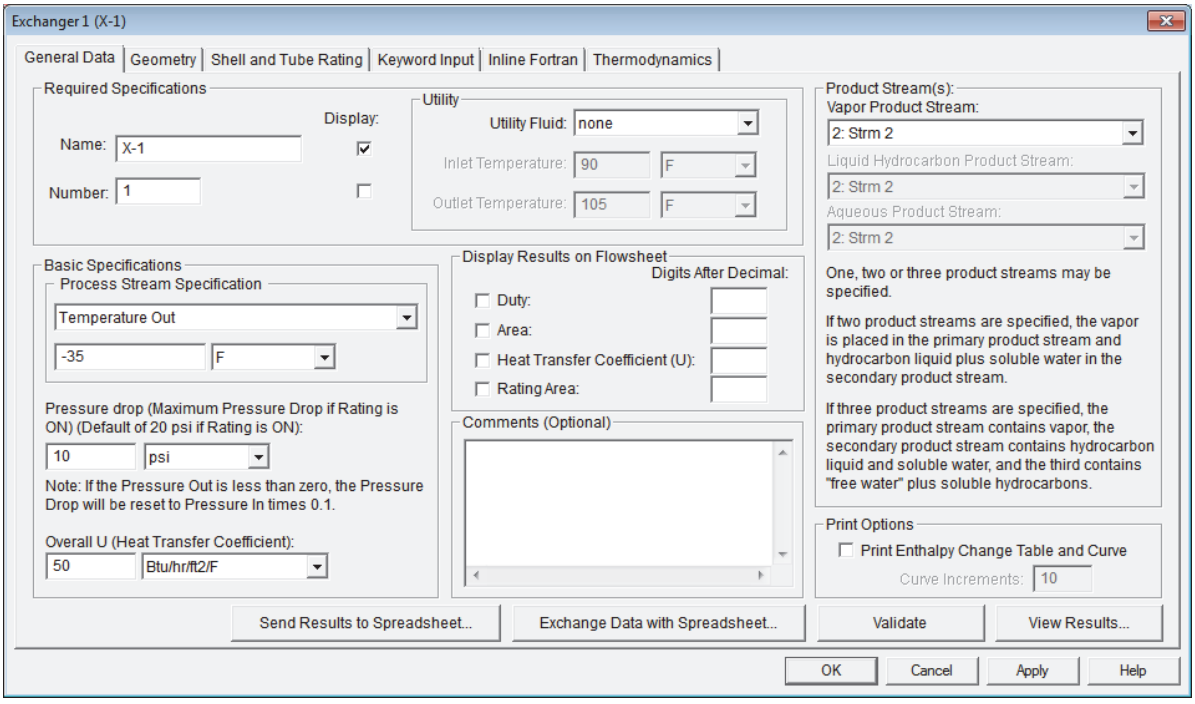


Figure 1 - 17: Heat Exchanger Dialog Box

3. Select the **Temp Out** specification and type *-35*. Make sure the temperature units are F.

4. Press **Tab**, moving the cursor to the Pressure Drop edit box; type *10*. Make sure the pressure units are psi.
5. Click on **OK** to accept the Exchanger 1 specifications dialog box.
6. Now, you will need to enter specifications for the remaining equipment. Although the dialog names and required information are different than the Heat Exchanger procedures outlined above, you can apply the basic concepts from Steps 1-7 to enter the specifications for the feed chiller and expander.

   Since an adiabatic flash with zero pressure drop is the default specification for the vertical separators, you do not need to supply specifications for them.

   Enter the data from the table below.

Equipment	Tag	Specifications
Feed Chiller	X-1	Outlet Temperature = -35°F Delta Pressure = 10 psi
Expander	E-2	Pressure = 275 psig Efficiency = .80
Flash	F-3	Vapor Outlet Stream = Stream 3 Liquid Outlet Stream = Stream 7
Flash	F-4	Vapor Outlet Stream = Stream 5 Liquid Outlet Stream = Stream 6

   After specifying the last equipment item, deselect the item by clicking once in an open area of the drawing.

Specify the Thermodynamic Methods

The Thermodynamic Methods can be selected from the **Specify** menu by choosing the **Basic Thermo** option.

1. Click on **Specify** and select **Basic Thermo.** The Thermodynamic and Transport Properties dialog box is displayed.

   Please note your choice of a thermodynamic method is critical to the quality of the process simulation. Make sure the thermodynamic method is appropriate to the system you are modeling. You can click the **Help** button to get a detailed description of the thermodynamic methods available in DESIGN II.

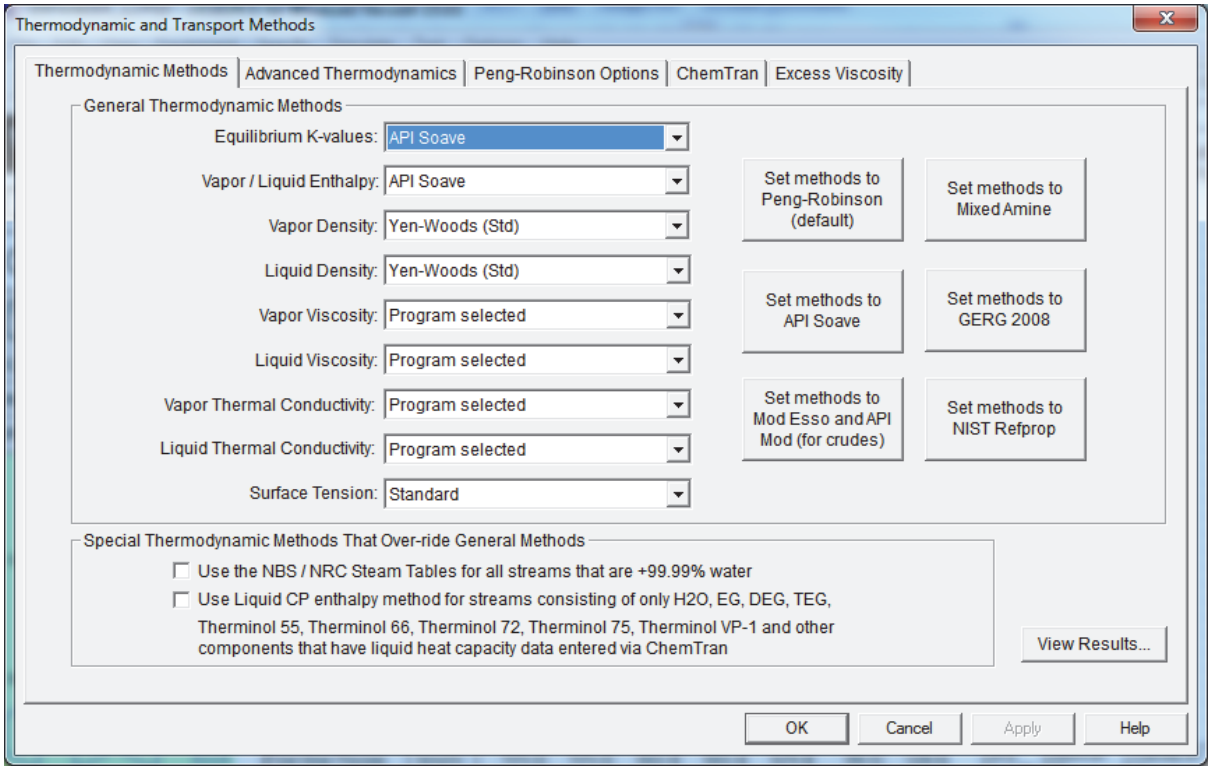


Figure 1 - 18: General Data - Basic Thermodynamics Dialog Box (from expander.psd)

2. Alternative calculation methods may be selected by clicking on the **down arrow** with an underscore next to the calculation type you want to change.

   A combo box is opened, containing all methods available for that type of calculation; click on the desired method.

NOTE: The defaults for Enthalpy and K-value methods have been changed to APISOAVEK and APISOAVEH (default thermodynamic methods for the keyword input versions of DESIGN II are STDH and STDK).

3. The following are default selections; they do not need to be changed for our example.

Enthalpy	Density	Viscosity	Thermal Conductivity
API Soave	Yen Woods (STD)	Program selected	Program selected

4. Click on **OK** when done.

   For viscosity and thermal conductivity calculations, Program selected means standard DESIGN II defaults are used. Methods are changed for input files with Crude feed sections. Please check your online ***DESIGN II User Guide*** for a description of these defaults.

Save and Name the Flowsheet

To name and save the flowsheet:

1. Click on **File** and select **Save As**.
2. Type ***Expander*** and click on **Save** or press **Enter**.

   The flowsheet is saved as EXPANDER.PSD, replacing Untitled at the top of the window.

   In the next section, when you view the input file or run the simulation, an input file is created and saved as EXPANDER.IN.

View the Input File

The input file for DESIGN II is automatically prepared using the specifications you entered. This step is optional; if you do not want to view the input file, turn to the ***Execute the Simulation*** section in this chapter.

To view the input file:

1. Click on **Simulate** and select **View Input**. The file appears in a window on-screen. Use the **Scroll bar** or the **Page Down** key to view more material if the input file is longer than one screen.
2.

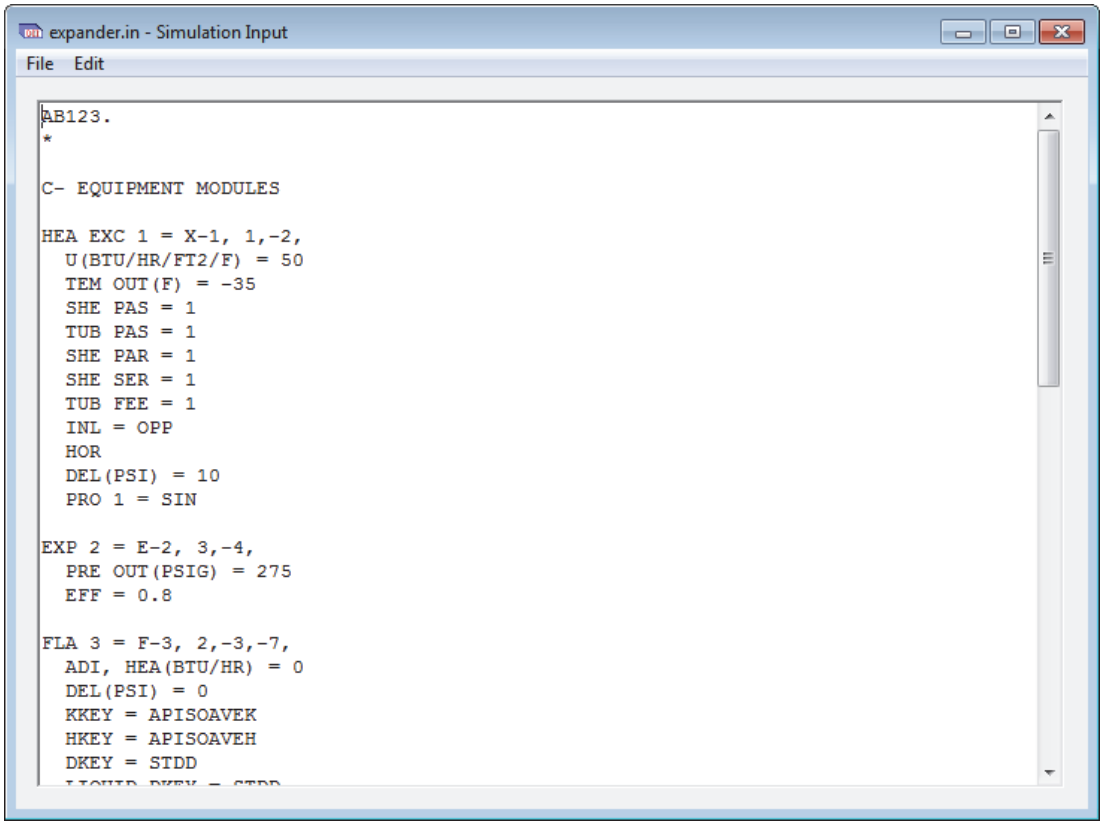


Figure 1 - 19: Expander Plant Input File (from expander.psd)

NOTE: If there are any omissions in the file, a checklist will appear instead of the input file. The list indicates the missing data or required items. Double click on a message to go directly to the dialog box where the missing data is needed.

2. After viewing the file, close its window by double clicking on its **control menu box** (the dash in the upper left corner). The input file window is closed.

Execute the Simulation

You will use DESIGN II to run a simulation using the input file you created.

1. Click on Simulate and select Execute Steady State or Execute Dynamic. For this example select Execute Steady State. DESIGN II loads the file, and after a few moments you will see DESIGN II run time messages.

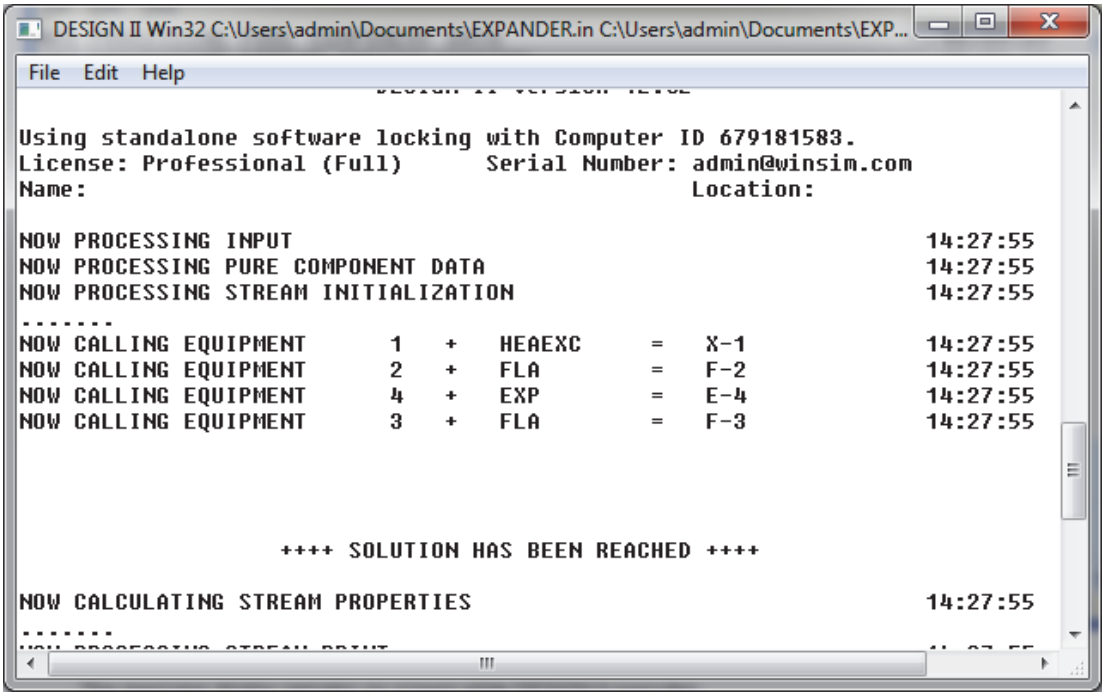


Figure 1 - 20: Simulation Status during Program Execution (from expander.psd)

The message display remains on-screen while DESIGN II executes.

2. When DESIGN II calculations are complete, a new dialog, the Simulation Summary, will appear on-screen containing the results of the simulation.

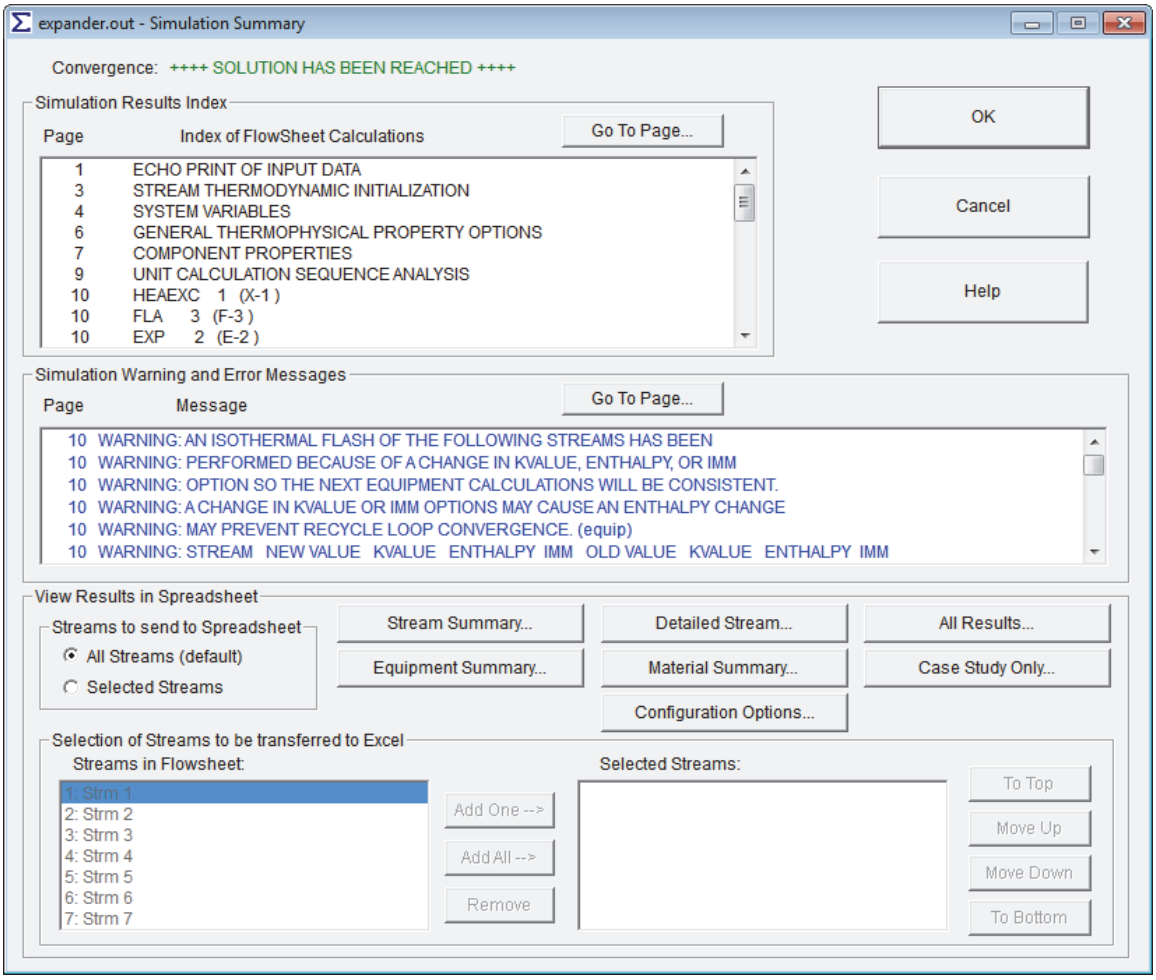


Figure 1 - 21: Simulation Summary Dialog Box (from expander.psd)

View the Simulation Results

The simulation summary consists of four parts: the Convergence message, the Index of Flowsheet Calculations, Simulation Warning and Error Messages, and the View Results in Spreadsheet showing transfer buttons.

The convergence message will be **++++ SOLUTION HAS BEEN REACHED ++++** if the simulation converged normally.

You can view any page of the DESIGN II output file by double clicking on a line in the **Index of Flowsheet Calculations**. You can also select a line in the Index of Flowsheet Calculations and click the **Go To Page...** button. This will pop up the output file viewer (Figure 3-23) with the selected output page.

To view subsequent pages of the file, use the **Page Down** button to move forward one page at a time; **Page Up** will move back one page at a time. If you click on the **Index** button, the first page of the Index of Flowsheet Calculations will be displayed. **Go To** allows you to input a specific page number to view. Use the scroll bar to move up or down within a page.

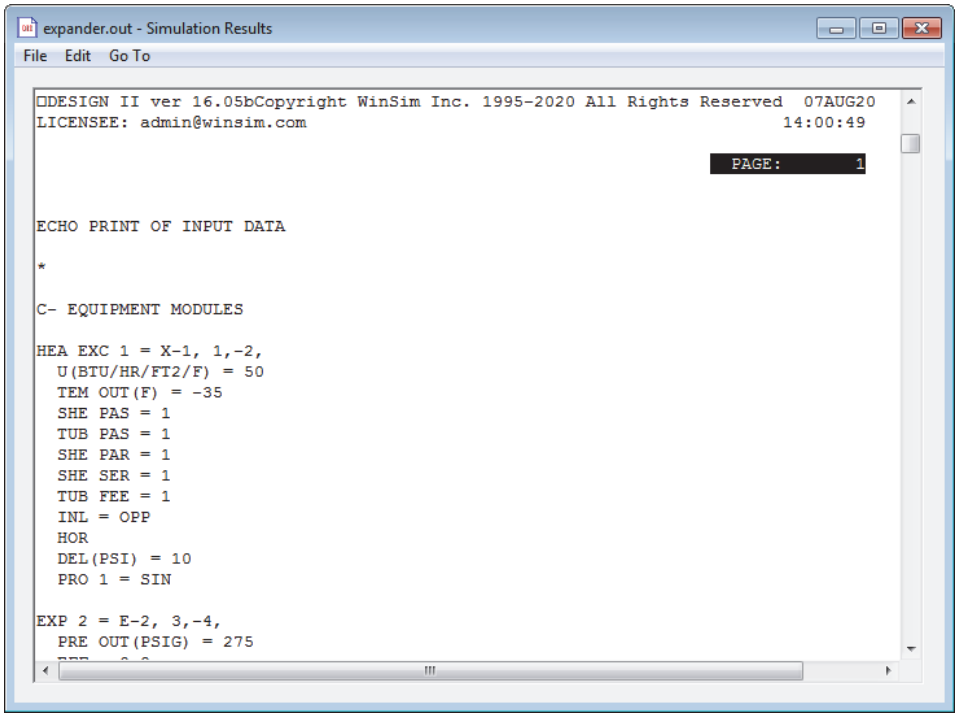
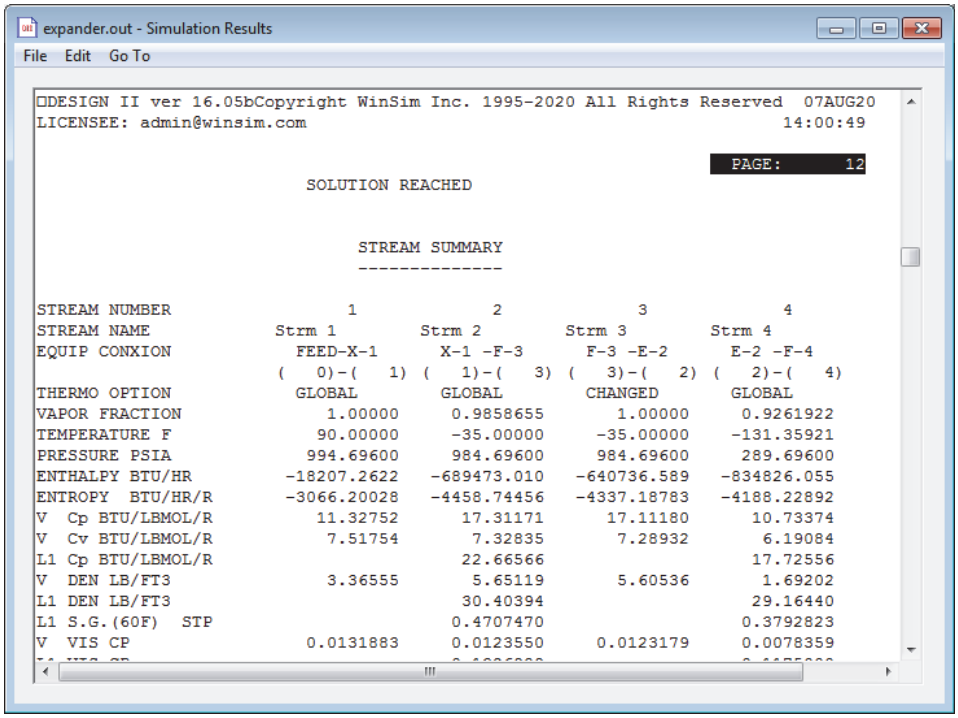


Figure 1 - 22: DESIGN II Output File (from expander.psd)

Close the output file by clicking on the **File** menu and selecting **Close**. This returns you to the Simulation Summary dialog. To view the Stream Summary, double click on **STREAM SUMMARY** line in the Index of Flowsheet Calculations. The output viewer will popup with the stream summary page in it.



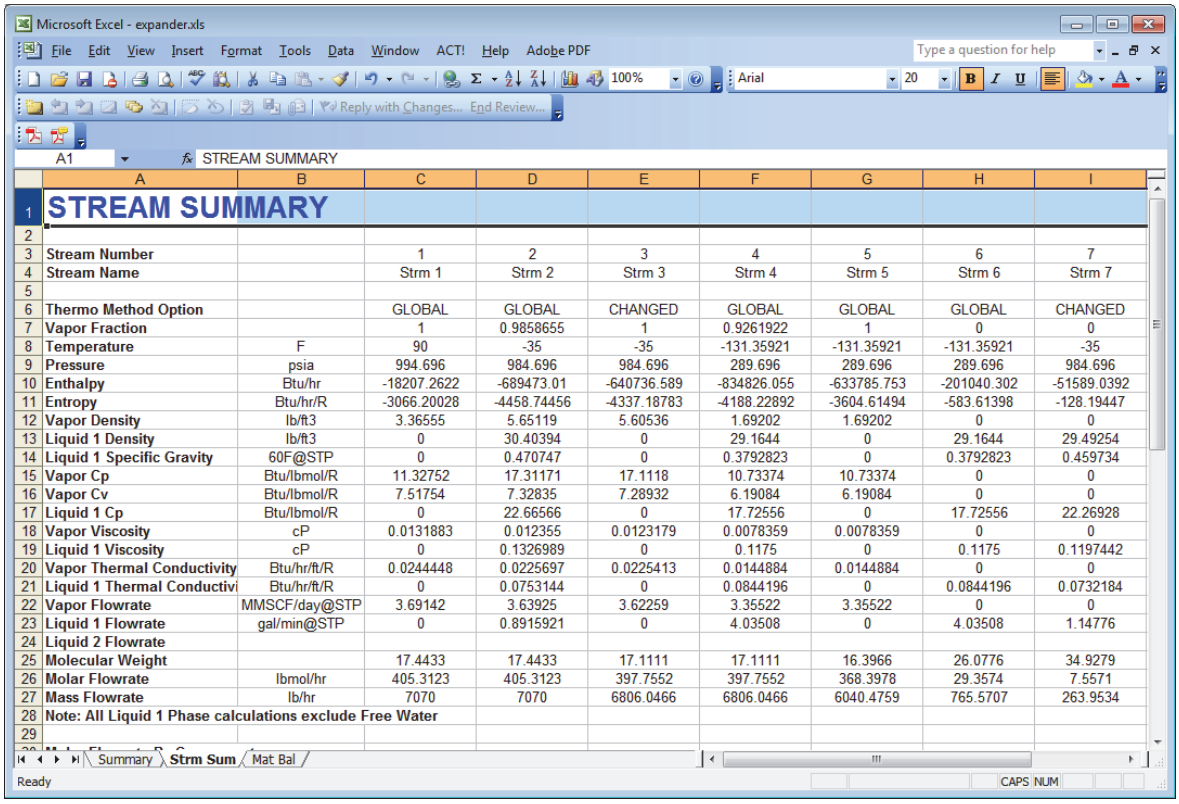
expander.out - Simulation Results

File Edit Go To

```
DESIGN II ver 16.05bCopyright WinSim Inc. 1995-2020 All Rights Reserved  07AUG20
LICENSEE: admin@winsim.com                                              14:00:49

                                                                  PAGE:       12

                          SOLUTION REACHED


                                STREAM SUMMARY
                                --------------

STREAM NUMBER                1            2            3            4
STREAM NAME            Strm 1       Strm 2       Strm 3       Strm 4
EQUIP CONXION            FEED-X-1     X-1 -F-3     F-3 -E-2     E-2 -F-4
                       (   0)-(   1) (   1)-(   3) (   3)-(   2) (   2)-(   4)
THERMO OPTION            GLOBAL       GLOBAL       CHANGED      GLOBAL
VAPOR FRACTION          1.00000      0.9858655      1.00000     0.9261922
TEMPERATURE F           90.00000     -35.00000     -35.00000    -131.35921
PRESSURE PSIA           994.69600    984.69600     984.69600    289.69600
ENTHALPY BTU/HR        -18207.2622  -689473.010   -640736.589  -834826.055
ENTROPY  BTU/HR/R      -3066.20028  -4458.74456   -4337.18783  -4188.22892
V  Cp BTU/LBMOL/R        11.32752     17.31171      17.11180     10.73374
V  Cv BTU/LBMOL/R         7.51754      7.32835       7.28932      6.19084
L1 Cp BTU/LBMOL/R                     22.66566                   17.72556
V  DEN LB/FT3             3.36555      5.65119       5.60536      1.69202
L1 DEN LB/FT3                         30.40394                   29.16440
L1 S.G.(60F)  STP                     0.4707470                  0.3792823
V  VIS CP               0.0131883    0.0123550     0.0123179    0.0078359
```

Figure 1 - 23: Stream Summary (from expander.psd)

If you have Microsoft Excel 95 or later, you can transfer your results to Excel by clicking one of the five buttons in the **View Results in Spreadsheet** group. Excel offers you another view of your flowsheet results. You can also perform additional calculations or create graphs and reports to meet your needs and specifications.



Microsoft Excel - expander.xls

File Edit View Insert Format Tools Data Window ACT! Help Adobe PDF

A1 STREAM SUMMARY

	A	B	C	D	E	F	G	H	I
1	**STREAM SUMMARY**								
2									
3	**Stream Number**		1	2	3	4	5	6	7
4	**Stream Name**		Strm 1	Strm 2	Strm 3	Strm 4	Strm 5	Strm 6	Strm 7
5									
6	**Thermo Method Option**		GLOBAL	GLOBAL	CHANGED	GLOBAL	GLOBAL	GLOBAL	CHANGED
7	**Vapor Fraction**		1	0.9858655	1	0.9261922	1	0	0
8	**Temperature**	F	90	-35	-35	-131.35921	-131.35921	-131.35921	-35
9	**Pressure**	psia	994.696	984.696	984.696	289.696	289.696	289.696	984.696
10	**Enthalpy**	Btu/hr	-18207.2622	-689473.01	-640736.589	-834826.055	-633785.753	-201040.302	-51589.0392
11	**Entropy**	Btu/hr/R	-3066.20028	-4458.74456	-4337.18783	-4188.22892	-3604.61494	-583.61398	-128.19447
12	**Vapor Density**	lb/ft3	3.36555	5.65119	5.60536	1.69202	1.69202	0	0
13	**Liquid 1 Density**	lb/ft3	0	30.40394	0	29.1644	0	29.1644	29.49254
14	**Liquid 1 Specific Gravity**	60F@STP	0	0.470747	0	0.3792823	0	0.3792823	0.459734
15	**Vapor Cp**	Btu/lbmol/R	11.32752	17.31171	17.1118	10.73374	10.73374	0	0
16	**Vapor Cv**	Btu/lbmol/R	7.51754	7.32835	7.28932	6.19084	6.19084	0	0
17	**Liquid 1 Cp**	Btu/lbmol/R	0	22.66566	0	17.72556	0	17.72556	22.26928
18	**Vapor Viscosity**	cP	0.0131883	0.012355	0.0123179	0.0078359	0.0078359	0	0
19	**Liquid 1 Viscosity**	cP	0	0.1326989	0	0.1175	0	0.1175	0.1197442
20	**Vapor Thermal Conductivity**	Btu/hr/ft/R	0.0244448	0.0225697	0.0225413	0.0144884	0.0144884	0	0
21	**Liquid 1 Thermal Conductivi**	Btu/hr/ft/R	0	0.0753144	0	0.0844196	0	0.0844196	0.0732184
22	**Vapor Flowrate**	MMSCF/day@STP	3.69142	3.63925	3.62259	3.35522	3.35522	0	0
23	**Liquid 1 Flowrate**	gal/min@STP	0	0.8915921	0	4.03508	0	4.03508	1.14776
24	**Liquid 2 Flowrate**								
25	**Molecular Weight**		17.4433	17.4433	17.1111	17.1111	16.3966	26.0776	34.9279
26	**Molar Flowrate**	lbmol/hr	405.3123	405.3123	397.7552	397.7552	368.3978	29.3574	7.5571
27	**Mass Flowrate**	lb/hr	7070	7070	6806.0466	6806.0466	6040.4759	765.5707	263.9534
28	**Note: All Liquid 1 Phase calculations exclude Free Water**								
29									

Summary | Strm Sum | Mat Bal

Ready CAPS NUM

Figure 1 - 24: Stream Summary in Microsoft Excel (from expander.psd)

When you have finished reviewing the DESIGN II results, click on the **Exit** button.

You can also review equipment and stream results from the flowsheet using the **View Results** button for each equipment or stream.

1. Select the appropriate mode from the **Toolbox** (Equipment or Stream).
2. When the cursor is pencil-shaped (for streams), double-click on the stream whose results you want to review.
3. The Stream dialog box will appear. Click on **View Results**. The detailed stream report will appear on-screen.
4. With the Equipment mode selected, steps 2 and 3 are used to view selected equipment's results.

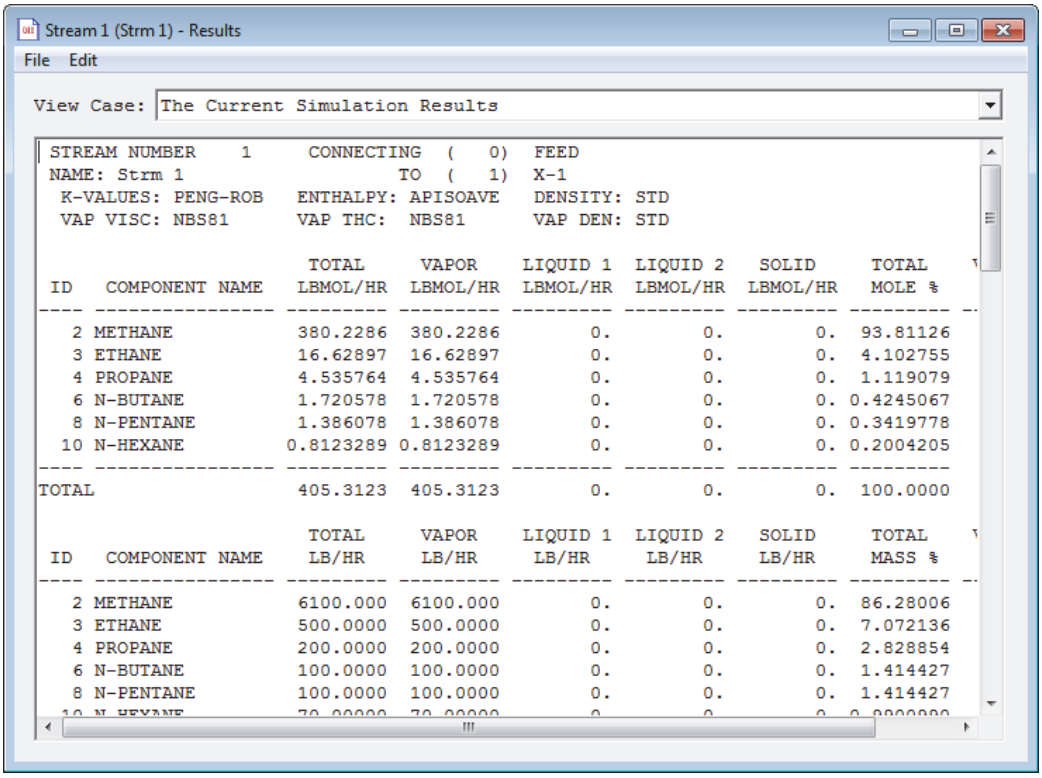


Figure 1 - 25: View Results for Stream 1 (from expander.psd)

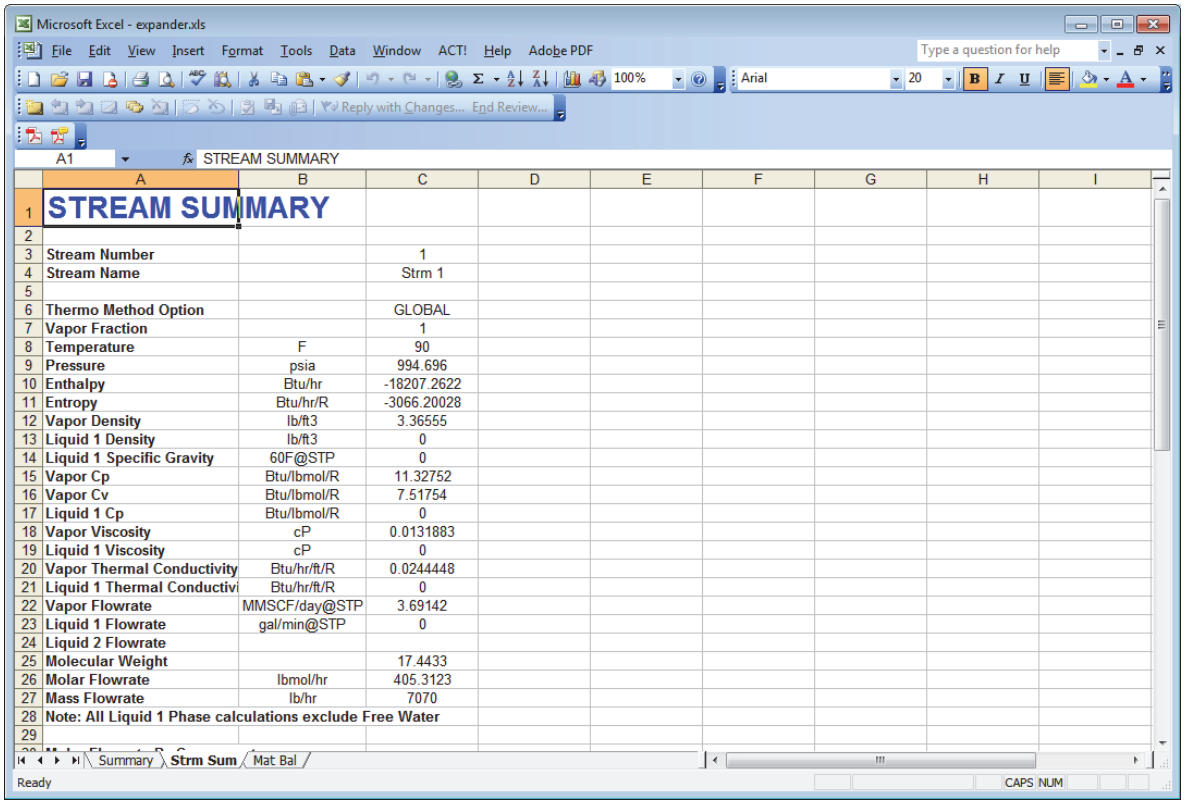


Figure 1 - 26: Results for Stream 1 in Microsoft Excel (from expander.psd)

5. For a brief review of equipment results, simply click on the **Equipment** icon in the **Toolbox**.
6. Right click on an equipment symbol in the drawing, such as Exchanger E-1 in the Expander plant drawing and click on the **View Results** button from the popup menu.

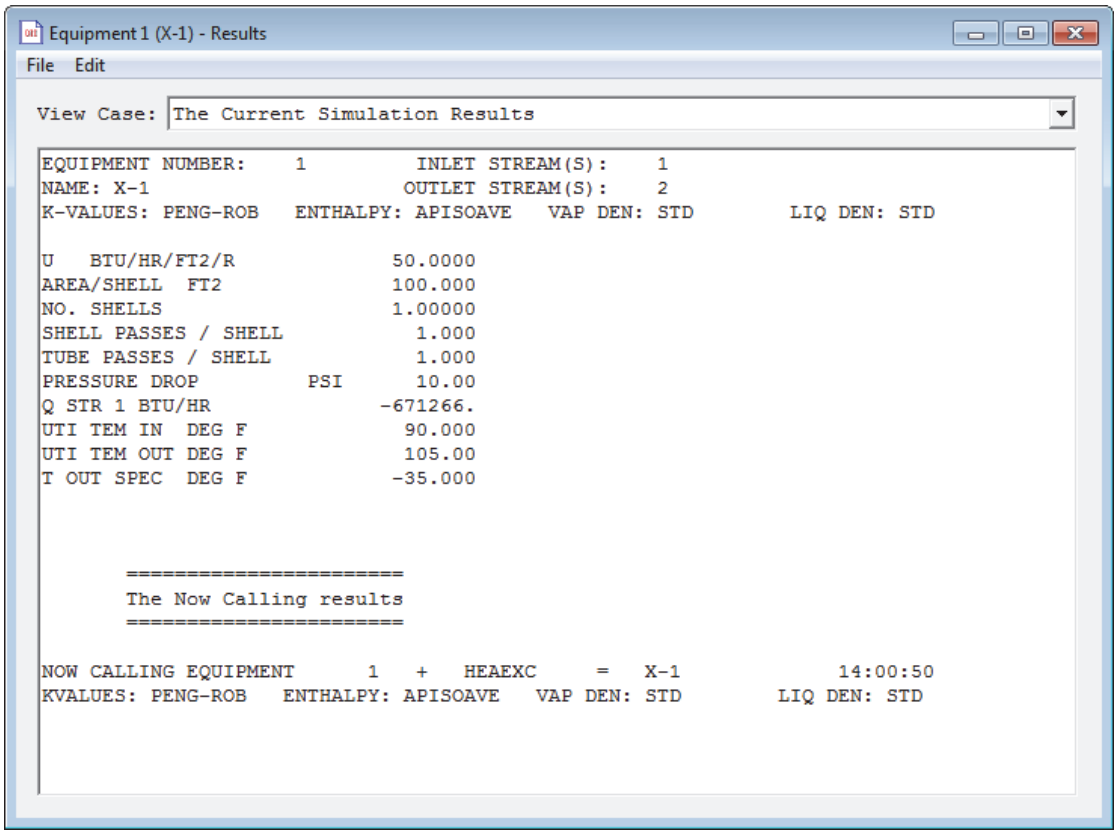


Figure 1 - 27: View Results for Exchanger 1 (from expander.psd)

Changing the PSD File

Making another run is easy. If you are working with an existing PSD file, follow the steps below.

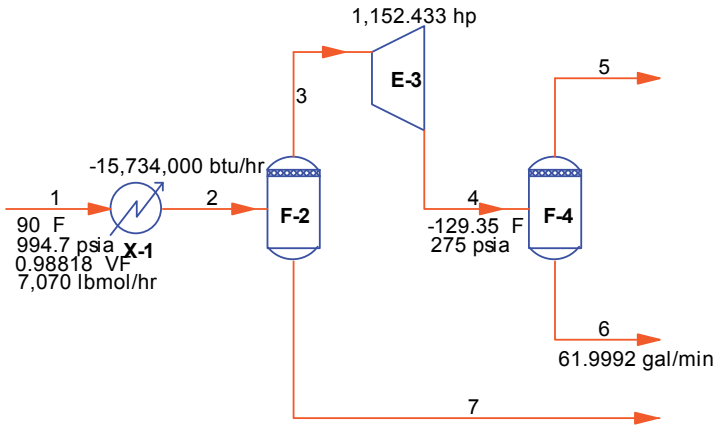
1. Make the desired changes in the equipment and stream specifications.
2. Save the PSD file under a different name unless you are sure you do not want the previous results. Click on the **File** menu and select **Save As**.
3. A dialog will appear on-screen displaying the PSD names for the current directory. Type a new name in the text box.
4. Click on the **Simulate** menu and click on **Execute**.
5. When the simulation is complete, review the results as discussed above.

Starting a new flowsheet is just as easy.

1. Close any Simulation files which may be open.
2. Click on the **File** menu and select **New**.
3. You will be prompted for saving changes to the current psd file. You may click on **Yes** to save changes; **Cancel** to return to the current PSD; or **No** to throw away changes since the last Save operation.
4. If you clicked on Yes or No, a dialog will appear on-screen which contains choices for paper size and orientation and drawing dimensional units preferences. Select the desired paper size and dimensional units, then click on **OK**.
5. The next dialog will request a name for the new PSD file. Type in a name and click on **OK** or press **Enter**.
6. Begin adding equipment symbols and streams to your drawing just as we did at the beginning of the tutorial.
7. Add the directional arrows. Add specifications for the equipment modules, then the feed stream(s). Enter specifications for any recycle streams at this time.
8. Click on the **File** menu and select **Save** before you simulate the flowsheet.
9. Now begin the simulation. Click on the **Simulate** menu and select **Execute**.
10. Review the results file when the simulation is completed.

Chapter 2: Samples - c:\designii

Expander.psd - Expander Plant Example

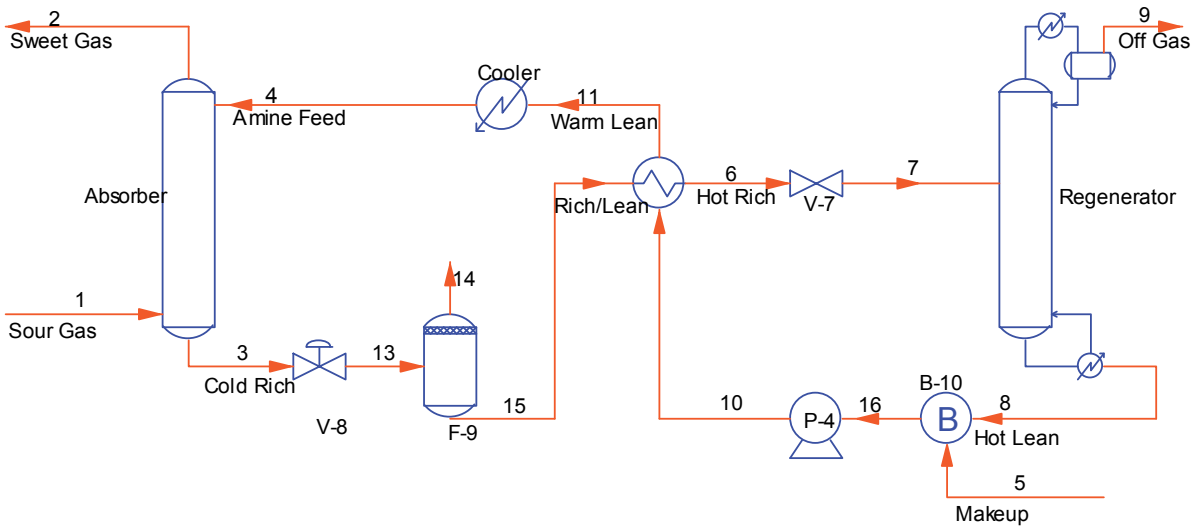


The purpose of this simulation is to cool the feed stream (90 F at 980 psig) to -35 F, separate the resulting stream into two phases, and then expand the vapor phase to 275 psia. The feed contains 6100 lbmol/hr of methane, 500 lbmol/hr of ethane, 200 lbmol/hr of propane, 200 lbmol/hr of butane, 200 lbmol/hr of pentane and 70 lbmol/hr of hexane.

```
AB123.
*EXPANDER PLANT PROBLEM
C- EQUIPMENT MODULES

HEA EXC 1 = X-1, 1,-2,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = -35
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 10

FLA 2 = F-2, 2,-3,-7,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

EXP 3 = E-3, 3,-4,
  PRE OUT(PSIA) = 275
  EFF = 0.8

FLA 4 = F-4, 4,-5,-6,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,6,8,10,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'N-BUTANE'
  NAM 8 = 'N-PENTANE'
  NAM 10 = 'N-HEXANE'
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
  THE CON VAP = NBS81
```

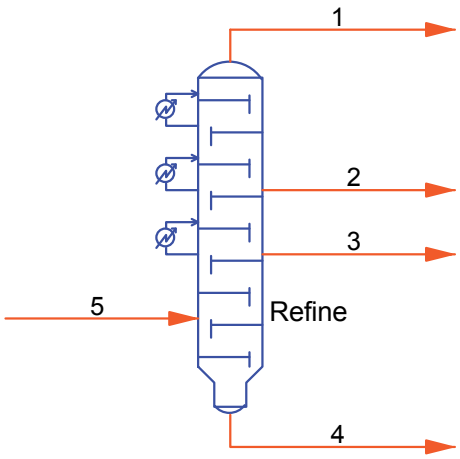
```
  THE CON LIQ = NBS81
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  BULK = 1,
  FLO(LBMOL/HR)1 = 6100,500,200,2*100,70,
  TP (F,PSIG) 1 = 90,980
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

MxpInt1.psd - Mixed Amine Plant for CO2 / H2S Removal



The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA).

Refi_ex3.psd - Pumparound Internal Reflux without Condenser (PUM REF)



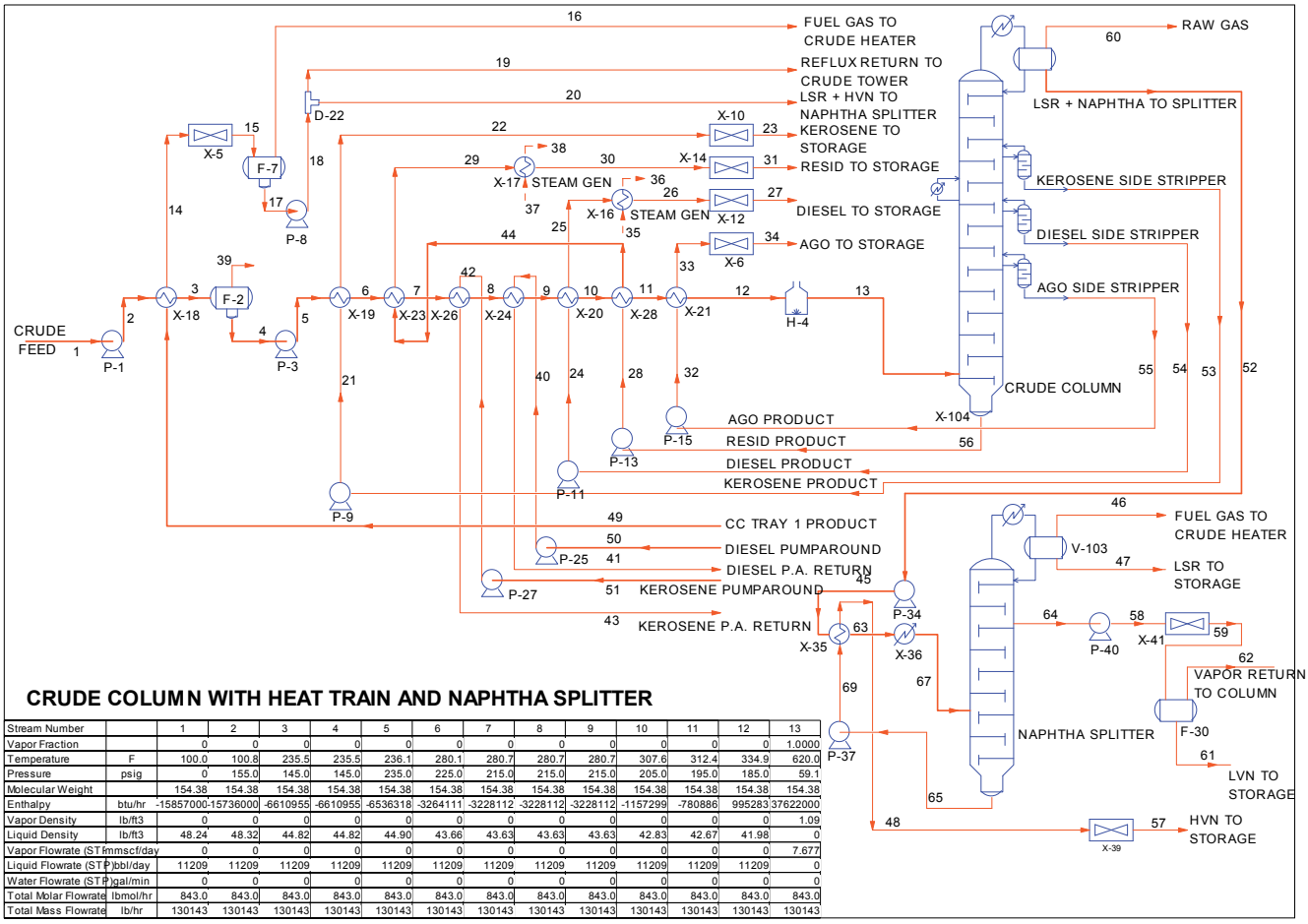
This simulation shows an internally refluxed column with no overhead condenser to provide reflux. Instead, a pumparound on tray 1 generates internal reflux. With this column type, there is no automatic decant of liquid water.

```
AB123.
```

```
*Refine Example 3: Pumparound Internal Reflux Without Condenser

C- EQUIPMENT MODULES

REFI 1 = Refine, 5,-1,-2,-3,-4,
  PUM REF
  STA = 8,
  TEM TOP(F) = 250,
  TEM BOT(F) = 600,
  PRE TOP(MMHG) = 75,
  PRE BOT(MMHG) = 100,
  LOC FEE = 7,
  PRO(BBL/DAY) = 0,461,4142,0,
  PRO GUE(LBMOL/HR) = 3,28,221,15,
  LOC STE = 9,
  PRE STE(PSIA) = 50,
  TEM STE(F) = 650,
  STE(LB/HR) = 1500,
  LOC PUM = 2,1,4,3,6,-1,
  PRO PUM(BBL/DAY) = 6000,19300,933,
  PRO PUM GUE(LBMOL/HR) = 460,1038,44,
  HEA PUM(BTU/HR) = 1000000,15850000,50000,
  PRO PUM CUR = 1,1,1,
  PRO NAM = TOP VAP,Side1,Side2,BOTTOM,
  PRO CUR = 0,0,0,0
  TEM PRO CUR(F) = 100,0,100,0,100,0,100,0,
  PRE PRO CUR(PSIA) = 0,0,0,0,0,0,0,0,
  LOC SID = 2,4,
  FIX TEM(F) 1 = 160,1,1
  FIX LIQ(BBL/DAY) 2 = 933,6,H1
  FIX GUE(LBMOL/HR) = 44,
  TOL = 0.0001
  MAX MAT = 3
  REP = 0,1,0,0,0,0,
  SWI = 1
  GLI SYS = 1
  GLI VAL MAT = CARB

C- CRUDE FEED CHARACTERIZATION
CRUDE
  FEE BLEND  5 = 1
  FEE VOL  5 = 2,5,10,20,30,40,50,60,70,80,90,95,98,
  FEE AST (F) 5 = 471,530,584,651,685,714,741,771,802,855,943,977,1059,
  FEE (BBL/DAY) 5 = 5000
  FEE GRA (API) 5 = 1.7
  FEE NAME  5 = Strm 5


C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,
  NAM 62 = 'WATER'
  MOD ESSO
  APIMOD
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  REI PRO = 4,
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  TP (F,PSIA) 5 = 650,95
  NAM STR 5 = Strm 5
API ENTHALPY
HEAT CURVE 1
END
```

Refinery.psd – Crude Column with Heat Train and Naphtha Splitter

This is a general purpose flowsheet for a typical refinery operation. Shows a combined simulation of a pre-heat train and light and heavy naptha refine columns.



CRUDE COLUMN WITH HEAT TRAIN AND NAPHTHA SPLITTER

Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13
Vapor Fraction		0	0	0	0	0	0	0	0	0	0	0	0	1.0000
Temperature	F	100.0	100.8	235.5	235.5	236.1	280.1	280.7	280.7	280.7	307.6	312.4	334.9	620.0
Pressure	psig	0	155.0	145.0	145.0	235.0	225.0	215.0	215.0	215.0	205.0	195.0	185.0	59.1
Molecular Weight		154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38
Enthalpy	btu/hr	-15857000	-15736000	-6610955	-6610955	-6536318	-3264111	-3228112	-3228112	-3228112	-1157299	-780886	995283	37622000
Vapor Density	lb/ft3	0	0	0	0	0	0	0	0	0	0	0	0	1.09
Liquid Density	lb/ft3	48.24	48.32	44.82	44.82	44.90	43.66	43.63	43.63	43.63	42.83	42.67	41.98	0
Vapor Flowrate (ST	mmscf/day	0	0	0	0	0	0	0	0	0	0	0	0	7.677
Liquid Flowrate (STP)	bbl/day	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	0
Water Flowrate (STP)	gal/min	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Molar Flowrate	lbmol/hr	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0
Total Mass Flowrate	lb/hr	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143

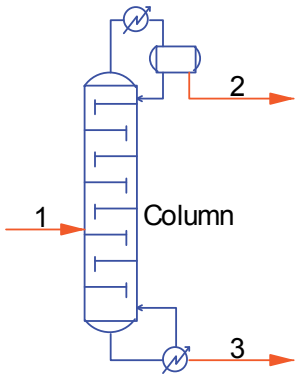
Chapter 3: Samples - c:\designii\samples

Agacomp.in – Polytropic Compressor Using AGA

This simulation shows a polytropic compressor simulation using the AGA thermodynamic method.

Decant.psd - Column with Water Decants

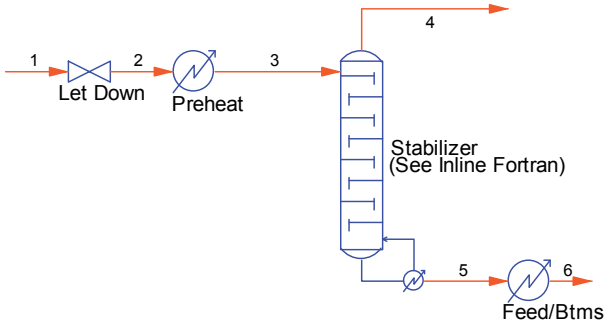
This simulation example is to illustrate the water decants at five trays including the condenser.




```
DIS 1 = Column , 1,-2,-3,
  TOT
  TRA = 20
  TEM TOP(F) = 115
  TEM BOT(F) = 220
  PRE TOP(PSIA) = 240
  PRE BOT(PSIA) = 246
  LOC FEE = 10,
  SUP
  PRO (LBMOL/HR) = 49.3,0.
  TOL = 1e-005
  REF = 3.17
  PRE CON(PSIA) = 235
  REB VAP GUE(LBMOL/HR) = 160
  GLI SYS = 1
  GLI VAL MAT = CARB
  LOC DEC = 0,1,2,3,10,
  DEC COM = 62, IMM = 62
  VAP PRO(LBMOL/HR) = 0.001,150,17*0,155,
  PRI SWI = 2
```

Disrvp.psd - Inline Fortran To Meet Reid Vapor Pressure Specification.

This example creates a Reid vapor pressure specification for the DIStillation column using Inline Fortran.



```
VAL 1 = Let Down, 1,-2,
  PRE OUT(PSIA) = 120

HEA EXC 2 = Preheat, 2,-3,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 120
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5

DIS 3 = Stabilizer, 3,-4,-5,
  ABS REB
  TRA = 12
  TEM TOP(F) = 140
  TEM BOT(F) = 275
  PRE TOP(PSIA) = 105
  PRE BOT(PSIA) = 110
  LOC FEE = 1,
  SUP
  PRO (LBMOL/HR) = 0.,116
  TOL = 1e-005
  PUR BOT (MOL, REL) 2 = 0.015
  GLI SYS = 1
  GLI VAL MAT = CARB
  PRI SWI = 2
C-
C-  MAIN COLUMN SPECIFICATION
C-
C-  THE PURITY SPECIFICATION FOR METHANE IN THE
C-  BOTTOMS PRODUCT WILL BE REPLACED BY THE
C-  INLINE FORTRAN EPS(1) SPECIFICATION WHICH
C-  WILL REQUEST A BOTTOMS PRODUCT RVP OF 10.0
C-  USING API TECHNICAL DATA BOOK PROCEDURE 5B1.2
F-      DURING
F-      CALL REICRU(FLOWL(1,1),RVP,IC,ERR)
F-      EPS(1)=(RVP - 10.)/10.

HEA EXC 4 = Feed/Btms, 5,-6,
  DUT(BTU/HR) = -1500000
  U(BTU/HR/FT2/F) = 50
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5
C-  INLINE FORTRAN COMMANDS WILL RETRIEVE
C-  CALCULATED DUTY FROM FEED PREHEAT
C-  EXCHANGER AND SET THE SPECIFICATION FOR
C-  THE FEED/BOTTOMS EXCHANGER
F-      PRE
F-      FPREHT = GET(HEAEXC(2),CAL DUT)
F-      SET (HEAEXC(4),DUT) = - FPREHT
```

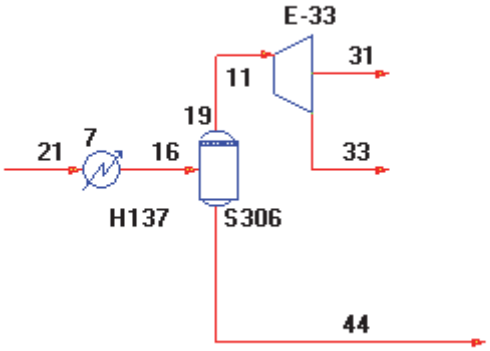
Emulsion.in – Line Pressure for Emulsion Mixture

This is a two-phase line pressure drop calculations for an oil in water emulsion mixture.

Example1.psd - Expander Plant Example

The purpose of this simulation is to cool the feed stream (100 F at 800 PSIA) to -40 F, separate the resulting stream into two phases, and then expand the vapor phase to 300 PSIA. The feed contains 10 LBMOL/HR of methane, 2 LBMOL/HR of ethane, and 1 LBMOL/HR of propane.

```
AB123.
*EXPANDER PLANT PROBLEM
C- HEAT EXCHANGER MODULE:    7   (H137)
HEAT EXCHANGER 7 = H137, 21,-16,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE OUT = -40,
C- OPTIONAL COMMANDS
DELTA PRESSURE = 10,

C- PHASE SEPARATOR: 19 (S306)
FLASH 19 = S306,16,-11,-44,
        ISOTHERMAL FLASH

C- EXPANDER: 37 (E33)

EXPANDER 37 = E33, 11,-31,-33,
        PRESSURE OUT = 300,

GENERAL
         COMPONENTS = 2,3,4,
C- THERMO PROPERTY OPTIONS
          KVAL,   COPE,

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE ID C- NUMBERS: 21 (STREAM 21)
         TP 21 = 100,   800,
         FLOW 21 = 10,2,1,
END
```



Example2.psd - Debutanizer Design Example

This simulation is a rigorous calculation of the debutanizer design problem presented in Flowsheet Procedure: Example Flowsheet Simulation-Debutanizer Column Design. For this version the rigorous distillation (DIS) Equipment Module is used, rather than the Shortcut fractionator (SHO).

From the results of the shortcut run, it is determined that a 23-tray column with feed on tray 9 and operating at a reflux ratio of 0.9 can be used. The overhead product rate guess is 2996 LBMOL/HR.

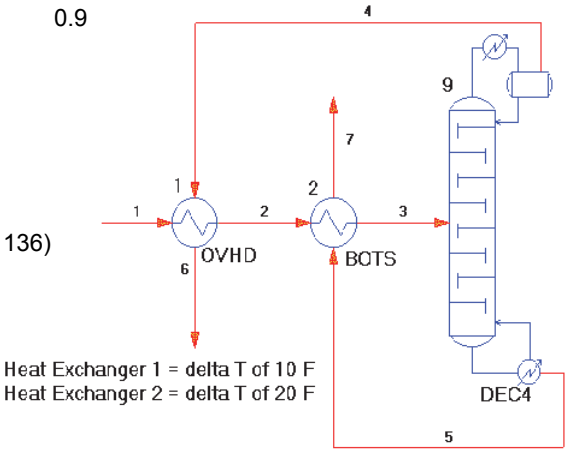
NOTE: The recovery specifications automatically invoke the SUPer convergence routine. Notice that the recovery specifications for DIS are rates, not fractions of total feed.

Feed Stream 1 **Estimate for recycle stream 3**
Temperature of 80 F Temperature F
Pressure 210 PSIA

Flowrate	**LBMOL/HR**	**Estimates for column calculation**	
C3	2000	Top product temperature	143 F
i-C4	400	Bottom product temperature	344F
n-C4	600	Top product rate	2996 LBMOL/HR
i-C5	150.3	Reflux ratio guess	0.9
n-C5	200		
C6	50.8		
C 7+	250		

Specifications for column calculation

Overhead recovery of n-C4 588 LBMOL/HR (98%)
Bottom recovery of i-C5 142.8 LBMOL/HR (95%)
(API gravity = 40, mean average boiling point = 350 F, molecular weight = 136)




```
AB123.
*DEBUTANIZER DESIGN

C- HEAT EXCHANGER MODULE:    1   (OVHD)
HEAT EXCHANGER 1 = OVHD, 4,1,-6,-2,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE APPROACH = 10,
```

```
C- HEAT EXCHANGER MODULE:    2  (BOTS)
HEAT EXCHANGER 2 = BOTS, 5,2,-7,-3,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE APPROACH = 20,
C- PARTIAL FRACTIONATOR: 3 (DEC4)
DIS 3 = DEC4,3,-4,-5,
        PARTIAL FRACTIONATOR,
        TRAYS = 23,
        PRESSURE TOP = 200,
        TEMPERATURE TOP = 143,  TEMP BOTTOM = 344,
        LOCATION OF FEEDS = 9,
        REFLUX GUESS = 0.9,
        PRODUCT FLOW RATES = 2996,0,
        RECOVERY IN TOP 6 = 588,
        RECOVERY IN BOT 7 = 142.8,
GENERAL
        COMPONENTS = 4,5,6,7,8,10,100,
C- FLOWSHEET OPTIONS
        MAXIMUM ITERATIONS = 5,
        CONVERGENCE TOLERANCE = 0.01,
        RECYCLE STREAMS = 3,
        CONVERGE TEMPERATURE OF STREAM = 3,
        AMB 100 = 350,  API (API) 100 = 40,
        AMW 100 = 136, NAM 100 = C7+

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE ID C- NUMBERS: 1 (STREAM 1)
        TP 1 = 80,  210,
        FLOW 1 = 2000,400,600,150.3,200,50.8,250,

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE ID C- NUMBERS: 3 (STREAM 3)
        TP 3 = 154,  200,
        FLOW 3 = FLOW 1,
END
```

Example3.psd - Propane-Propylene Separation Example

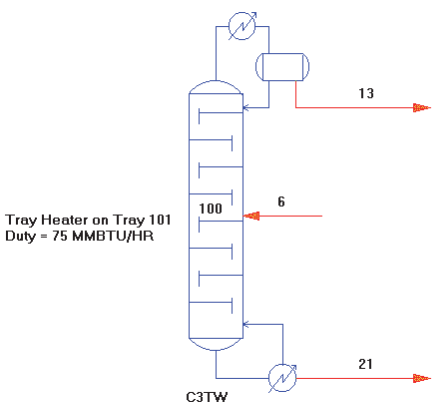
The objective of this simulation is to extract 99.6 mole-percent of pure propylene from a propane-propylene mixture. The simulation uses the rigorous distillation (DIS) module.

Guesses
Top product rate = 1480 LBMOL/HR
Top product temperature = 110 F Bottom product temperature = 125 F

Specifications
Reflux ratio = 22.5
Top product purity = 99.6% propylene

Problem
We want to extract 99.6 mole-percent pure propylene from a propane-propylene mixture.

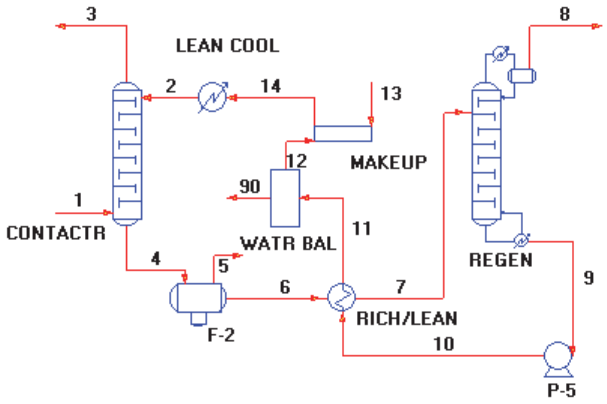
```
AB123.
*C3 SPLITTER

C- TOTAL FRACTIONATOR: 1(C3TW)
DIS 1 = C3TW,6,-13,-21,
        TOTAL FRACTIONATOR,
        TRAYS = 155,
        PRESSURE TOP = 264,  DELTA PRESSURE = 15,
        TEMPERATURE TOP = 110,  TEMP BOTTOM = 125,
        LOCATION OF FEEDS = 100,
        PRODUCT FLOW RATES = 1480,0,
        REFLUX RATIO = 22.5,
        PURITY IN TOP 23 = .996,
        PRINT SWITCH = 2,
GENeral Commands
GENERAL
        COMPONENTS = 4,23,
C- THERMO PROPERTY OPTIONS
        PENK,  PENH,
        BINARY PARAMETER = PENG1,
GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE ID
C- NUMBERS: 6 (STREAM 6)
        TP 6 = 120,  275,
```



```
        FLOW (LB/HR) 6 = 7724,62304,
END
```

Example4.psd - Gas Sweetening with MEA Example

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a monoethanolamine solution. Residual H2S in the sweet gas stream is to be no greater than 1/4 grains of H2S per 100 standard cubic feet (SCF) of gas. The reboiler and condenser duties will also be determined for the amine regenerator.



Sour Gas	LBMOL/HR	Assumptions
C1	5351.87	Use 15 weight % solution of MEA.
CO2	125.7	Amine flow rate is based on rich amine
H2S	31.32	loading of 0.45 moles of acid gas
H2O	0	per mole of MEA and13% residual acid gas in lean amine.
MEA	0	
Total	5518.95	

Objectives

0.25 grains H2S per 100 SCFof sweet gas, 120 F condenser temperature

Specification:

Reflux ratio = 2.5

Problem:

Determine the Lean Amine Flow rate. Acid gas in feed is 157.08 lbmole/hr.

1. (157.08 + residual acid gas) / lbmoles of MEA = 0.45
2. Residual Acid Gas = 0.13 X MEA flow
3. (0.32 MEA = 157.08) = 490.875 lbmoles/hr MEA
4. Molecular weight for MEA is 61.0861. Mass flow for amine = 61.0861 X 490.875 lbmoles/hr = 29985.639 lb/hr MEA
5. For a 15 wt% MEA total solution = 29985.639 / 0.15 = 199904.26 lb/hr of solution.
6. Molar flow of water = (199904.26 - 29985.639) = 169918.62 lb/hr./ 18.015 lb/lbmol = 9432.06 lbmol/hr of water.
7. Add 63.81 (0.13 X 490.875) moles of Residual Acid Gas to Lean Amine Solution.

The regenerator calculation will be most stable if you specify RECovery TOP for water, which will control the top product temperature. To calculate the correct amount of water in the overhead product, use the following steps:

1. Determine the desired condenser temperature (120 F in this case).
2. Find the partial pressure of water at 120 F from the steam tables (.16945 PSIA).
3. Divide the water partial pressure by the pressure of the condenser (20 PSIA in this example). The resultant value (.0857) is the mole fraction for water in the overhead product.
4. Estimate the amount of acid gas (and light hydrocarbons) in the overhead product. Unless you have a separator (FLAsh or VALve) removing some of the acid gas between the contactor and regenerator, this will simply be the CO2and/or H2S plus light hydrocarbons picked up by the lean amine steam. Divide this value by (1- water mole fraction) to get the total product rate overhead. Subtract the acid gas flow from the product rate to get the water flow rate. For this case we will estimate an acid gas plus light hydrocarbon flow rate of 161 lbmol/hr.

   Overhead Product of: 161/0.9153 = 175.9 lbmol/hr
   Water Overhead of: 175.9 - 161 = 14.9 lbmol/hr

```
AB123.
*SAMPLE ACID GAS STIPPING/ABSORPTION (MEA)

C- EQUIPMENT MODULES
```

```
DIS 1 = Contactor, 15,1,-3,-2,
  ABS
  TRA = 6
  TEM TOP(F) = 110.
  TEM BOT(F) = 140.
  PRE TOP(PSIA) = 914.7
  DEL(PSI) = 0.
  LOC FEE = 1,7,
  PRO (LBMOL/HR) = 0.,10146.
  COM CON = 30,3
  TOL = 1.E-006
  GLI SYS = 1.
  GLI VAL MAT = CARB
C-
C-  ADD COMMANDS TO STABILIZE COLUMN CALCULATIONS
C-  AND SPEED RECYCLE CONVERGENCE
C-  TIGHTENING COLUMN CONVERGENCE REDUCES CHANGES
C-  IN COMPOSITION AND TEMPERATURE ACROSS ITERATIONS
C-
  SAVE PROFILES
  PRI SWI=-1,SMI SIZ OFF
C-

FLA 2 = Flash, 2,-4,-5,
  ADI, HEA(BTU/HR) = 0.
  PRE OUT(PSIA) = 22.

HEA EXC 3 = Rich/Lean, 5,9,-6,-10,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 220.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,

DIS 4 = Regenerator, 6,-7,-8,
  PAR
  TRA = 6
  TEM TOP(F) = 230.
  TEM BOT(F) = 260.
  PRE TOP(PSIA) = 22.
  DEL(PSI) = 0.
  LOC FEE = 2,
  PRO (LBMOL/HR) = 176.,0.
  SUP
  COM CON = 10,2
  DVS = 0.4
  TOL = 1.E-006
  REF = 2.5
  PRE CON(PSIA) = 20.
  REC TOP (LBMOL/HR, REL) 62 = 14.9
  GLI SYS = 1.
  GLI VAL MAT = CARB
C-
C-  RECOVERY TOP OF WATER SETS CONDENSER TEMPERATURE
C-  AND STABILIZES THE COLUMN CALCULATIONS
C-
SAVE PROFILES
PRI SWI=-1,SMI SIZ OFF
C-

PUM 5 = Pump, 8,-9,
  PRE OUT(PSIA) = 914.7
  EFF = 0.9
  ELE DRI

COM SPL 6 = Water Bal., 10,-12,-11,
  REC TOP(FRA) = 3*1.,0.,1.,

MIX 7 = Make-up, 13,12,-14,

HEA EXC 8 = COOL, 14,-15,
  U(BTU/HR/FT2/F) = 50.
```

```
  TEM OUT(F) = 110.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,49,50,62,4155,
  MDEAK
  MDEAH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 8
  CON TOL = 5.E-003
  REC STR = 15,9,
  FLO(LBMOL/HR)1 = 5351.87,125.76,31.32,2*0.,
  TP (F,PSIA) 1 = 90.,914.7
  NAM STR 1 = Acid Gas
  NAM STR 2 = Strm 2
  NAM STR 3 = Methane
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = CO2
  NAM STR 8 = Strm 8
  FLO(LBMOL/HR)9 = 0.,62.4,1.41,9432.06,490.875,
  TP (F,PSIA) 9 = 235.,914.7
  NAM STR 9 = Recycle
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  FLO(LBMOL/HR)13 = 3*0.,9431.1,0.,
  TP (F,PSIA) 13 = 236.26,914.7
  NAM STR 13 = Water
  NAM STR 14 = Strm 14
  FLO(LBMOL/HR)15 = 0.,62.4,1.41,9432.06,490.875,
  TP (F,PSIA) 15 = 110.,914.7
  NAM STR 15 = MEA
END
```

NOTE: The water makeup rate will be the difference between streams 13 and 90.Check the results of the regeneration column calculation to see if the bottoms product has the expected residual moles of acid gas/moles of amine.

NOTE: A typical residue for MEA is an acid gas/amine ratio of 10-12%; for DEA the typical residue is 5-8%. The higher the reflux ratio or reboiler duty, the cleaner the lean amine will be. Neither MEA nor DEA is ever regenerated completely.

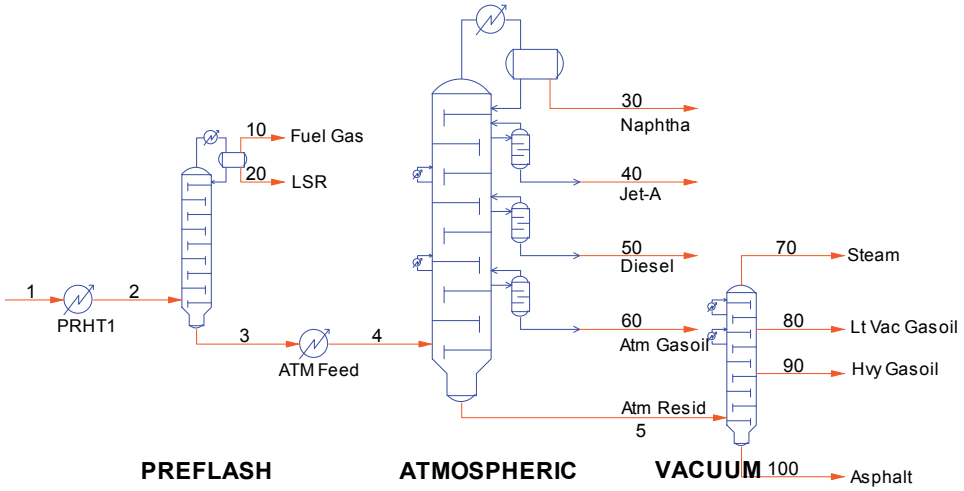
Also check to be sure the correct number of moles of CO2 and H2S are leaving in the overhead product. If not, the initial guess for CO2 and H2S in the lean amine feed to the contactor was probably incorrect. CO2 and H2S are not equally attracted to or rejected from amines. CO2 absorption is favored by both equilibria and kinetics.

Some physical properties for monoethanolamine, diethanolamine and water are shown below.

	MEA	DEA	Water
Molecular Weight	61.08	105.14	18.015
Boiling point @ 760mm Hg,F	338.9	516.2	212
Freezing point, F	50.9	82.4	2
Density @ 20C, gm/cc	1.018	1.095	0.998
Weight, lb/gal @ 60 F	8.48	9.09	8.33

Example5.psd - Crude Unit Example

The purpose of this simulation is to determine the quality of products which can be made from a series of crude distillation columns for a feed of 100000 barrels per day. The amount of duty required for feed preheat (including feed furnaces for the atmospheric and vacuum columns) will also be calculated.



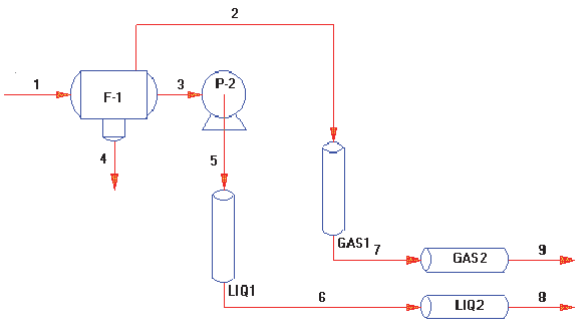

```
AB123.
*CRUDE UNITS -PREFLASH; ATMOSPHERIC AND VACUUM

C- HEAT EXCHANGER MODULE:    1  (PRHT)

HEAT EXCHANGER 1 = PRHT, 1,-2,
C- HEAT EXCHANGER SPECIFICATION
         TEMPERATURE OUT = 450,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 10,
C- REFINE CRUDE TOWER 2  (PRFL)

REFINE 2 = PRFL, 2,-10,-20,-3,

C-      PARTIAL CONDENSER
        PARTIAL CONDENSER
        STAGES = 8


        LOCATION OF FEEDS = 7,
        PRESSURE OF TOP = 41.7
        TEMPERATURE OF TOP TRAY = 340,
        TEMPERATURE OF BOTTOM = 450,
        PRODUCT RATES (BBL/DAY) = 0,0,80000,
        PRODUCT RATE GUESS = 280,1950,3900,
        TEMPERATURE OF CONDENSER = 160,

C-      PRESSURE DROPS FOR MAIN COLUMN AND
C-      SIDE STRIPPERS
        DELTA PRESSURE = 3,

C-      CONDENSER DESCRIPTION
        PRESSURE OF CONDENSER = 39.7,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 8,
        TEMPERATURE OF STEAM = 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 5000,

C-      COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = FUEL GAS,LSR,ATM.FEED,
C- HEAT EXCHANGER MODULE:    3  (PRHT 2)
HEAT EXCHANGER 3 = PRHT 2, 3,-4,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE OUT = 625,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 15,

C- REFINE CRUDE TOWER 4  (ATM)
REFINE4 = ATM, 4,-30,-40,-50,-60,-5,
C-      TOTAL CONDENSER
        TOTAL CONDENSER
        STAGES = 24,4,3,2,
        LOCATION OF FEEDS = 21,
```

```
        PRESSURE OF TOP = 20.7
        TEMPERATURE OF TOP TRAY = 360,
        TEMPERATURE OF BOTTOM = 620,
        PRODUCT RATES (BBL/DAY) = 0,35410,
        PRODUCT RATE GUESS = 730,1050,

C-      PRESSURE DROPS FOR MAIN COLUMN AND SIDE C-    STRIPPERS
        DELTA PRESSURE = 4,

C-      CONDENSER DESCRIPTION
        PRESSURE OF CONDENSER = 15.7,

C-      SIDE STRIPPER DESCRIPTION AND
C-      SPECIFICATION
        LOCATION OF DRAW TRAYS = 5,12,17
        LOCATION OF VAPOR RETURN = 5,12,17

C-      SIDE STRIPPER SPECIFICATIONS
        PRODUCT RATES FROM STRIPPER (BBL/DAY) = 8165,
        16967,10000,
        PRODUCT FROM STRIPPER GUESS = 615,1050,470,

C-      PUMPAROUND DESCRIPTIONS
        LOCATION OF PUMPAROUND = 7,5,13,12,
C-      PUMPAROUND SPECIFICATIONS
        PRODUCT OF PUMPAROUND (BBL/DAY) = 49000,11000,
        HEAT FROM PUMPAROUND = 40E6,15E6,
C-      PUMPAROUND MOLAR RATE ESTIMATES
        PRODUCT OF PUMPAROUND GUESS = 3600,700,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 25,29,32,34,
        TEMPERATURE OF STEAM = 4*400,
        PRESSURE OF STEAM = 4*60,
        STEAM RATES (LB/HR) = 12000,3300,1000,800,

C-    PRODUCT PROPERTY SPECIFICATIONS
        PRODUCT SPEC 1 = 625, ASTM95PT , PROD3, PROD4,

C- TRAY FIX COMMANDS
                FIX LIQUID (BBL/DAY) 1 = 1500, 20, H1,
                FIX GUESS = 50,

C-      COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = NAPHTHA,JET,DIESEL,AGO,RESID,
C- REFINE CRUDE TOWER 15  (VTWR)

REFINE 15 = VTWR, 5,-70,-80,-90,-100,

C-      PUMPAROUND INTERNAL REFLUX
        PUMREF
        STAGES = 6
        LOCATION OF FEEDS = 6,
        PRESSURE OF TOP (MMHG) = 75
        TEMPERATURE OF TOP TRAY = 450,
        TEMPERATURE OF BOTTOM = 700,

C-     SIDE DRAW DESCRIPTION
        LOCATION OF SIDE DRAWS = 2, 4,
        PRODUCT RATES (BBL/DAY) = 0,6200,18600,10600,
        PRODUCT RATE GUESS = 1.2,269.4,592.3,195.2,

C-      PRESSURE DROPS FOR MAIN COLUMN AND
C-      SIDE STRIPPERS
        DELTA PRESSURE (MMHG) = 30,

C-      PUMPAROUND DESCRIPTIONS
        LOCATION OF PUMPAROUND = 2,1,4,3,
C-      PUMPAROUND SPECIFICATIONS
        PRODUCT OF PUMPAROUND (BBL/DAY) = 20000,
        49000,
        HEAT FROM PUMPAROUND = 14E6,54E6,
C-      PUMPAROUND MOLAR RATE ESTIMATES
        PRODUCT OF PUMPAROUND GUESS = 870,1560,
```

```
C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 7,
        TEMPERATURE OF STEAM= 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 20000,

C- TRAY FIX COMMANDS
        FIX TEMP 1 = 150, 1, 1,
        FIX LIQ NET 2 = 35, 2, 3,
        FIX LIQ NET 3 = 2, 5, H1,

C-     COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = STEAM,LVGO,HVGO,ASPHALT,

GENERAL
        COMPONENTS = 62,2,3,4,5,6,7,8,
C- THERMO PROPERTY OPTIONS
        ESSOTAB,  API,  STDD,
        VISCOSITY OPTION = STD,
        THERMAL CONDUCTIVITY OPTION = NBS81,
        PETROLEUM STREAMS =     10,20,30,40,
        50,60,70,80,90,100,

CRUDE PROPERTY SECTION
C- DATA FOR PSEUDO COMPONENTS WITHOUT C- ID NUMBERS (CUTS)
        TEMPERATURE INCREMENTS FOR TBP CURVES = 30,
        60,120,
        TEMPERATURE BREAKS FOR TBP CURVES (F) = 500,900,
CRUDE STREAM
C- BULK STREAM - PSEUDO COMPONENTS WITHOUT ID C- NUMBERS: 1 (STREAM 1)
        FEED NAME 1 = STREAM 1,
        TP 1 = 215,59.7,
        FEED DRY RATE (BBL/DAY) 1 = 1000000,
        FEED REAL PERCENTS 1 = .1,.2,.5,.5,1,1.5,2.5,
        FEED TBP 1 = -50,100,200,300,400,470,550,
                        650,750,850,1100,1300,1475,1670,

        FEED VOLUME PERCENTS 1 = 2,5,10,20,30,40,
                50,60,70,80,90,95,98,100,

        FEED GRAVITIES (API) 1 = 150,95,65,45,40,
                38,33,30,25,20,15,10,5,

        FEED GRAVITY VOLUME PERCENTS 1 = 2,5,10,20,
                30,40,50,60,70,80,90,95,98,

END
```

Example6.psd - Offshore Separation Example

This simulation will model the last stage of separation and compression from an offshore platform with two single-phase pipelines to shore. The water knockout drum will be sized.

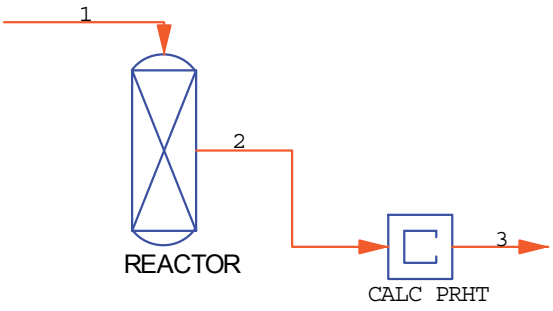



```
AB123.
*SOUTH PASS BLOCK 40 - OFFSHORE TRANSMISSION

C- PHASE SEPARATOR: 10 (KO3)
```

```
FLASH 10 = KO3,1,-2,-4,-5,
        ISOTHERMAL FLASH
C- SEPARATOR SIZING COMMANDS
        HORI,
        LENGTH SHELL = 10,
        BOOT DIAMETER = 2.5,

C- PUMP: 11 (PUMP)

PUMP 11 = PUMP,4,-7
        PRESSURE OUT = 1600,

C- LINE PRESSURE DROP: 13(LIQ1)

LINE 13 = LIQ1,7,-11,
        NOMINAL DIAMETER (IN) = 8,
        LENGTH OF LINE = 150,
        DOWN,
C- LINE PRESSURE DROP: 14(LIQ2)
LINE 14 = LIQ2,11,-12,
        NOMINAL DIAMETER (IN) = 8,
        LENGTH OF LINE (MILE) = 10*1,
        ELEVATION = 3*10,3*10,15,10,10,5,

C- LINE PRESSURE DROP: 15(GAS1)
LINE 15 = GAS1,2,-6,
        NOMINAL DIAMETER (IN) = 8,
        LENGTH OF LINE (FT) = 150,
        DOWN,

C- LINE PRESSURE DROP: 16(GAS2)
LINE 16 = GAS2,6,-10,
        NOMINAL DIAMETER (IN) = 8,
        LENGTH OF LINE (MILE) = 10*1,
        ELEVATION = 3*10,3*10,15,10,10,5,

GENERAL
        COMPONENTS =    46,49,2,3,4,5,6,7,8,10,11,
        12,13,14,62,
C- THERMO PROPERTY OPTIONS
        APISOAVEK,  APISOAVEH,
        PLOT PRE AND ELE VS LENGTH OF LINES = 13,14,
        PLOT PRE AND ELE VS LENGTH OF LINES = 15,16,

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE
C- ID NUMBERS: 1 (STREAM 1)
        TP 1 = 120,  1400,
        FLOW 1 =        5,3,6000,1500,500,350,500,115,160,
        115,210,105,130,120,500,
END
```

Example7.psd - Ammonia Synthesis Loop (Step 1)

The purpose of this simulation is to model an ammonia synthesis loop. The amount of heat to be added to the feed and the rate of feed to each bed of the reactor need to be determined. The feed preheat will be determined first using a CONTroller to vary the TEMperature OUT of REActor 1.



Feed:

Components	Flow (lb/mol/hr)
H2	650
N2	216

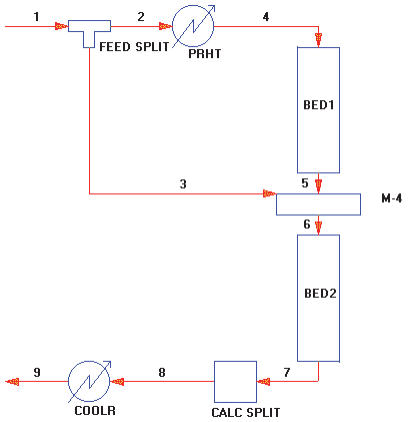
```
       NH3             34
       AR              30
       CH4             70
Temperature = 380 F
Pressure = 3400 PSIA

AB123.
*FEED PREHEAT REQUIRED FOR 14.98% NH3 PRODUCT

C- SPECIAL EQUIL. REACTOR: 1(REACTOR) @CODE 1 = Equipment Module Commands
 REACTOR 1 = REACTOR,1,-2
       ADIABATIC REACTOR,
       TEMPERATURE OUT = 925,  AMMONIA SYNTHESIS,
       NATIONAL DATA,
C-CONTROLLER (VARY) MODULE:  2 (HEAT)
CONTROLLER 2 = HEAT,2,-3,
       VARY TEM OUT(MIN=700,MAX=1000,STE=15) OF REAC 1
       UNT FLO COM 63 FROM STR2/FLO FROM STR2 = .1498 (TOL=.0002)
C-   GENeral Commands
GENERAL
        COMPONENTS = 1,46,63,1171,2,
C- THERMO PROPERTY OPTIONS
       PENK, PENH, PEND,

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE C- ID NUMBERS: 1 (STREAM 1)
       TP 1 = 380, 3400,
       FLOW 1 = 650,216,34,30,70,

END
```

The results from this calculation show a reactor duty of -2.055E6 BTU/HR (which means feed 1 is too cold). A duty of 2.055E6 will be set for HEAEXC2 in Step 2.

Example8.psd - Ammonia Synthesis Reactor (Step 2)

In Step 2, the problem is to define what feed split from DIVider 1 will result in the same feed temperature to beds 1 and 2 of the ammonia synthesis reactor. A CONTroller will be used to vary the FLOw RAT (FRAction) of DIVider 1.

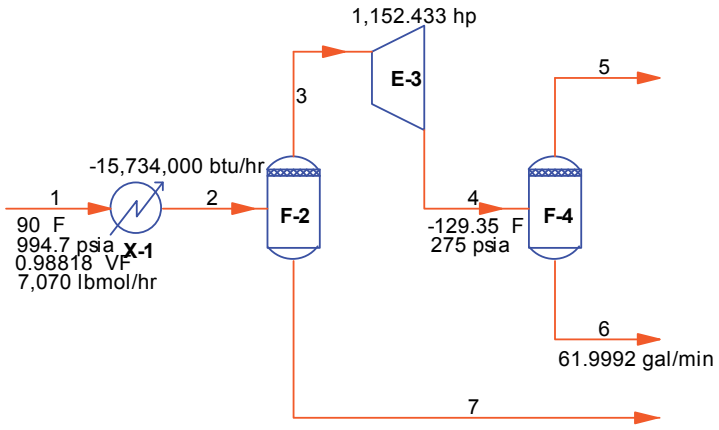
```
AB123.
*TWO-BED AMMONIA SYNTHESIS REACTOR
C- DIVIDER: 1 (SPLT)

Equipment Module Commands
DIVIDER 1 = SPLT,1,-2,-3,
       FLOW RATE = 0.7,

C- HEAT EXCHANGER MODULE:    2   (PRHT)

HEAT EXCHANGER 2 = PRHT, 2,-4,
C- HEAT EXCHANGER SPECIFICATION
       DUTY OF EXCHANGER = 2.055E6,

C- SPECIAL EQUIL. REACTOR: 3(BED 1)
REACTOR 3 = BED 1,4,-5
       ADIABATIC REACTOR,
       AMMONIA SYNTHESIS,
       NATIONAL DATA,
       DELTA PRESSURE = 20,

C- MIXER:4 (MIX)

MIXER 4 = MIX,5,3,-6

C- SPECIAL EQUIL. REACTOR: 5(BED 2)
REACTOR 5 = BED 2,6,-7
       ADIABATIC REACTOR,
       AMMONIA SYNTHESIS,
       NATIONAL DATA,
       DELTA PRESSURE = 20,

C-CONTROLLER (VARY) MODULE:  6 (SPLT)

CONTROLLER 6 = SPLT,7,-8,
       VARY FLO RAT(MIN=.01,MAX=.99,STE=.08) OF DIV1
       UNT TEM FROM STR4/TEM FROM STR6 =
```



```
       1.0(TOL=.00001)

C- HEAT EXCHANGER MODULE:    7   (COOL)
HEAT EXCHANGER 7 = COOL, 8,-9,
C- HEAT EXCHANGER SPECIFICATION
       DUTY OF EXCHANGER = -2.055E6,

GENERAL
       COMPONENTS = 1,46,63,1171,2,
C- THERMO PROPERTY OPTIONS
       PENK,   PENH,   PEND,

GENERAL
C- COMPONENT STREAM - ALL COMPONENTS HAVE
C- ID NUMBERS: 1 (STREAM 1)
       TP 1 = 380,   3400,
       FLOW 1 = 650,216,34,30,70,
END
```

Expander.psd - Expander Plant Example

The purpose of this simulation is to cool the feed stream (90 F at 980 psig) to -35 F, separate the resulting stream into two phases, and then expand the vapor phase to 275 psia. The feed contains 6100 lbmol/hr of methane, 500 lbmol/hr of ethane, 200 lbmol/hr of propane, 200 lbmol/hr of butane, 200 lbmol/hr of pentane and 70 lbmol/hr of hexane.




```
AB123.
*EXPANDER PLANT PROBLEM
C- EQUIPMENT MODULES

HEA EXC 1 = X-1, 1,-2,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = -35
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 10

FLA 2 = F-2, 2,-3,-7,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

EXP 3 = E-3, 3,-4,
  PRE OUT(PSIA) = 275
  EFF = 0.8

FLA 4 = F-4, 4,-5,-6,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

C- GENERAL SECTION
```

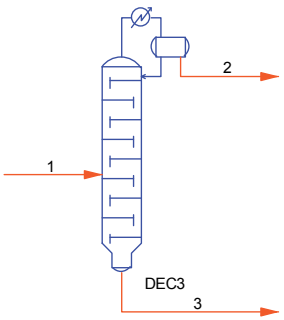
```
GENERAL
  COMPONENTS = 2,3,4,6,8,10,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'N-BUTANE'
  NAM 8 = 'N-PENTANE'
  NAM 10 = 'N-HEXANE'
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
  THE CON VAP = NBS81
  THE CON LIQ = NBS81
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  BULK = 1,
  FLO(LBMOL/HR)1 = 6100,500,200,2*100,70,
  TP (F,PSIG) 1 = 90,980
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Fmtraga.in - Flowmeter for Calculating Orifice Discharge

This is a flowmeter simulation with an API method for calculating the orifice discharge coefficient using AGA density option.

H2ochk.psd - Free Water Check On Refine

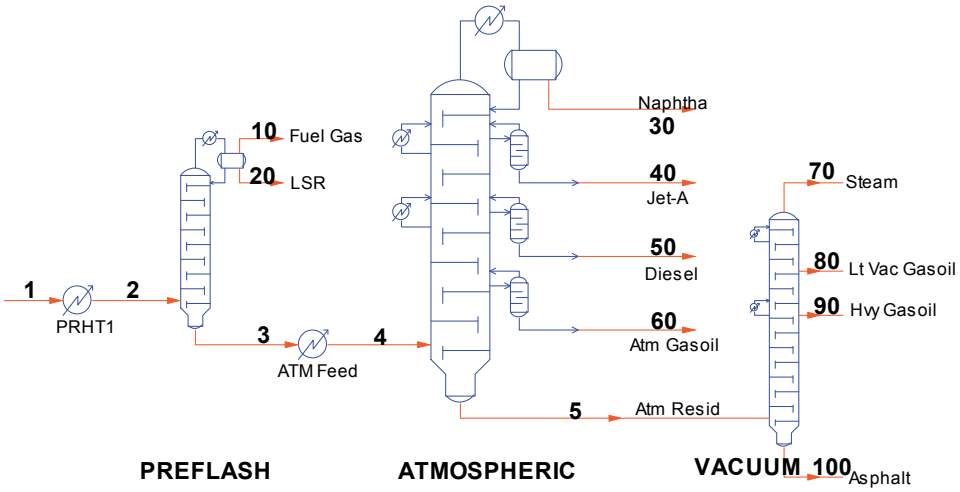
This Total refine column simulation is set-up to show free water balance closure.



Heatpump.in – Crude Unit Example

This is the same as c:\designii\samples\general\CRUDE.PSD in an input file form.

The purpose of this simulation is to determine the quality of products which can be made from a series of crude distillation columns for a feed of 100000 barrels per day. The amount of duty required for feed preheat (including feed furnaces for the atmospheric and vacuum columns) will also be calculated.



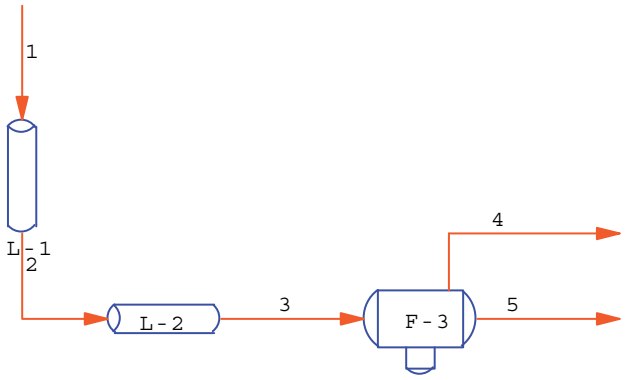

```
AB123.
*CRUDE UNITS -PREFLASH; ATMOSPHERIC AND VACUUM

C- HEAT EXCHANGER MODULE:    1   (PRHT)

HEAT EXCHANGER 1 = PRHT, 1,-2,
C- HEAT EXCHANGER SPECIFICATION
         TEMPERATURE OUT = 450,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 10,
C- REFINE CRUDE TOWER 2   (PRFL)

REFINE 2 = PRFL, 2,-10,-20,-3,

C-      PARTIAL CONDENSER
        PARTIAL CONDENSER
        STAGES = 8
        LOCATION OF FEEDS = 7,
        PRESSURE OF TOP = 41.7
        TEMPERATURE OF TOP TRAY = 340,
        TEMPERATURE OF BOTTOM = 450,
        PRODUCT RATES (BBL/DAY) = 0,0,80000,
        PRODUCT RATE GUESS = 280,1950,3900,
        TEMPERATURE OF CONDENSER = 160,

C-      PRESSURE DROPS FOR MAIN COLUMN AND
C-      SIDE STRIPPERS
        DELTA PRESSURE = 3,

C-      CONDENSER DESCRIPTION
        PRESSURE OF CONDENSER = 39.7,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 8,
        TEMPERATURE OF STEAM = 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 5000,

C-      COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = FUEL GAS,LSR,ATM.FEED,
C- HEAT EXCHANGER MODULE:    3   (PRHT 2)
HEAT EXCHANGER 3 = PRHT 2, 3,-4,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE OUT = 625,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 15,

C- REFINE CRUDE TOWER 4   (ATM)
REFINE4 = ATM, 4,-30,-40,-50,-60,-5,
C-      TOTAL CONDENSER
        TOTAL CONDENSER
        STAGES = 24,4,3,2,
        LOCATION OF FEEDS = 21,
        PRESSURE OF TOP = 20.7
```

```
       TEMPERATURE OF TOP TRAY = 360,
       TEMPERATURE OF BOTTOM = 620,
       PRODUCT RATES (BBL/DAY) = 0,35410,
       PRODUCT RATE GUESS = 730,1050,

C-     PRESSURE DROPS FOR MAIN COLUMN AND SIDE C-    STRIPPERS
       DELTA PRESSURE = 4,

C-     CONDENSER DESCRIPTION
       PRESSURE OF CONDENSER = 15.7,

C-     SIDE STRIPPER DESCRIPTION AND
C-     SPECIFICATION
       LOCATION OF DRAW TRAYS = 5,12,17
       LOCATION OF VAPOR RETURN = 5,12,17

C-     SIDE STRIPPER SPECIFICATIONS
       PRODUCT RATES FROM STRIPPER (BBL/DAY) = 8165,
       16967,10000,
       PRODUCT FROM STRIPPER GUESS = 615,1050,470,

C-     PUMPAROUND DESCRIPTIONS
       LOCATION OF PUMPAROUND = 7,5,13,12,
C-     PUMPAROUND SPECIFICATIONS
       PRODUCT OF PUMPAROUND (BBL/DAY) = 49000,11000,
       HEAT FROM PUMPAROUND = 40E6,15E6,
C-     PUMPAROUND MOLAR RATE ESTIMATES
       PRODUCT OF PUMPAROUND GUESS = 3600,700,

C-     STEAM FEED SPECIFICATIONS
       LOCATION OF STEAM = 25,29,32,34,
       TEMPERATURE OF STEAM = 4*400,
       PRESSURE OF STEAM = 4*60,
       STEAM RATES (LB/HR) = 12000,3300,1000,800,

C-    PRODUCT PROPERTY SPECIFICATIONS
       PRODUCT SPEC 1 = 625, ASTM95PT , PROD3, PROD4,

C- TRAY FIX COMMANDS
              FIX LIQUID (BBL/DAY) 1 = 1500, 20, H1,
              FIX GUESS = 50,

C-     COLUMN PRINT CONTROL
       REPORT = 1, 0, 0, 0, 0, 0,
       PLOT OFF,
       PRODUCT NAMES = NAPHTHA,JET,DIESEL,AGO,RESID,
C- REFINE CRUDE TOWER 15  (VTWR)

REFINE 15 = VTWR, 5,-70,-80,-90,-100,

C-     PUMPAROUND INTERNAL REFLUX
       PUMREF
       STAGES = 6
       LOCATION OF FEEDS = 6,
       PRESSURE OF TOP (MMHG) = 75
       TEMPERATURE OF TOP TRAY = 450,
       TEMPERATURE OF BOTTOM = 700,

C-     SIDE DRAW DESCRIPTION
       LOCATION OF SIDE DRAWS = 2, 4,
       PRODUCT RATES (BBL/DAY) = 0,6200,18600,10600,
       PRODUCT RATE GUESS = 1.2,269.4,592.3,195.2,

C-     PRESSURE DROPS FOR MAIN COLUMN AND
C-     SIDE STRIPPERS
       DELTA PRESSURE (MMHG) = 30,

C-     PUMPAROUND DESCRIPTIONS
       LOCATION OF PUMPAROUND = 2,1,4,3,
C-     PUMPAROUND SPECIFICATIONS
       PRODUCT OF PUMPAROUND (BBL/DAY) = 20000,
       49000,
       HEAT FROM PUMPAROUND = 14E6,54E6,
C-     PUMPAROUND MOLAR RATE ESTIMATES
       PRODUCT OF PUMPAROUND GUESS = 870,1560,

C-     STEAM FEED SPECIFICATIONS
```

```
        LOCATION OF STEAM = 7,
        TEMPERATURE OF STEAM= 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 20000,

C- TRAY FIX COMMANDS
        FIX TEMP 1 = 150, 1, 1,
        FIX LIQ NET 2 = 35, 2, 3,
        FIX LIQ NET 3 = 2, 5, H1,

C-     COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = STEAM,LVGO,HVGO,ASPHALT,

GENERAL
        COMPONENTS = 62,2,3,4,5,6,7,8,
C- THERMO PROPERTY OPTIONS
        ESSOTAB,  API,  STDD,
        VISCOSITY OPTION = STD,
        THERMAL CONDUCTIVITY OPTION = NBS81,
        PETROLEUM STREAMS =     10,20,30,40,
        50,60,70,80,90,100,

CRUDE PROPERTY SECTION
C- DATA FOR PSEUDO COMPONENTS WITHOUT C- ID NUMBERS (CUTS)
        TEMPERATURE INCREMENTS FOR TBP CURVES = 30,
        60,120,
        TEMPERATURE BREAKS FOR TBP CURVES (F) = 500,900,
CRUDE STREAM
C- BULK STREAM - PSEUDO COMPONENTS WITHOUT ID C- NUMBERS: 1 (STREAM 1)
        FEED NAME 1 = STREAM 1,
        TP 1 = 215,59.7,
        FEED DRY RATE (BBL/DAY) 1 = 1000000,
        FEED REAL PERCENTS 1 = .1,.2,.5,.5,1,1.5,2.5,
        FEED TBP 1 = -50,100,200,300,400,470,550,
                        650,750,850,1100,1300,1475,1670,
        FEED VOLUME PERCENTS 1 = 2,5,10,20,30,40,
                50,60,70,80,90,95,98,100,
        FEED GRAVITIES (API) 1 = 150,95,65,45,40,
                38,33,30,25,20,15,10,5,
        FEED GRAVITY VOLUME PERCENTS 1 = 2,5,10,20,
                30,40,50,60,70,80,90,95,98,

END
```

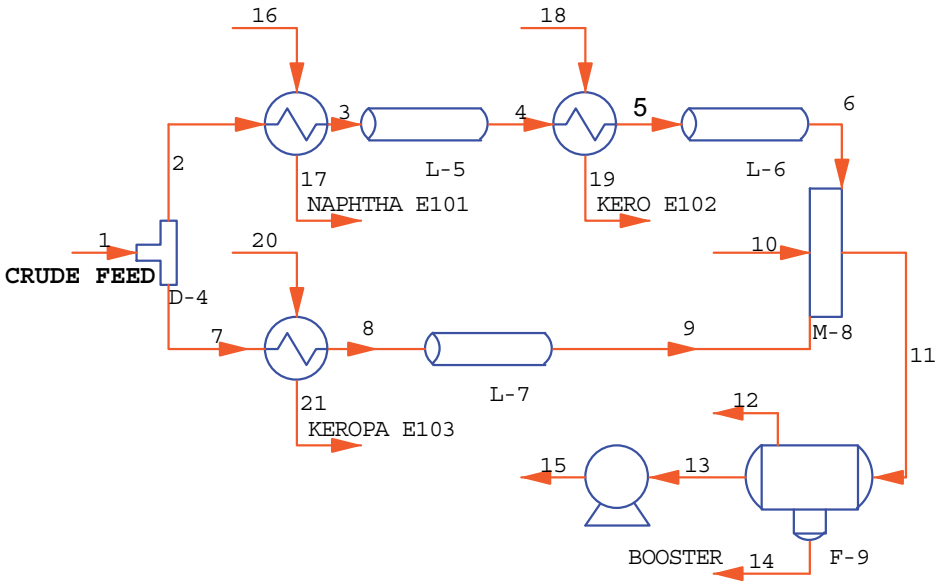
Holdup.psd - Holdup And Heat Transfer In Buried Pipeline

The simulation shows the overall calculation of pressure drop due to holdup, in addition to head & friction loss for a natural gas mixture.



Httrain.psd - Crude Pre-Heat Train

The simulation illustrates the crude pre-heat train problem. Feed crude in stream #1 is divided and pre-heated before it reaches the booster pump for further refining. All three exchangers are being rated by the BELL method.



Insulate.in – Two-Phase Flow Line

This is a two-phase line to onshore separator simulation with some in-built fortran calculations.

Lineftrn.in - Two-Phase Flow

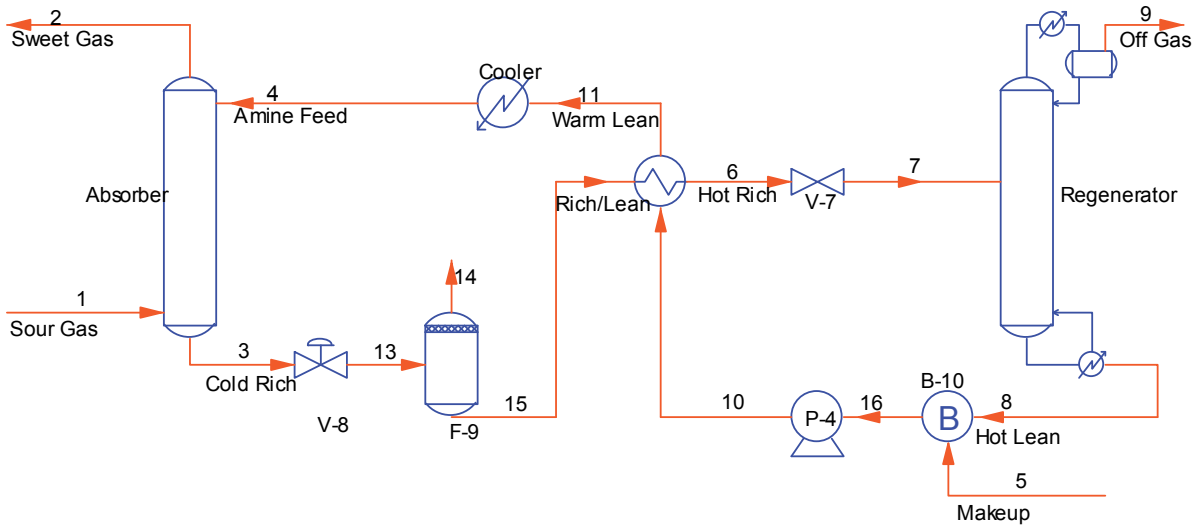
This is a two-phase flow in a buried line pressure calculations using heat transfer to the surroundings. Flash vessel sizing is also performed.

Linemeth.in - needs description

This is a two-phase steam line using the Beggs and Brill method for pressure drop.

MxpInt1.psd - Mixed Amine Plant for CO2 / H2S Removal

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA).



Newrefi.in – Crude Refine Column

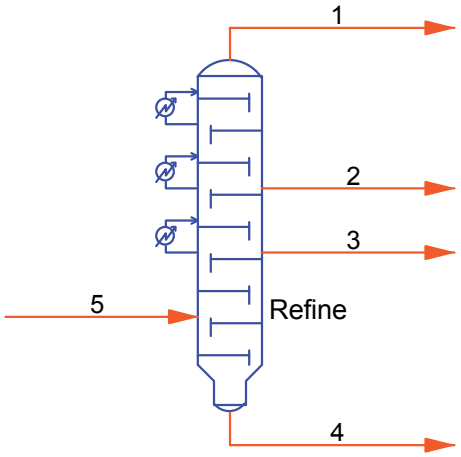
This is crude refine column simulation with some Inline fortran to retrieve vapor and liquid mass fractions for a Tray.

Pentium.in – Math Coprocessor Check

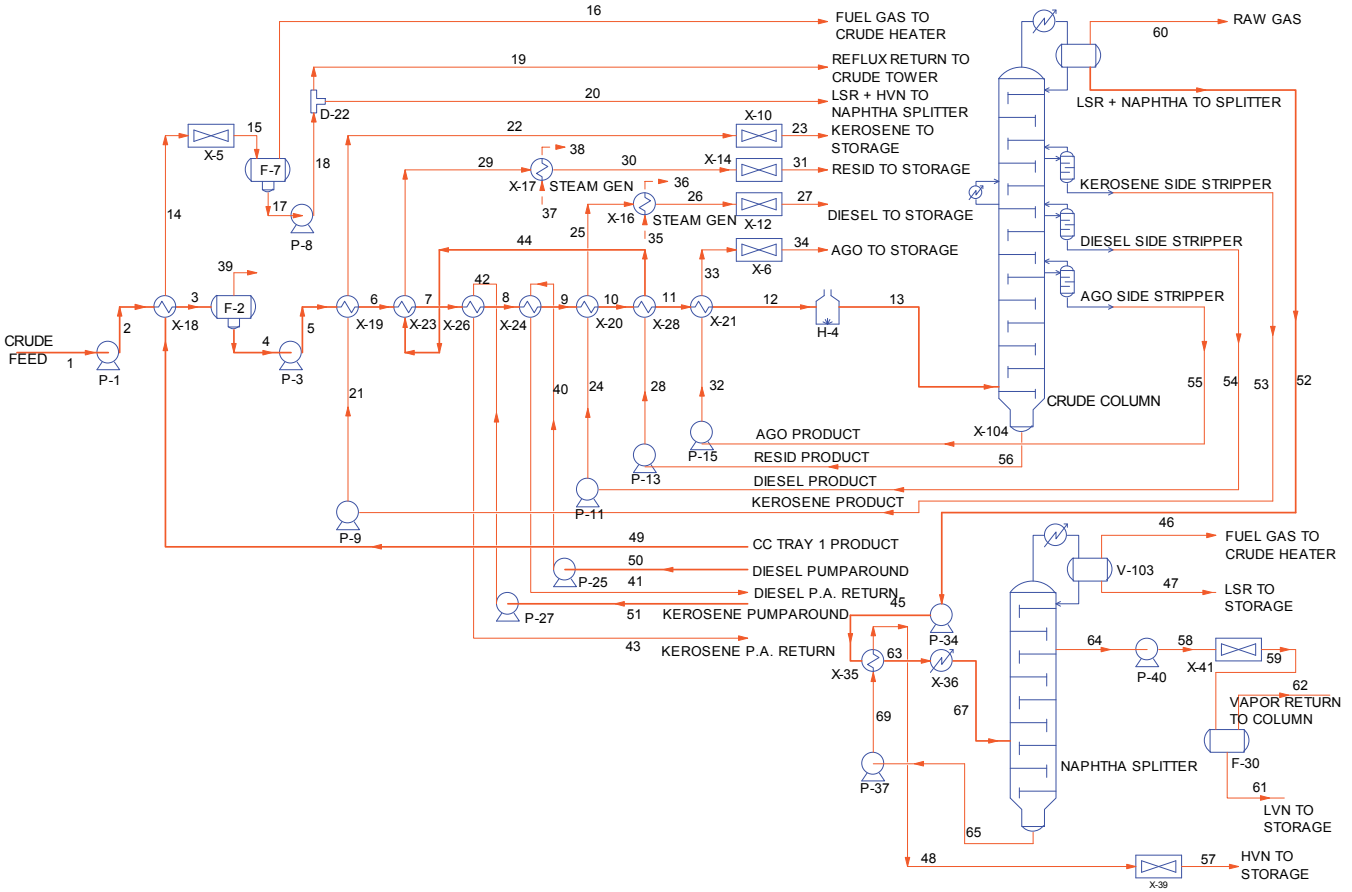
This file can be used to check whether the CPU has a good math coprocessor for the Pentium II math bug.

Refi_ex3.psd - Pumparound Internal Reflux without Condenser (PUM REF)

This simulation shows an internally refluxed column with no overhead condenser to provide reflux. Instead, a pumparound on tray 1 generates internal reflux. With this column type, there is no automatic decant of liquid water.



Refinery.psd - Crude Column with Heat Train and Naphtha Splitter



This is a general purpose flowsheet for a typical refinery operation. Shows a combined simulation of a pre-heat train and light and heavy naptha refine columns.

Skdmphse.in – Multiple Phase Flash Using SKD K-values Method

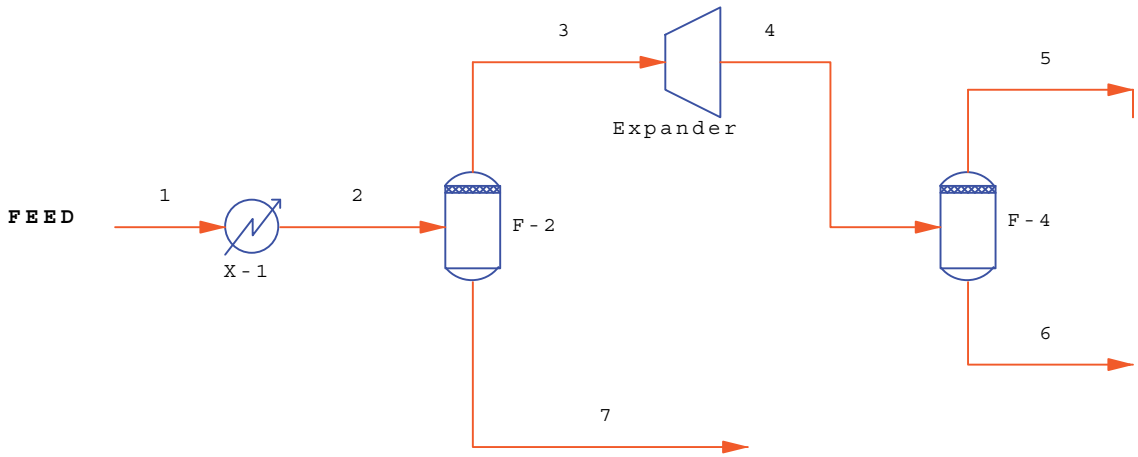
This is multiple phase flash simulation with SKD K-values method.

Surround.in – Line Pressure Drop Calculations

This is input file for line pressure drop calculations with heat transfer to surroundings.

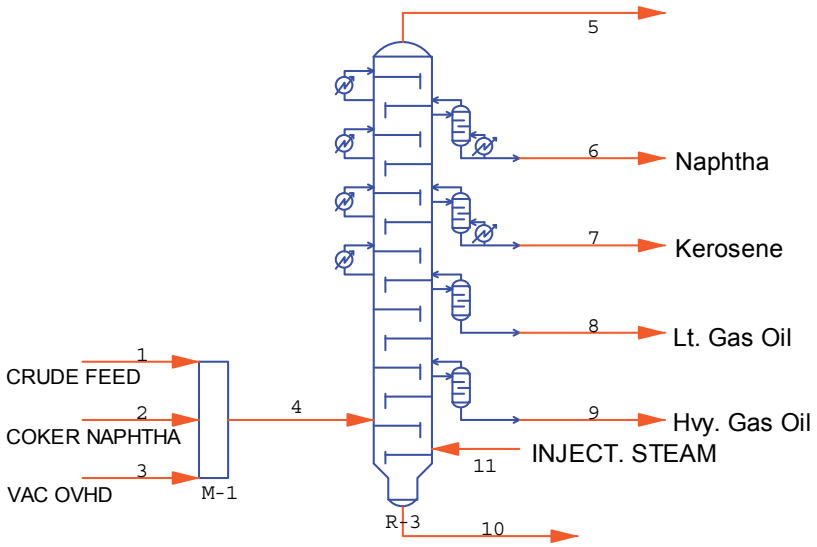
Tutorial.psd - Tutorial Flowsheet

This flowsheet is set up for the benefit of a new user. The example can be used to explore various options in DESIGN II.



Typ3reb.psd - Refine Column Hotdraw Thermosiphon Reboiler

This flowsheet is an example of a refine column where the side strippers use thermosiphon hotdraw reboilers to purify the required products.


5
6
Naphtha
7
Kerosene
8
Lt. Gas Oil
9
Hvy. Gas Oil
INJECT. STEAM
11
1
CRUDE FEED
2
COKER NAPHTHA
3
VAC OVHD
M-1
4
R-3
10

Chapter 4: Activated MDEA Samples - c:\designii\samples\activatedMDEA

Figure1.jpg – Flow Diagram for a Ammonia Plant

A flow diagram for a typical CO2 Removal System commonly used in an Ammonia plant is shown in Figure 1. Ammonia plants generate tremendous amounts of CO2 to be removed in the manufacture of Syn-Gas, bearing Hydrogen and Nitrogen.

Figure 1
MDEA Based CO2 Removal System for an Ammonia Plant

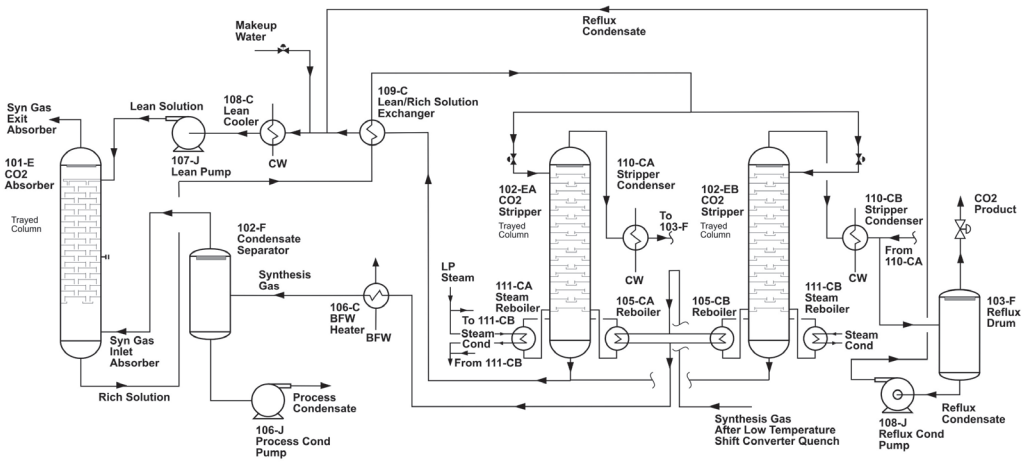


Figure2.psd – Ammonia Plant CO2 Removal System

In Figure 2, an Ammonia plant CO2 Removal System operates in a plant at 1550 Short Tons Per Day (STPD) Ammonia production (1406 mt/d). It is the Base Case, the CO2 Removal System operates at an efficiency of 50,050 Btu/LbMole (27800 Kcal/KgMole) of CO2 removed. These operating conditions and the operating efficiency is typical of many aMDEA based systems.

Figure 2 Example 1
Activated MDEA (aMDEA) Based CO2 Removal System - Base Case at 1550 Short Tons/Day Ammonia Plant
(Modeled with WinSim, Inc. Design II)

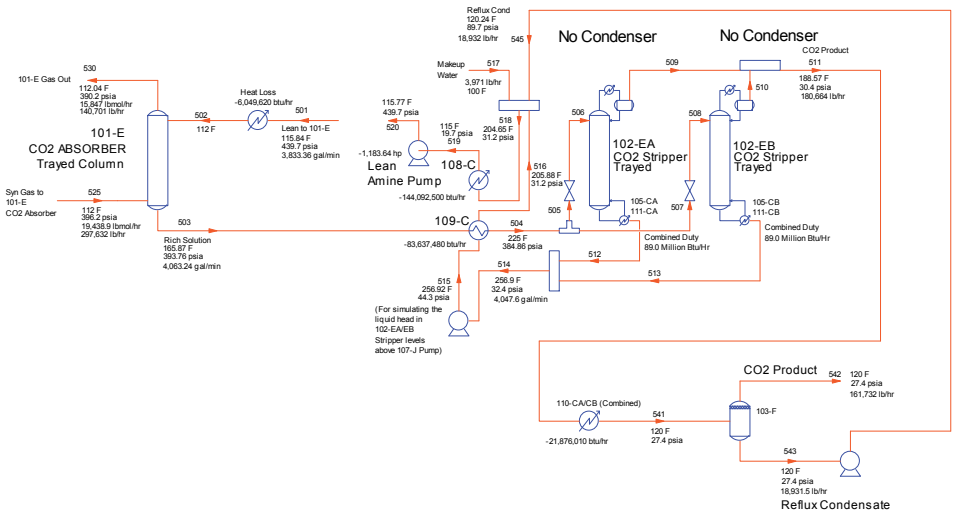


Figure3.psd – Optimizing the CO2 Removal System

Figure 3 illustrates how Design II process simulation software can be used to optimize the CO2 Removal System process efficiency and energy use.

Figure 3 Example 2
Activated MDEA (aMDEA) Based CO2 Removal System Optimization Case
at 1550 Short Tons/Day Ammonia Plant
(Modeled with WinSim, Inc. Design II)

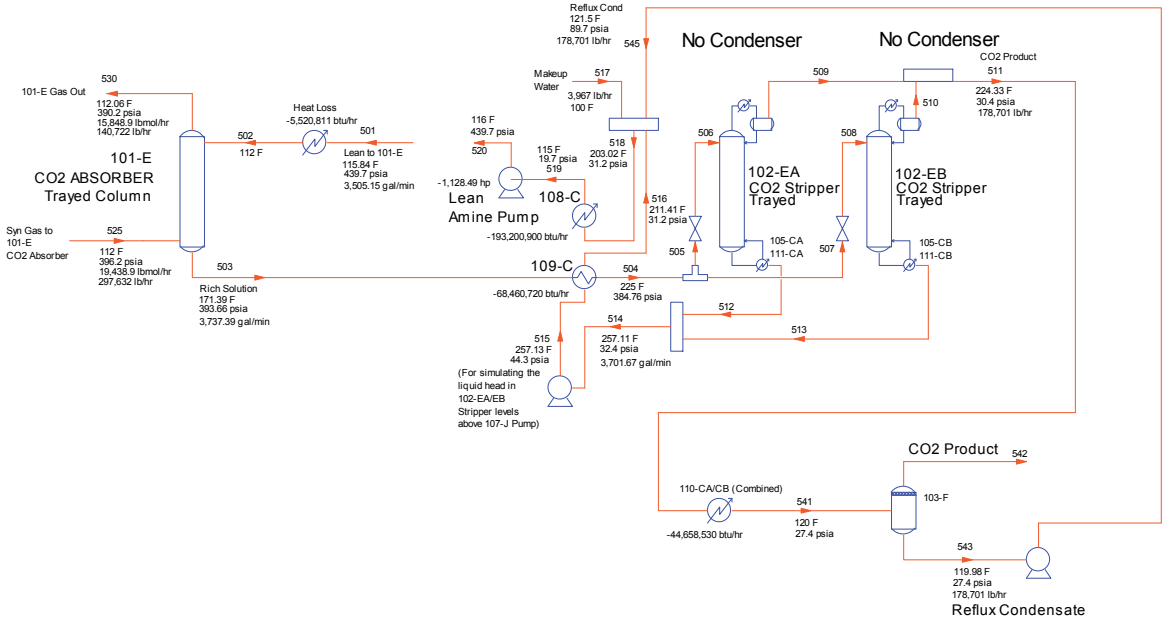


Figure4.psd – Developing Equipment Upgrades to Expand Production

Figure 4 illustrates how Design II process simulation software can be used to develop equipment upgrades needed to expand plant production. The plant illustrated in the Base Case aims to expand production capacity from 1550 STPD (1406 mt/d) to 1800 STPD (1633 mt/d) Ammonia production

Figure 4 Example 3
Activated MDEA (aMDEA) Based CO2 Removal System - Expanded to 1800 Short Tons/Day Ammonia Plant Production
(Modeled with WinSim, Inc. Design II)

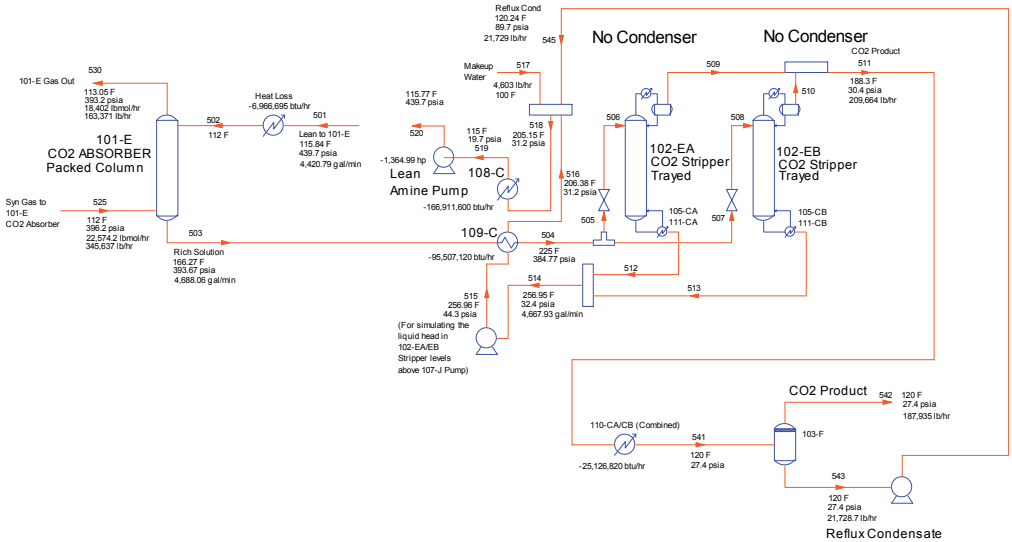
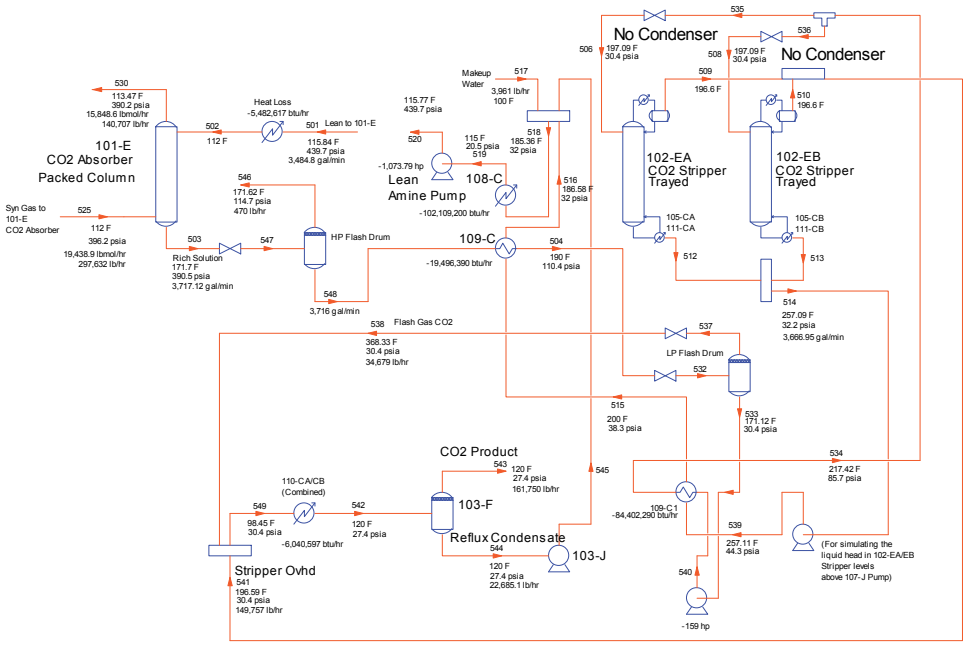


Figure5.psd – CO2 Removal System Flow Diagram

Figure 5 shows a CO2 Removal System flow diagram developed using Design II, for an energy saving retrofit for the Base Case plant at 1550 STPD (1406 mt/d) Ammonia production.

Figure 5 Example 4
Activated MDEA (aMDEA) Based CO2 Removal System - Retrofit Case at 1550 Short Tons/Day Ammonia Plant
(Modeled with WinSim, Inc. Design II)



Value of Energy Savings.xls – Energy Reduction

These examples illustrate how process simulation software can aid plants that use CO2 Removal System equipment optimize energy use, design plant expansions and accomplish major energy reduction projects.

aMDEA Based CO2 Removal System
Value of Energy Savings

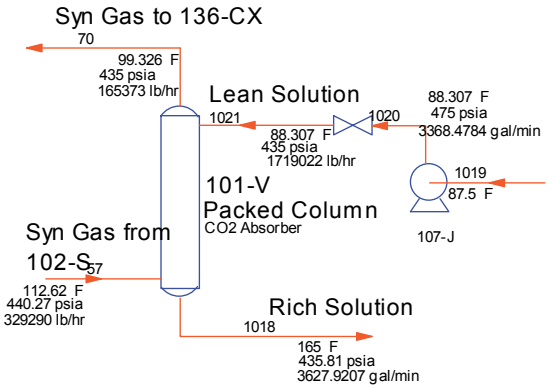
Case	Optimize	Retrofit	
Base Energy, Reboilers MM Btu/Hr	178	178	
New Case Energy, Reboilers MM Btu/Hr	171	141.2	
Energy Savings, MM Btu/Hr	7	36.8	
Fuel System HHV %Eff	83	83	
HHV Energy MM Btu/Hr	8.434	44.337	
Operating Days/Year	350	350	
Energy Cost, $/MM Btu, HHV	$7.00	$7.00	Typical current US energy cost, July-Aug. 2006
Energy Savings, $/Year	**$496,000**	**$2,607,000**	

Optimize - refers to optimizing the Base Case plant operating conditions to save energy
Retrofit - refers to addition of new equipment to reduce energy
use

Chapter 5: Ammonia Plant Samples - c:\designii\samples\ammonia

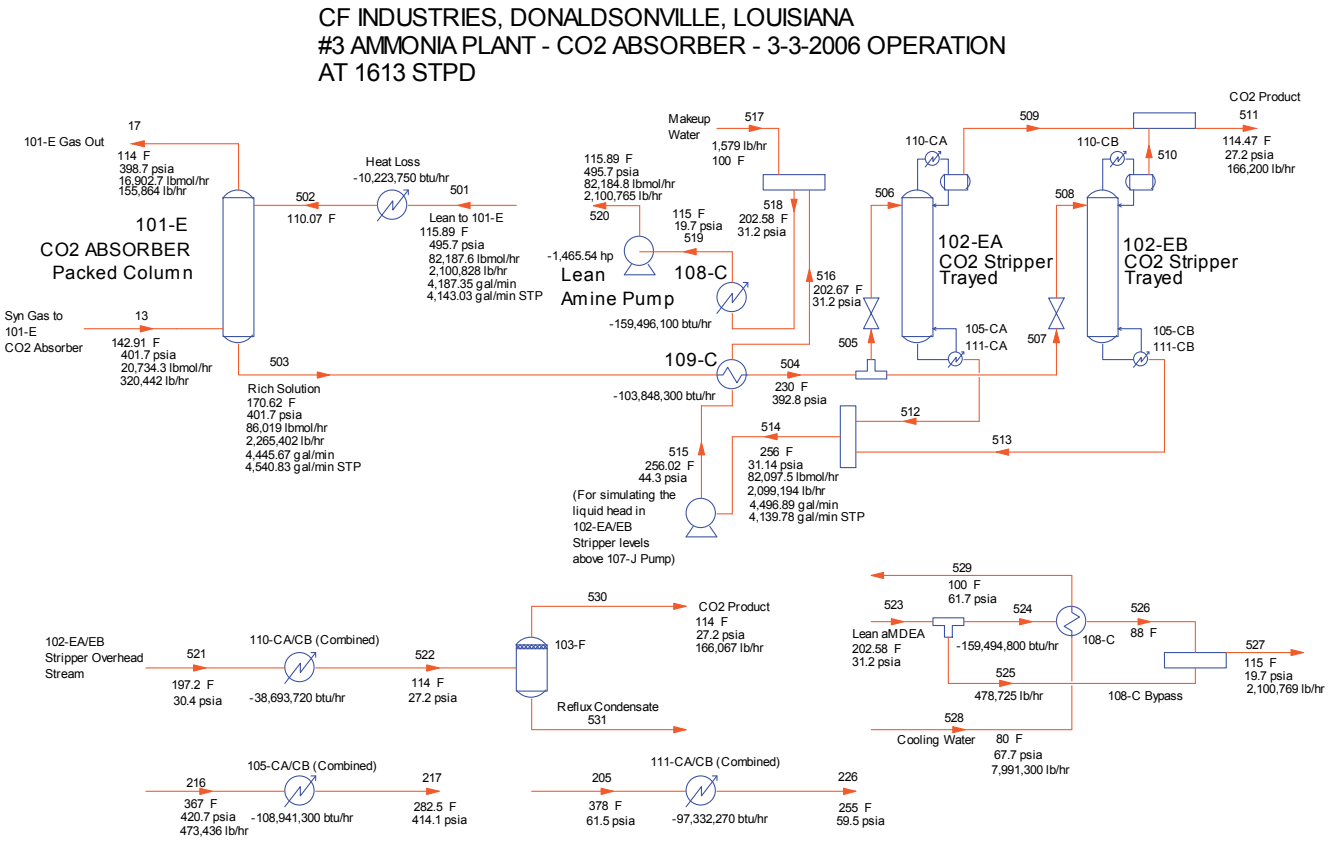
CO2 Absorber-Frontend with Purifier.psd – Stand-alone CO2 Absorber

This is a stand-alone CO2 Absorber psd for the Frontend using Syn Gas Purifier psd, permitting quick solutions of the synthesis gas preparation calculations, without the need for a complete CO2 Removal System simulation.



CO2remvl.psd - CO2 Removal System Simulation Details

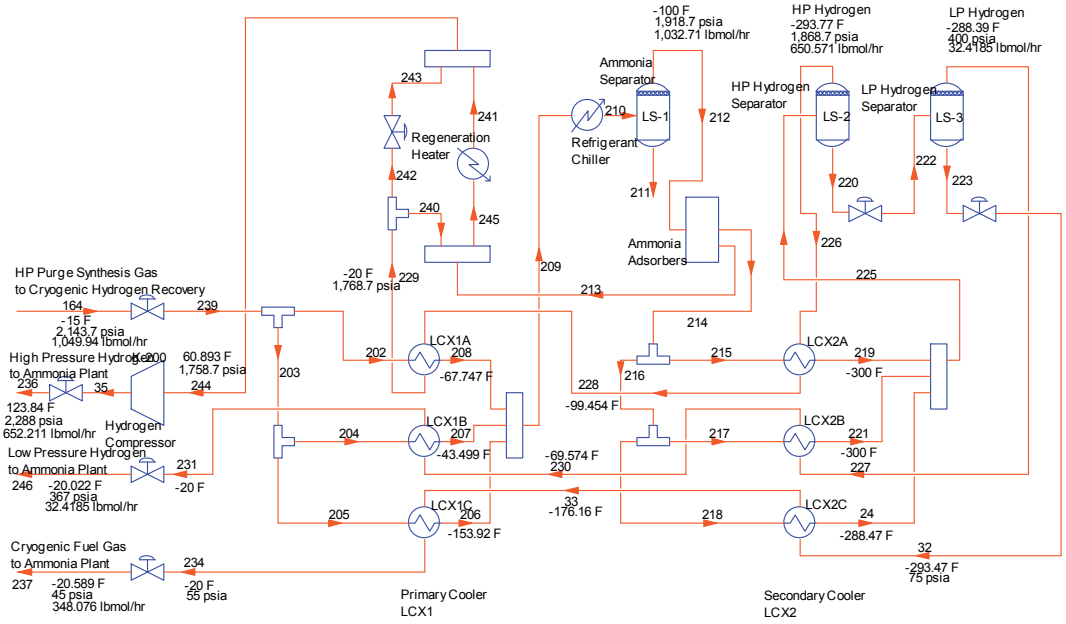
Design II software provides accurate assessment of carbon-dioxide removal systems modeling including thermodynamic, vapor liquid equilibria and component balances for diverse arrangements of lean solvent pumping, high pressure carbon dioxide absorbtion, lean/rich solvent heat exchange, rich solvent stripping, carbon dioxide cooling, condensate separation and solvent heat rejection.



Cryogenic H2 Recovery Unit.psd - Cryogenic Hydrogen Recovery System Simulation Details

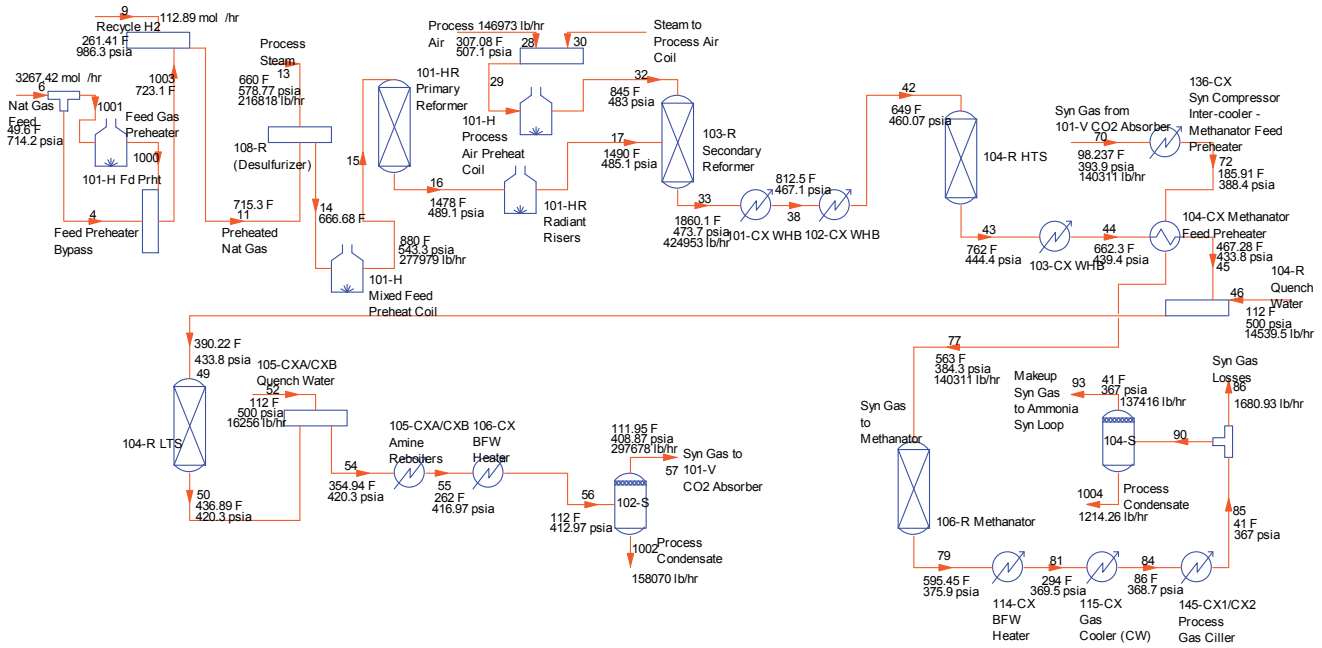
Cryogenic Primary Feed Cooler, Refrigerant Chiller, Ammonia Separator, Ammonia Absorbers, Cryogenic Secondary Feed Cooler, High Pressure Hydrogen Separator, Low Pressure Hydrogen Separator, High Pressure Hydrogen Regeneration Heater, High Pressure Hydrogen Compressor

1550 TPD AMMONIA PLANT HYDROGEN RECOVERY PROCESS



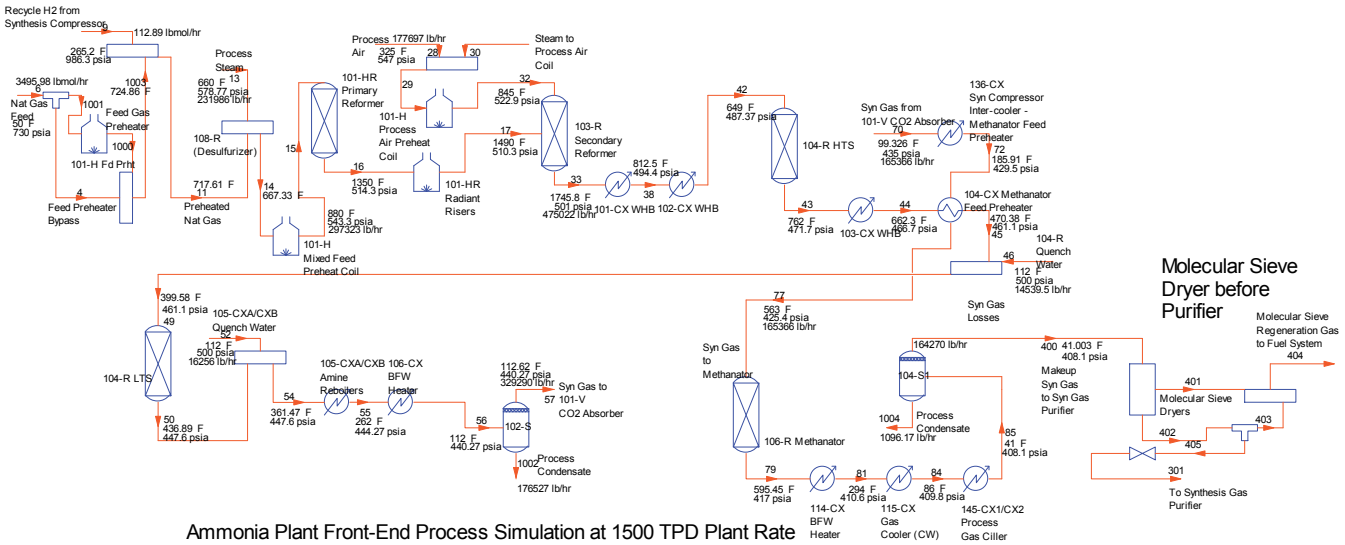
FrontEnd.psd - Front-End Simulation Details

Design II provides thermodynamic, equilibrium and component material balance calculations for all ammonia plant technologies including processes which incorporate various combinations of feedstock preheating, steam-hydrocarbon and auto-thermal reforming, waste heat steam generation, carbon-monoxide shifting, boiler water preheating, carbon dioxide removal, methanation, synthesis gas cooling and condensate recovery.



Ammonia Plant Front-End Process Simulation at 1550 TPD Plant Rate

Frontend using Syn Gas Purifier.psd – Variation on the Synthesis Gas Purifier

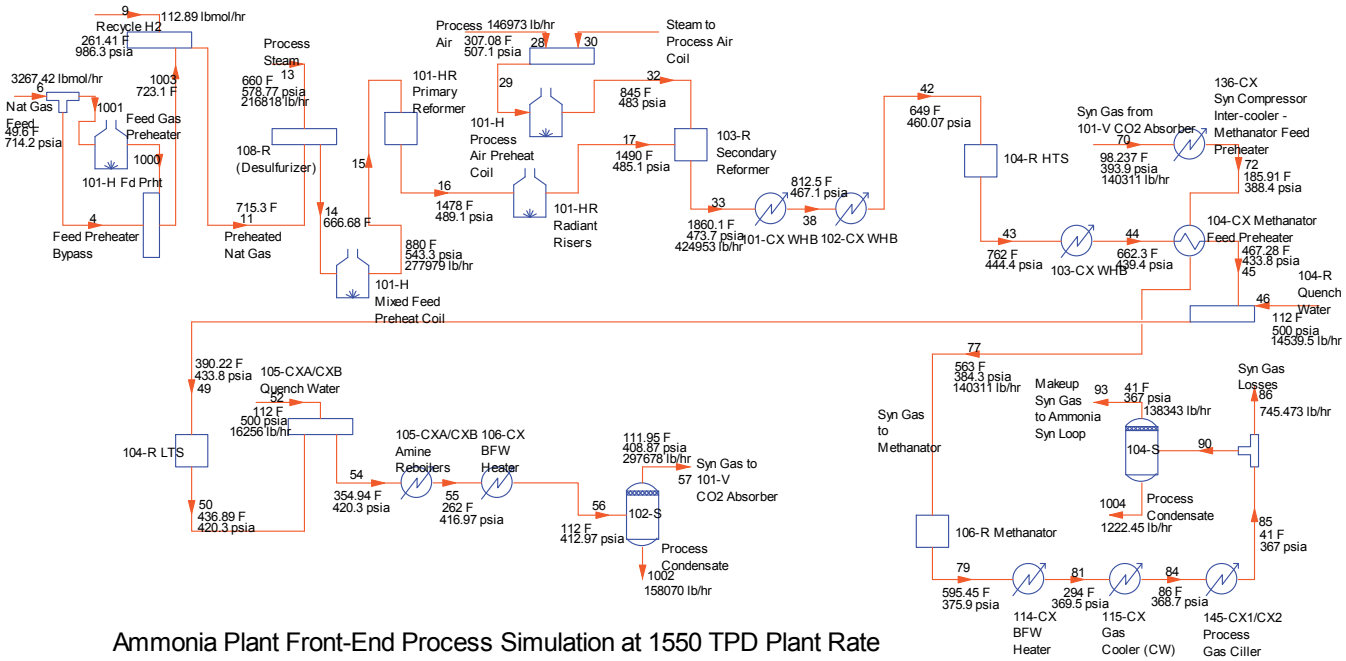


Ammonia Plant Front-End Process Simulation at 1500 TPD Plant Rate using Sythesis Gas Purifier

The synthesis gas preparation is different in two ways. First, the reforming conditions in the Primary and Secondary Reformer's are milder than traditional equipment, with lower operating temperatures yielding higher outlet Methane compositions. Second, the amount of Process Air is higher than stoichiometric requirements for ammonia synthesis to produce an excess of Nitrogen to the Synthesis Gas Purifier equipment to drive the cryogenic cooling.

FrontEnd with Cryogenic H2 Recovery.psd - Front-End Simulation Details

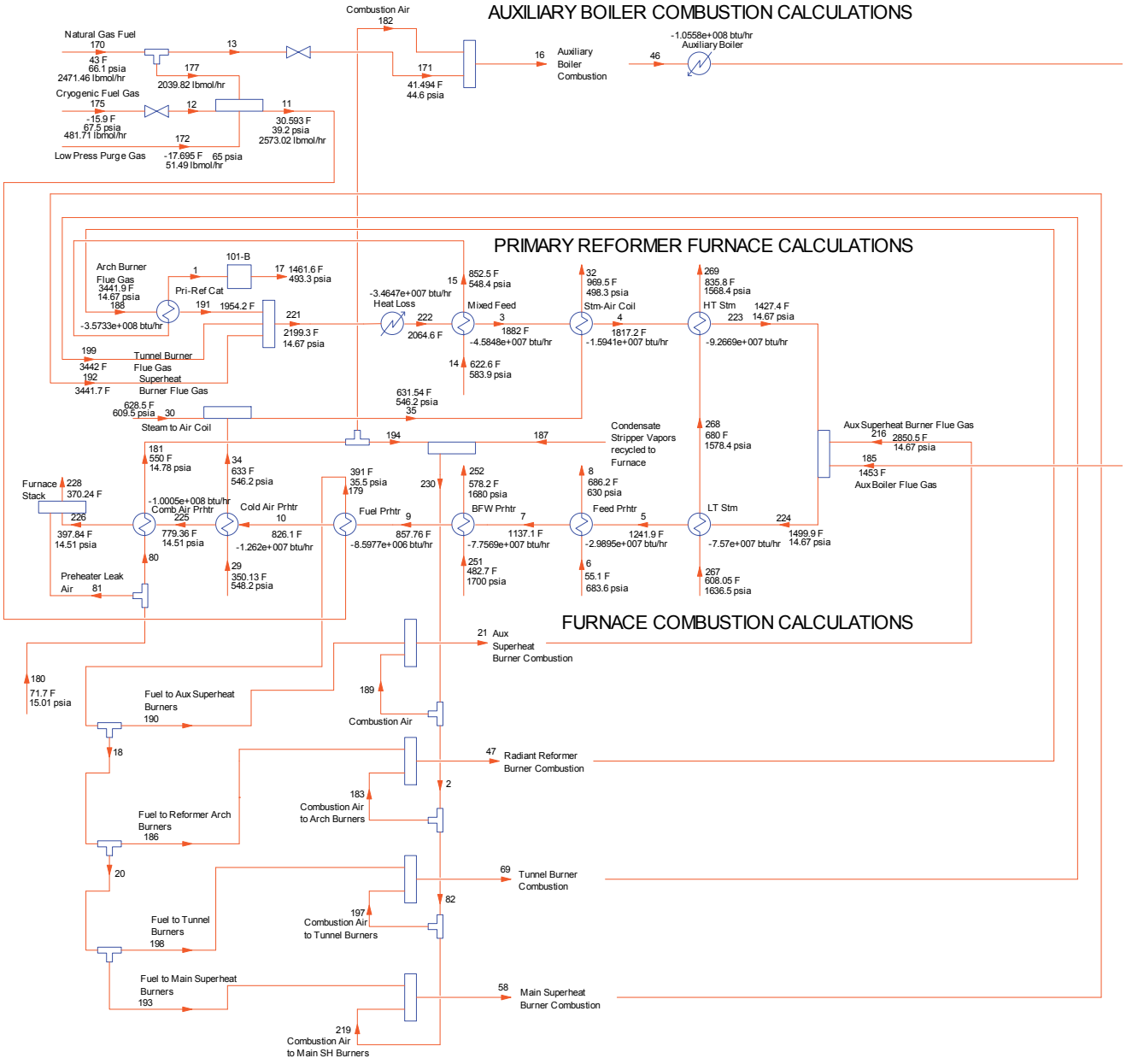
The calculations are the same as those for Frontend.psd, except integration with a Cryogenic Hydrogen recovery process is provided with a Cryogenic H2 Recovery.psd. The process highlighted here is quite elegant and has very high hydrogen recovery, 99%. Manual iterative calculations are performed between the Ammonia Synthesis with Cryogenic H2 Recovery.psd and the Cryogenic H2 Recovery Unit.psd until satisfactory reconciliation of their combined material and energy balance is achieved. Linking the two flowsheets could be possible, but the combined calculations with multiple recycle streams would be arduous.



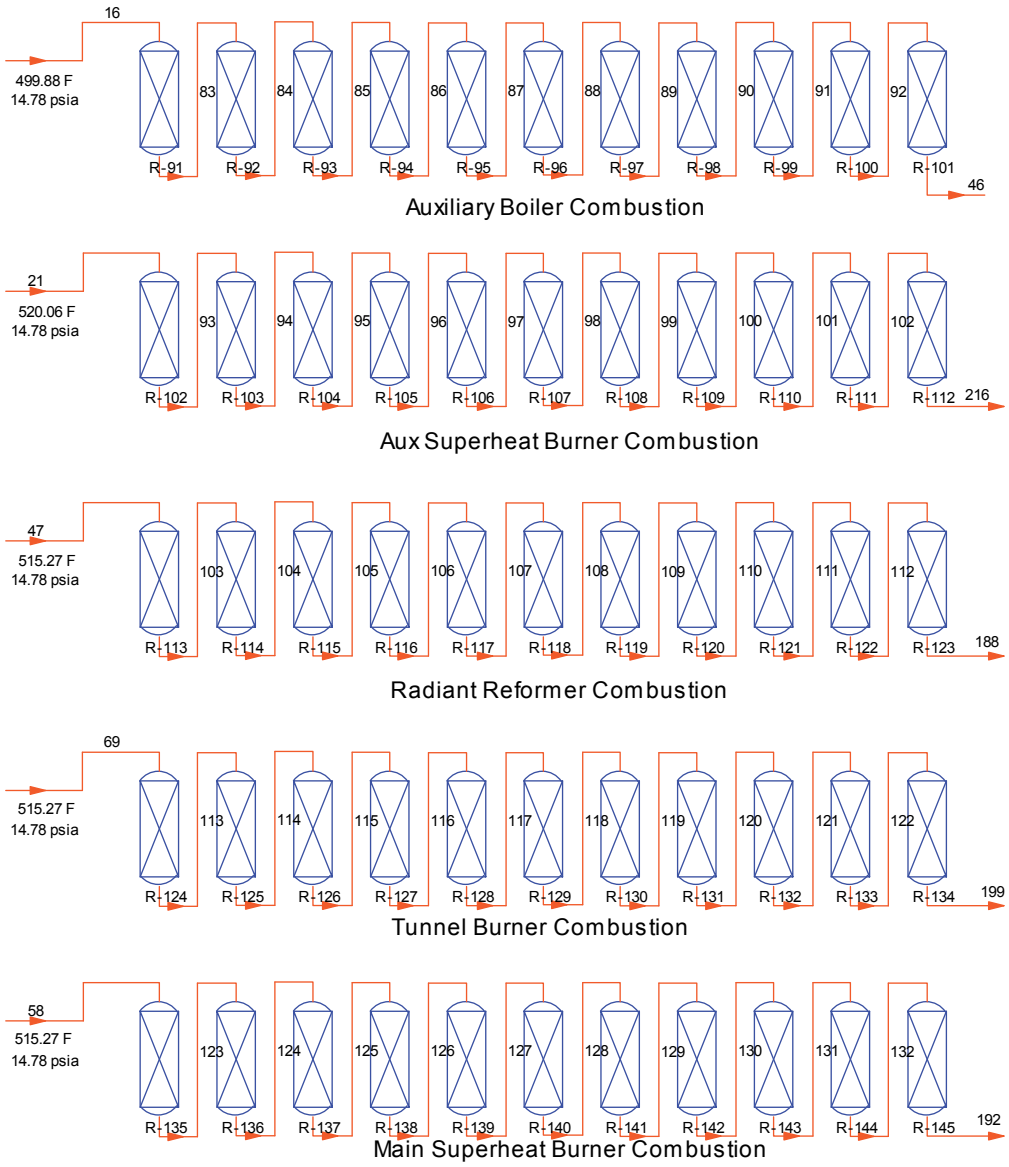
Ammonia Plant Front-End Process Simulation at 1550 TPD Plant Rate

Fuel System at 1790 TPD.psd - Ammonia Plant Fuel System Simulation at 1790 Tons Per Day

Fuel System sheet: Includes fuel gas mixing for Natural Gas, Cryogenic Fuel Gas and Low Pressure Purge Gas; Fuel Gas Combustion for Mixed Fuel to Reformer Arch,Tunnel and Superheat Burners; Natural Gas Fuel Combustion in direct fired boiler; Flue gas Heat Recovery in Primary (Radiant Heat Transfer) Reformer Reactor; Flue Gas Heat Recovery in preheating Mixed Feed Gas, High Temperature Steam/Air, High Temperature Steam Superheating; Flue Gas Blending to Cold Convection Section; Flue Gas Heat Recovery in preheating Low Temperature Steam Superheating, Natural Gas Feedstock Preheating, Boiler Feedwater Preheating, Mixed Fuel Gas Preheating, Cold Process Air Preheating and finally, Combustion Air Preheating for Primary Reformer Furnace and Steam Boiler combustion.

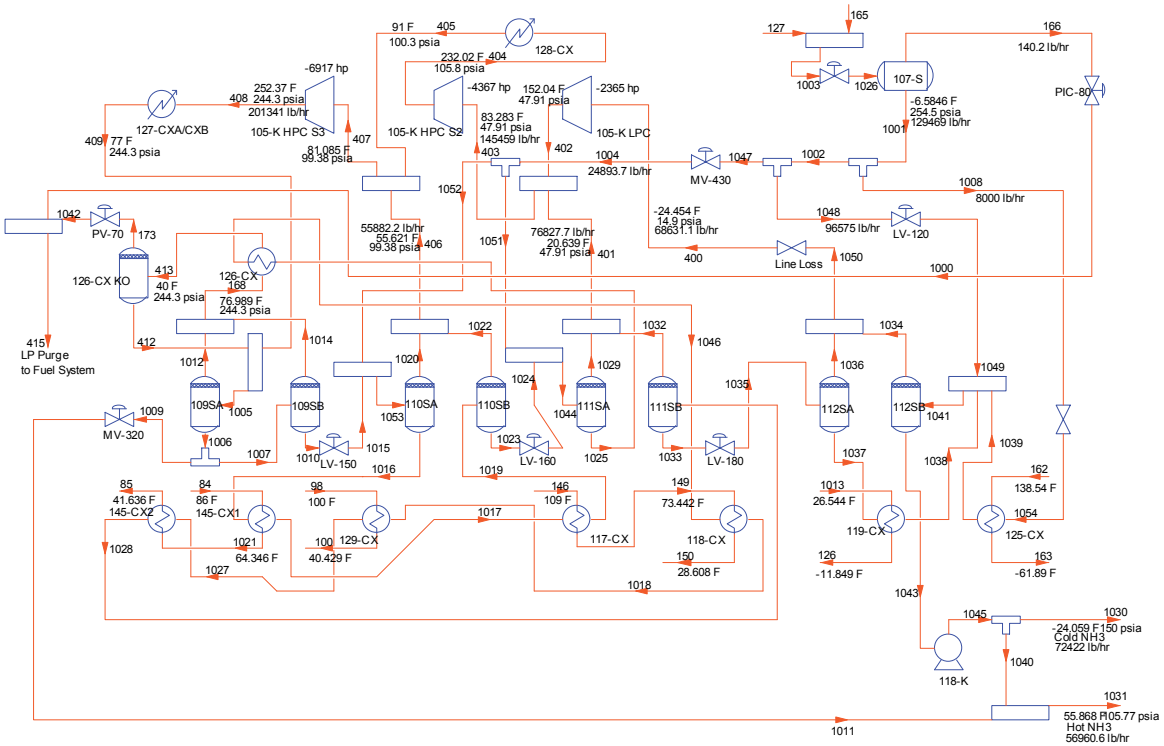


Combustion Calcs sheet: Includes adiabatic combustion calculations for Natural Gas and Mixed Fuel Gas to Reformer Furnace Arch Burners, Main and Auxiliary Superheat Burners, Tunnel Burners and Steam Boiler Burners. Combustion calculations for adiabatic reactions of Mixed Fuel and Natural Gas Fuel are provided generating flue gases for the combustion of Methane, Ethane, Propane, Iso-Butane, Normal-Butane, Iso-Pentane, Normal-Pentane, Normal-Hexane, Carbon Monoxide, Hydrogen and Ammonia fuel constituents.



NH3Refrg.psd - Ammonia Refrigeration System Simulation Details

Design II rigorously models energy usage, vapor-liquid equilibria, thermodynamics and component balances for multi-stage refrigeration systems using inter-cooled compressors rejecting heat from process gas chillers via refrigerant ammonia condensing, including ammonia product let-down, flash drum equilibria and loads, inerts purge gas rejection, hot/cold ammonia product pumps and flash drum liquid and vapor flows.

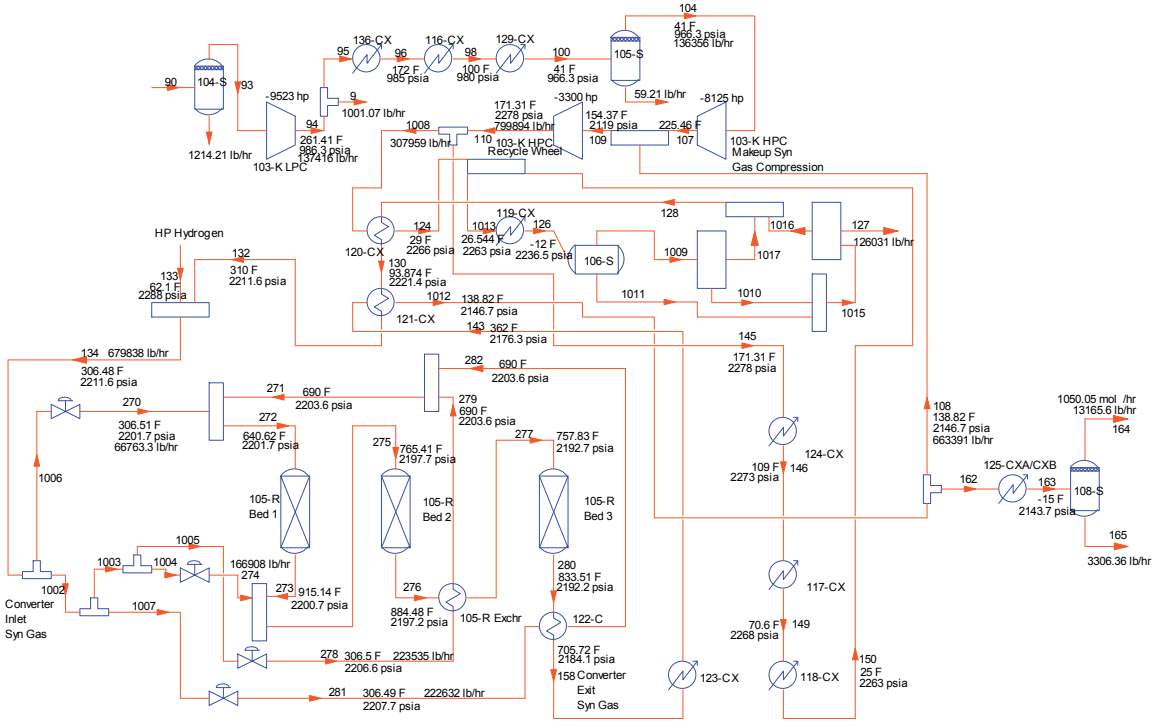


AMMONIA REFRIGERATION SYSTEM SIMULATION AT 1550 TPD

NH3Synth.psd - Ammonia Synthesis Process Simulation Details

Design II readily handles ammonia synthesis equipment calculations, including predictions of energy usage, reaction equilibria, thermodynamics, component balances and vapor-liquid equilibrium for complex plant arrangements and technologies which incorporate synthesis gas multi stage compressors with inter-cooling, synthesis gas cooling using water and refrigeration chilling, refrigeration recovery exchange, ammonia product separation, process heating and cooling, inter-cooled and quench ammonia converters with internal exchangers, steam generation and boiler water preheating, synthesis gas purge and the synergies of recycle of synthesis gas.

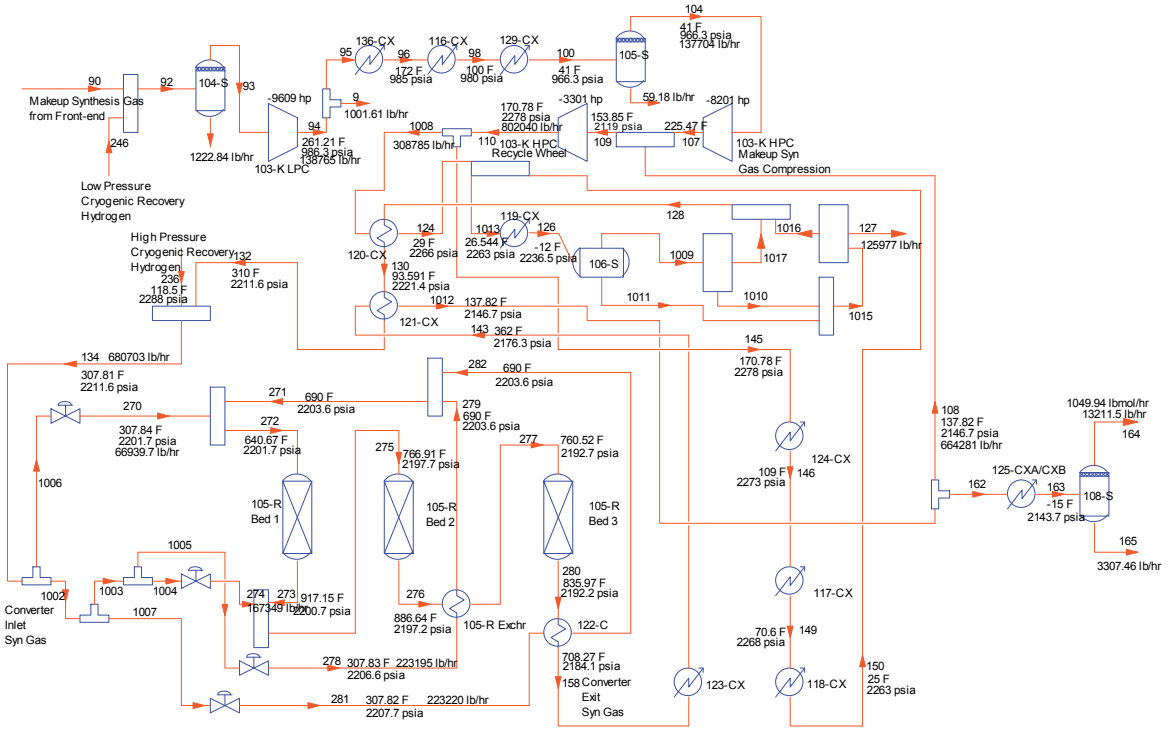
AMMONIA SYNTHESIS PROCESS SIMULATION AT 1550 TPD RATE



NH3 Synthesis with Cryogenic H2 Recovery.psd - Ammonia Synthesis Process Simulation Details

The calculations are the same as those for NH3Synth.psd, except integration with a Cryogenic Hydrogen recovery process is provided with a Cryogenic H2 Recovery.psd. The process highlighted here is quite elegant and has very high hydrogen recovery, 99%. Manual iterative calculations are performed between the Ammonia Synthesis with Cryogenic H2 Recovery.psd and the Cryogenic H2 Recovery Unit.psd until satisfactory reconciliation of their combined material and energy balance is achieved. Linking the two flowsheets could be possible, but the combined calculations with multiple recycle streams would be arduous.

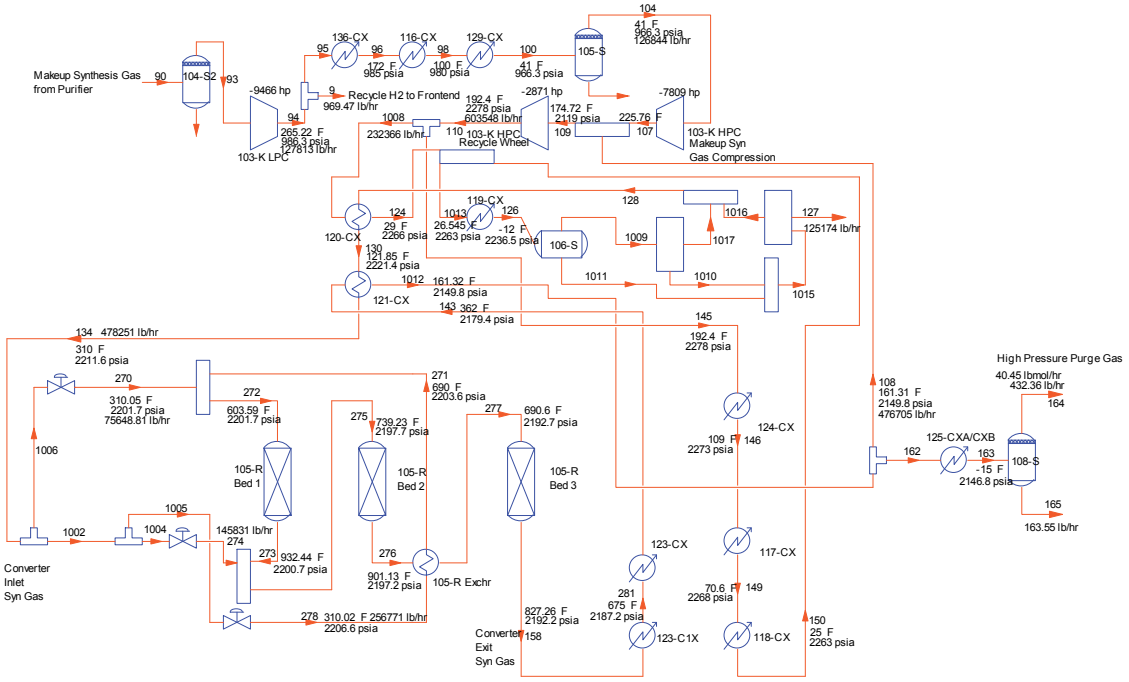
AMMONIA SYNTHESIS PROCESS SIMULATION AT 1550 TPD RATE WITH CRYOGENIC HYDROGEN RECOVERY



NH3 Synthesis with Purifier.psd – Ammonia Synthesis with Low Inerts

This Ammonia Synthesis process is similar to the conventional process, except that the Ammonia Converter operates with low (CH4 and Ar) inerts, achieving substantially higher than traditional conversion. The traditional 3rd bed feed/effluent inter-cooler exchanger is eliminated and the high temperature level heat is recovered in high pressure steam generation, downstream of the converter, prior to a conventional boiler feedwater heater. The high process efficiency generates substantial energy savings for a grass-roots or retrofit plant project.

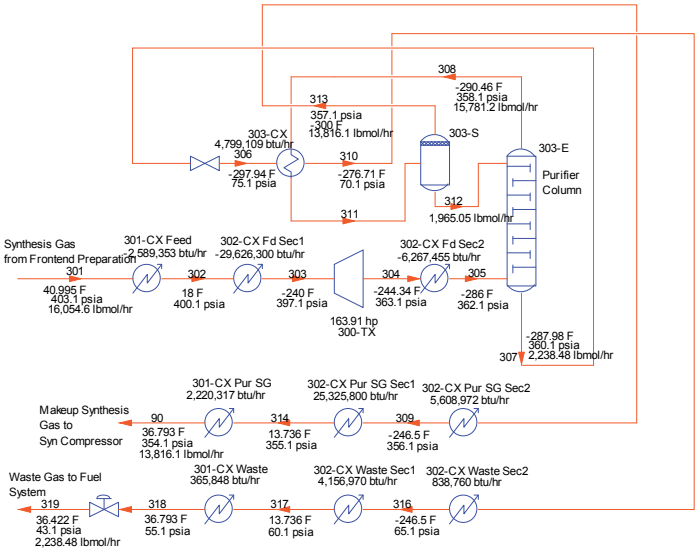
AMMONIA SYNTHESIS PROCESS FROM FRONT-END SYN GAS PREPARATION WITH PURIFIER AT 1500 TPD RATE



Syn Gas Purifier.psd – Cryogenic Process for Purifying Feed to Ammonia Syntheis Process

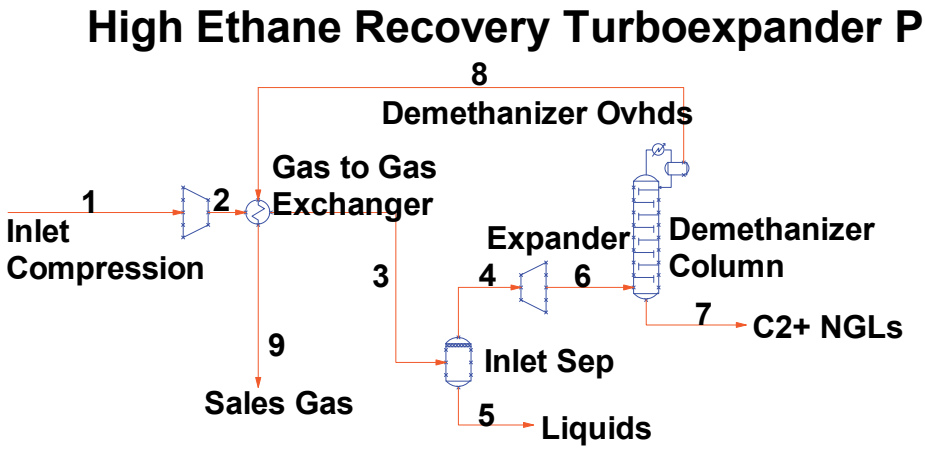
This cryogenic process removes 100% of the makeup Synthesis Gas Methane and more than 40% of the makeup Synthesis Gas Argon to the Ammonia Synthesis Process, promoting improved conversion in the Ammonia Synthesis process. The high inerts stream, including the excess Nitrogen used for cooling in the Purifier is sent to the fuel system of the plant for recovering its heating value, thus displacing natural gas fuel.

FRONT-END SYNTHESIS GAS PREPARATION - SYN GAS PURIFIER



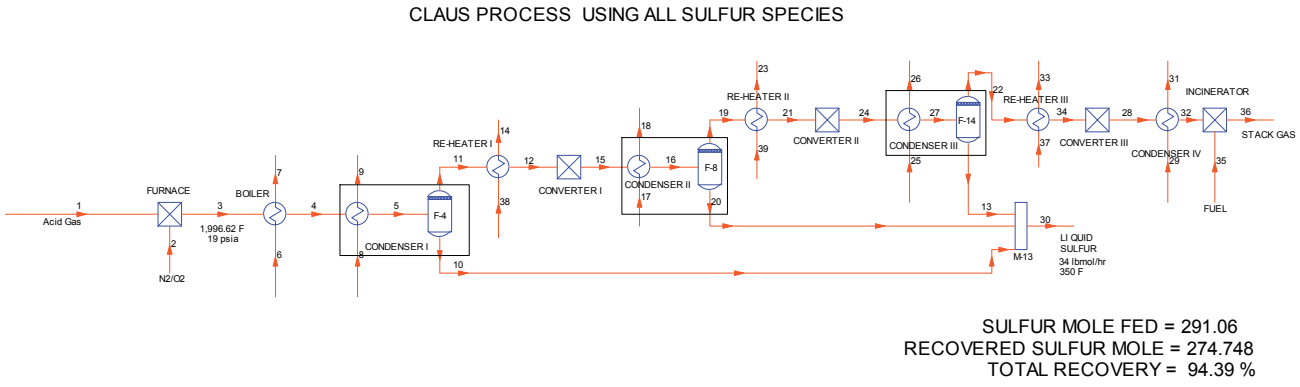
Chapter 6: Brochure Samples - c:\designii\brochure\samples

C:\designii\samples\brochure\C2plant.psd



C:\designii\samples\brochure\ClausAllSulfurSpecies.psd

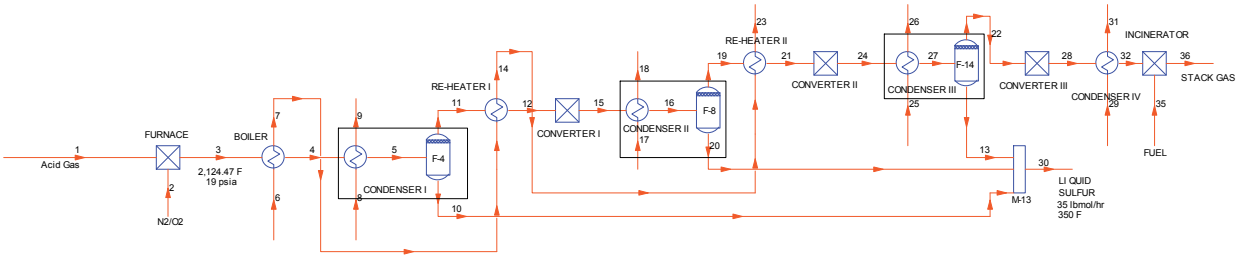
This sample flowsheet models the Claus Process in the GPSA data book using Sulfur 2, Sulfur 3, Sulfur 4, Sulfur 5, Sulfur 6, Sulfur 7, and Sulfur 8.



C:\designii\samples\brochure\ClausGPSA.psd

This sample flowsheet models the Claus Process in the GPSA data book using Sulfur 2, Sulfur 6, and Sulfur 8.

GPSA CLAUS PROCESS USING SULFUR 2, 6 and 8

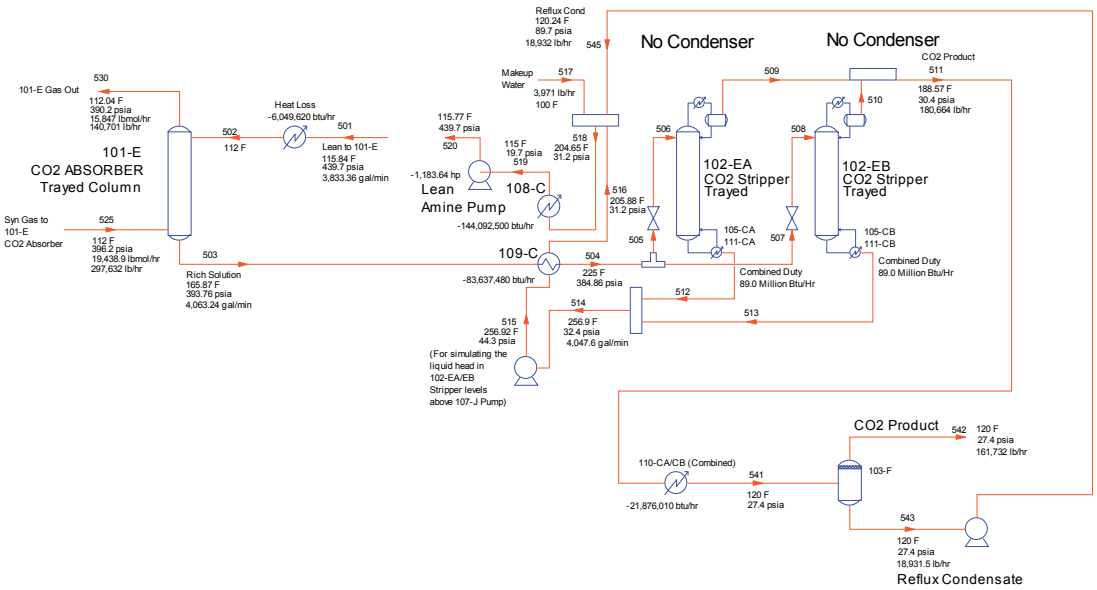


SULFUR MOLE FED = 291.06
RECOVERED SULFUR MOLE = 280.48
TOTAL RECOVERY = 96.36 %

C:\designii\samples\brochure\Co2remvl.psd

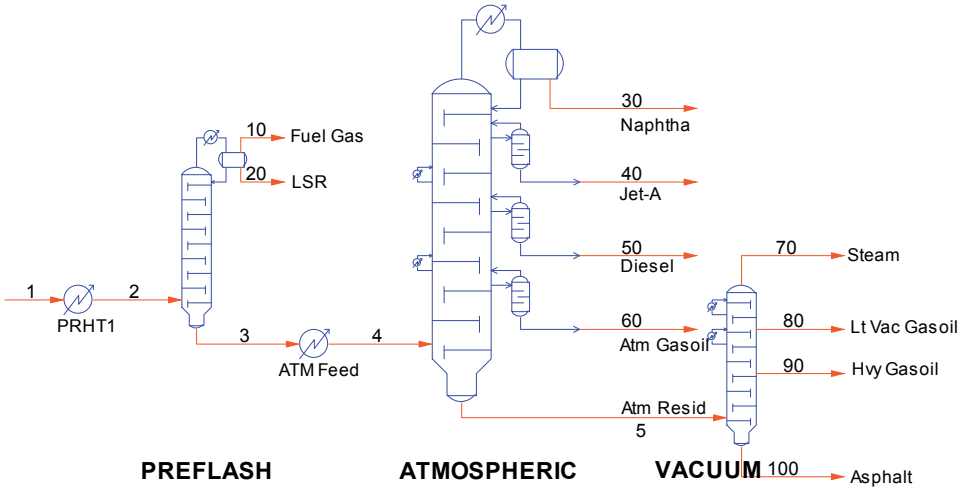
In Figure 2, an Ammonia plant CO2 Removal System operates in a plant at 1550 Short Tons Per Day (STPD) Ammonia production (1406 mt/d). It is the Base Case, the CO2 Removal System operates at an efficiency of 50,050 Btu/LbMole (27800 Kcal/KgMole) of CO2 removed. These operating conditions and the operating efficiency are typical of many aMDEA based systems.

Figure 2 Example 1
Activated MDEA (aMDEA) Based CO2 Removal System - Base Case at 1550 Short Tons/Day Ammonia Plant
(Modeled with WinSim, Inc. Design II)



C:\designii\samples\brochure\Crude.psd

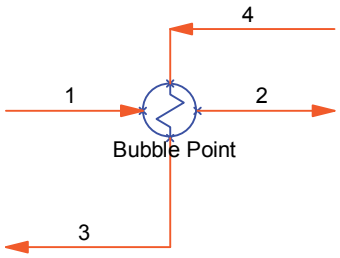
The purpose of this simulation is to determine the quality of products which can be made from a series of crude distillation columns for a feed of 100000 barrels per day. The amount of duty required for feed preheat (including feed furnaces for the atmospheric and vacuum columns) will also be calculated.



C:\designii\samples\brochure\Exchgr14.psd

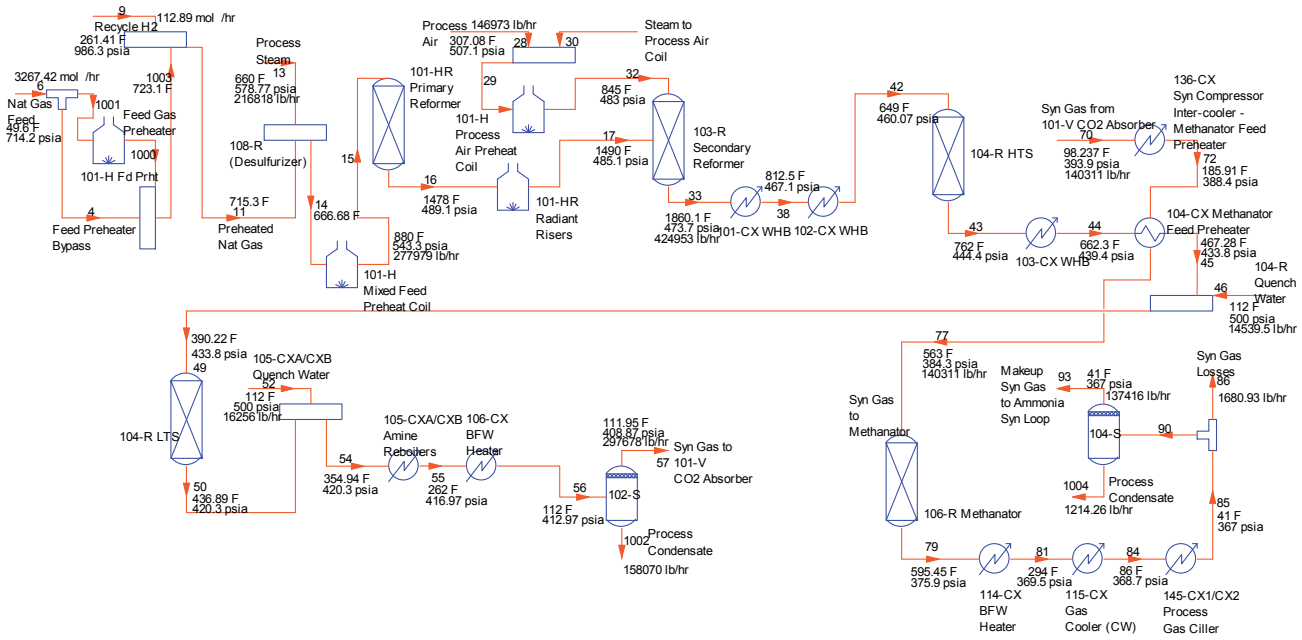
The purpose of this simulation is to determine the quality of products which can be made from a series of crude distillation columns for a feed of 100000 barrels per day. The amount of duty required for feed preheat (including feed furnaces for the atmospheric and vacuum columns) will also be calculated.

COUNTER-CURRENT: EXCHANGER RATING



C:\designii\samples\brochure\Frontend.psd

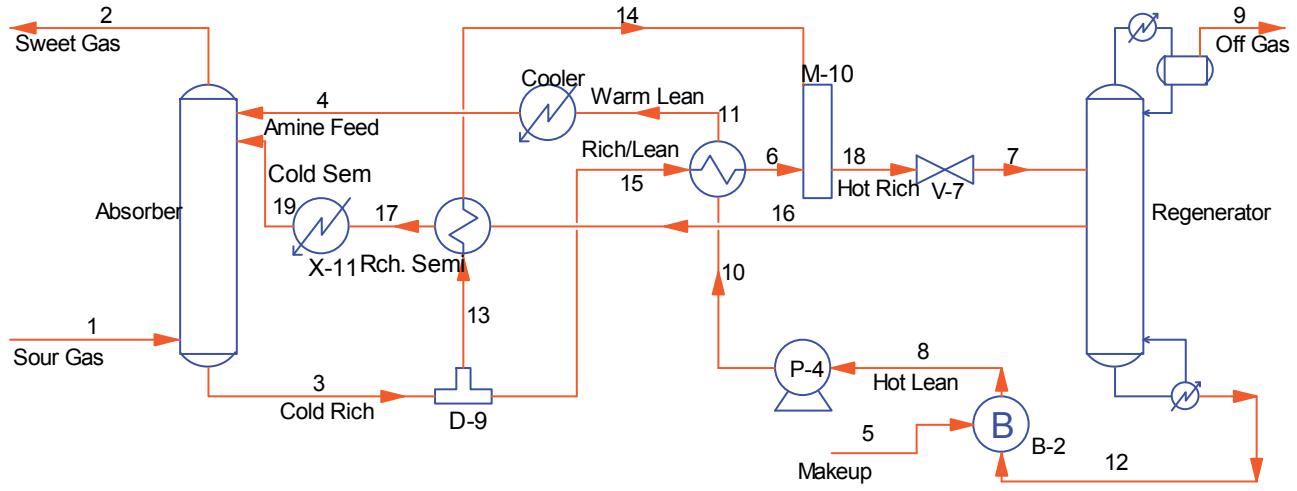
Design II provides thermodynamic, equilibrium and component material balance calculations for all ammonia plant technologies including processes which incorporate various combinations of feedstock preheating, steam-hydrocarbon and auto-thermal reforming, waste heat steam generation, carbon-monoxide shifting, boiler water preheating, carbon dioxide removal, methanation, synthesis gas cooling and condensate recovery.



Ammonia Plant Front-End Process Simulation at 1550 TPD Plant Rate

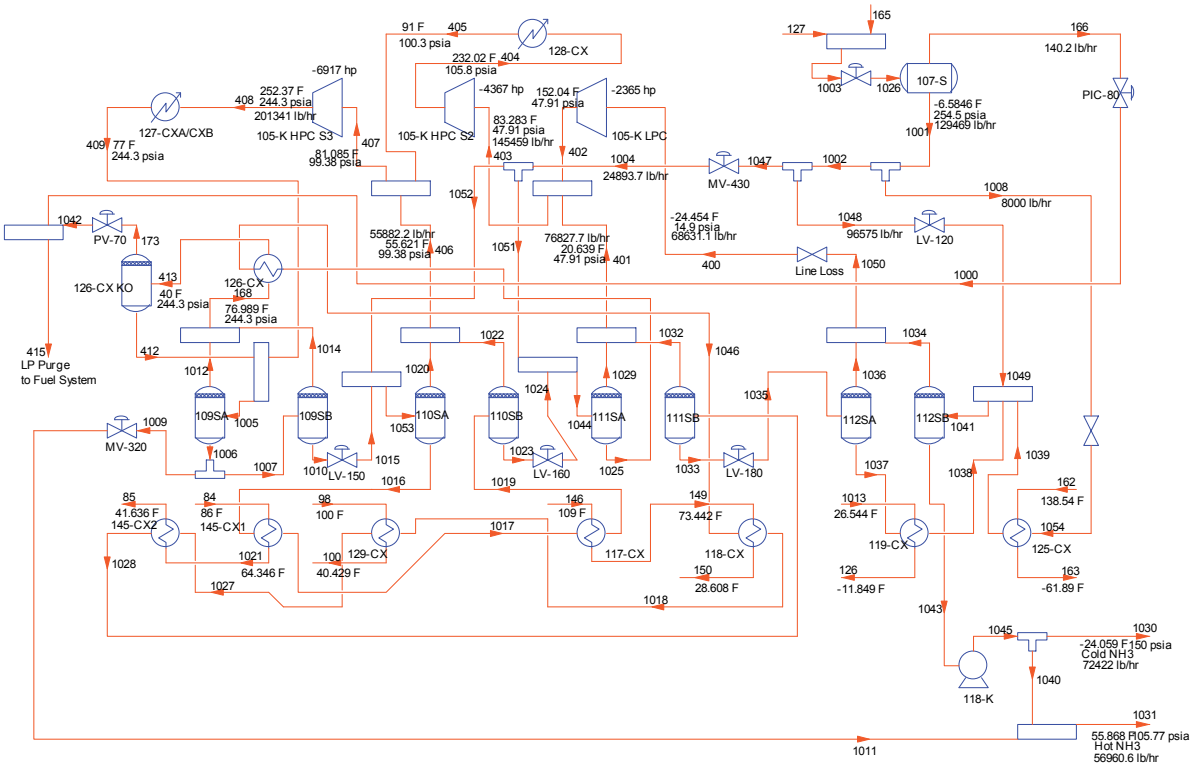
C:\designii\samples\brochure\MxpIntsd.psd

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



C:\designii\samples\brochure\Nh3refrg.psd

Design II rigorously models energy usage, vapor-liquid equilibria, thermodynamics and component balances for multi-stage refrigeration systems using inter-cooled compressors rejecting heat from process gas chillers via refrigerant ammonia condensing, including ammonia product let-down, flash drum equilibria and loads, inerts purge gas rejection, hot/cold ammonia product pumps and flash drum liquid and vapor flows.

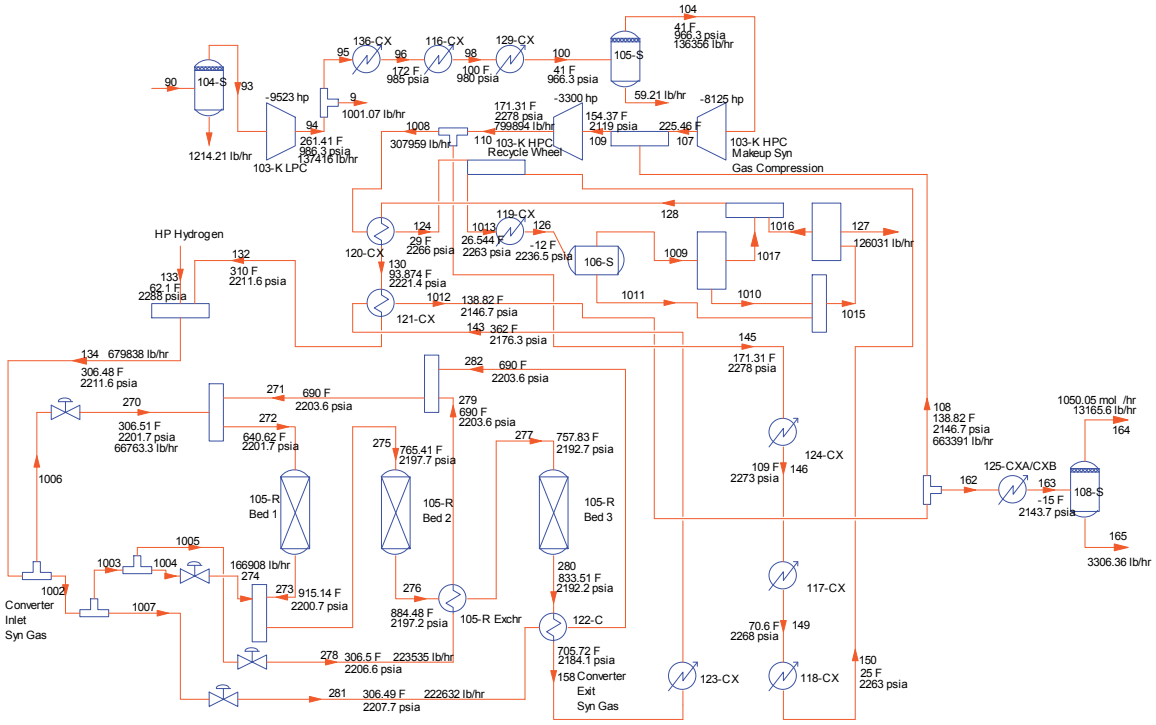


AMMONIA REFRIGERATION SYSTEM SIMULATION AT 1550 TPD

C:\designii\samples\brochure\Nh3synth.psd

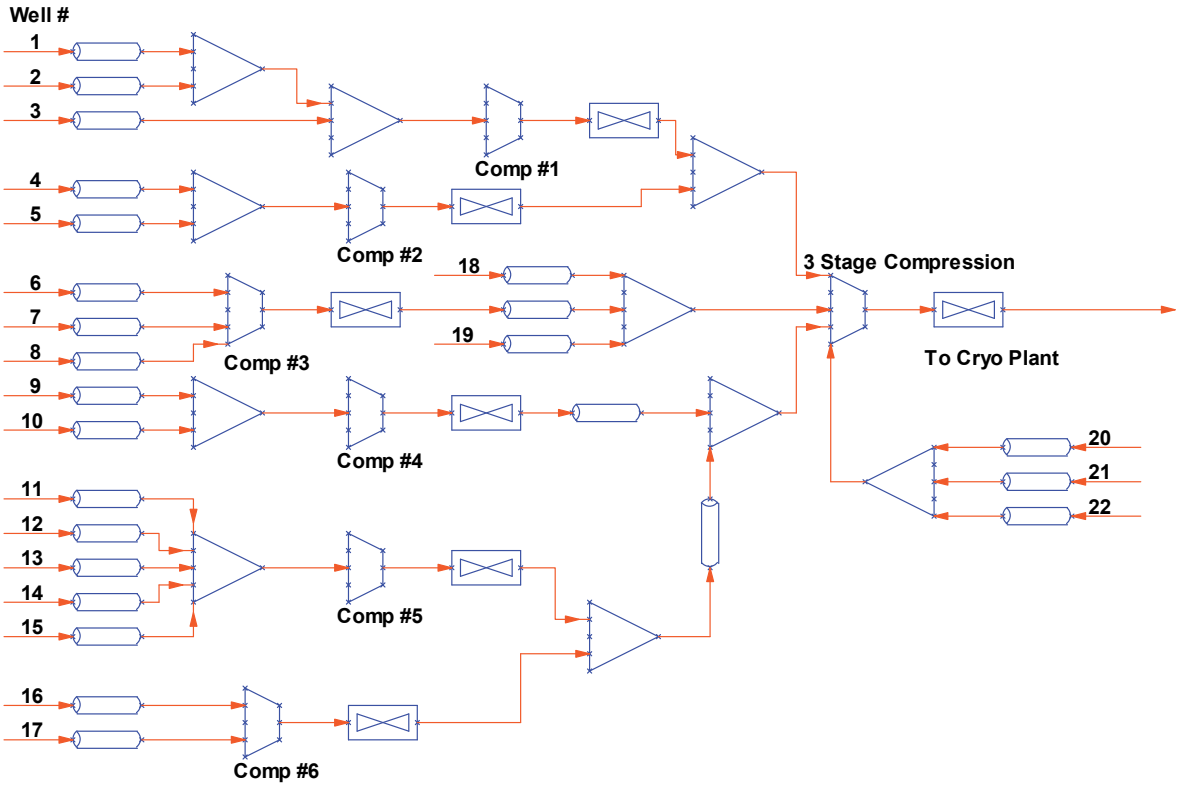
Design II readily handles ammonia synthesis equipment calculations, including predictions of energy usage, reaction equilibria, thermodynamics, component balances and vapor-liquid equilibrium for complex plant arrangements and technologies which incorporate synthesis gas multi stage compressors with inter-cooling, synthesis gas cooling using water and refrigeration chilling, refrigeration recovery exchange, ammonia product separation, process heating and cooling, inter-cooled and quench ammonia converters with internal exchangers, steam generation and boiler water preheating, synthesis gas purge and the synergies of recycle of synthesis gas.

AMMONIA SYNTHESIS PROCESS SIMULATION AT 1550 TPD RATE



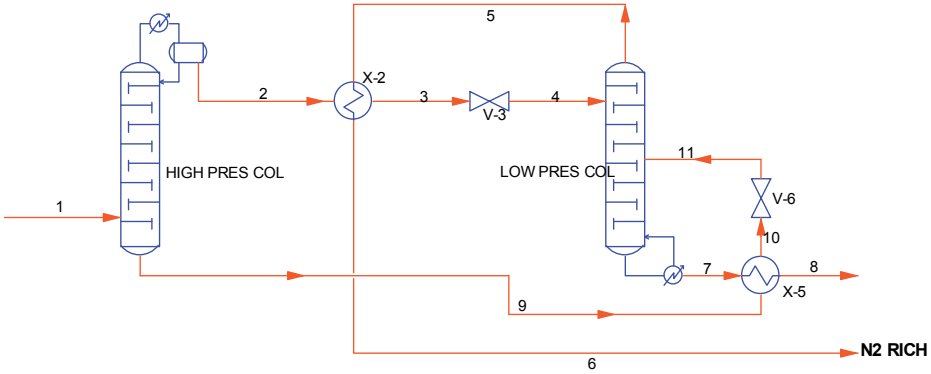
C:\designii\samples\brochure\Northrid.psd

Sample illustrating a gathering plant with multiple compressor stations.



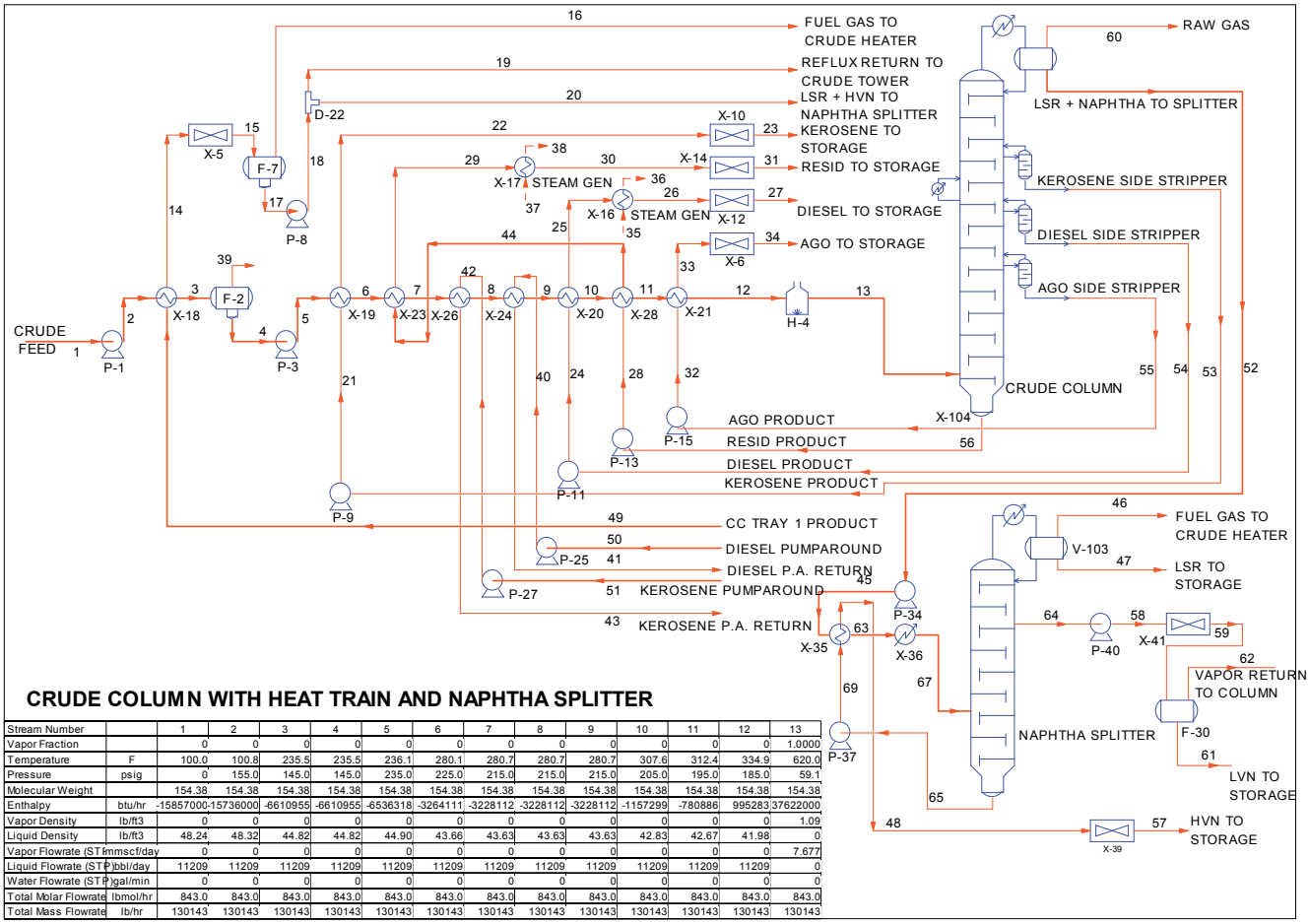
C:\designii\samples\brochure\NRU.psd

This sample simulation illustrates a scheme for rejecting pure Nitrogen from a natural gas mixture. The flowsheet uses a high pressure (up to 377 psia) column to process a natural gas mixture feed and the top & the bottom products are fed to the low pressure (about 37 psia) column. Pure nitrogen is rejected to the top of the low pressure column.



C:\designii\samples\brochure\Refinery.psd

This is a general purpose flowsheet for a typical refinery operation. Shows a combined simulation of a pre-heat train and light and heavy naptha refine columns.

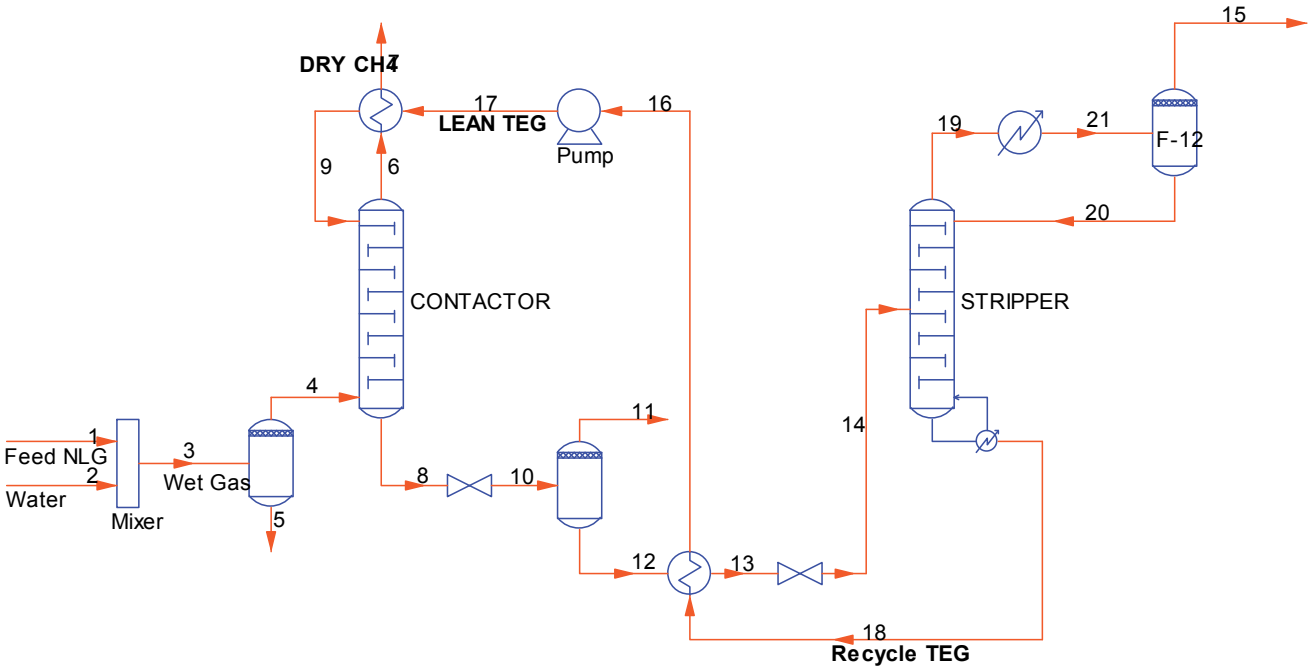


CRUDE COLUMN WITH HEAT TRAIN AND NAPHTHA SPLITTER

Stream Number		1	2	3	4	5	6	7	8	9	10	11	12	13
Vapor Fraction		0	0	0	0	0	0	0	0	0	0	0	0	1.0000
Temperature	F	100.0	100.8	235.5	235.5	236.1	280.1	280.7	280.7	280.7	307.6	312.4	334.9	620.0
Pressure	psig	0	155.0	145.0	145.0	235.0	225.0	215.0	215.0	215.0	205.0	195.0	185.0	59.1
Molecular Weight		154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38	154.38
Enthalpy	btu/hr	-15857000	-15736000	-6610955	-6610955	-6536318	-3264111	-3228112	-3228112	-3228112	-1157299	-780886	995283	37622000
Vapor Density	lb/ft3	0	0	0	0	0	0	0	0	0	0	0	0	1.09
Liquid Density	lb/ft3	48.24	48.32	44.82	44.82	44.90	43.66	43.63	43.63	43.63	42.83	42.67	41.98	0
Vapor Flowrate (STP)	mmscf/day	0	0	0	0	0	0	0	0	0	0	0	0	7.677
Liquid Flowrate (STP)	bbl/day	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	11209	0
Water Flowrate (STP)	gal/min	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Molar Flowrate	lbmol/hr	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0	843.0
Total Mass Flowrate	lb/hr	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143	130143

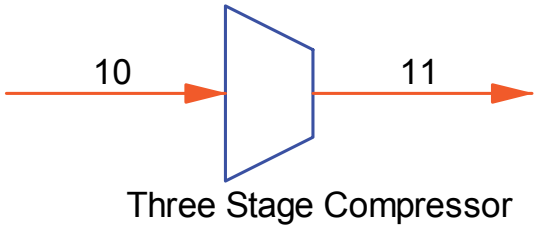
C:\designii\samples\brochure\TEGBTEX.psd

The sample problem is a simulation scheme for tri-ethylene glycol/water dehydration for BTEX system. The objective is to evaluate the amount of TEG recycled in order to dehydrate a mixture of Benzene, Toluene, Ethyl benzene, o-Xylene, (BTEX) and methane gas to sufficient purity often encountered in gas processing industry.



Chapter 7: Case Study Samples - c:\designii\samples\casestdy

Case1.psd – Case Study Example for Three Stage Compressor

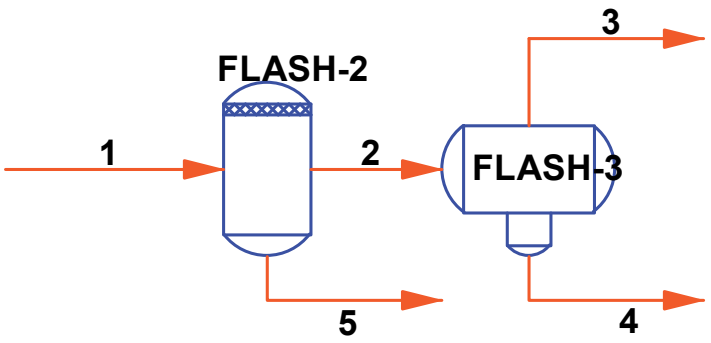


The simulation is set up to run a case study of a multi-stage compressor for various second stage pressures with 25 psia increments and to change the intercooler temperature from base case 125 F to 120 F. All properties of stream 10 are known. Each case is tabulated and plotted based on the temperature of stream 11, calculated work, and polytropic coefficient.

```
AB123.
*
C-
C-  ****************************************
C-  SEVERAL PRINT CONTROL COMMANDS HAVE BEEN
C-  ADDED TO MINIMIZE THE AMOUNT OF OUTPUT
C-  GENERATED BY CASE STUDIES AND LARGE RECYCLE
C-  PROBLEMS. YOU NOW HAVE CONTROL OF THE
C-  STREAM SUMMARY PAGE PRINT - ALL, NONE, OR
C-  SPECIFIC STREAM NUMBERS.  THESE NUMBERS CAN
C-  BE ENTERED IN THE ORDER YOU WISH THEM
C-  PRINTED, NOT JUST ASCENDING ORDER.  FOR
C-  THIS EXAMPLE BOTH THE STREAM AND DETAILED
C-  STREAM SUMMARIES WILL BE ELIMINATED.
C-  ****************************************

COMPRE 76 = STAGES, 10, -11,
       PRESSURE OF STAGES  = 150, 300, 600
       STAGES = 3, PRESSURE OUT = 600
       TEMPERATURE OF INTERCOOLER = 125
       CALCULATE POL
       EFFICIENCY = .70
C-
C-GENERAL
GEN,
COM=46,49,2,3,4,5,6,7,8,10,11,12,13,62
C-  THERMODYNAMIC OPTIONS
SOAVEK,COPE
C-  FEED CONDITIONS
FLO(FRA)10=.003,.011,.781,.105,.062,.007,.017,.003, .004,.002,.0016,.0003,.0001,.003
TOTFLO(MMSCF/DAY)10=300
TP10=100,50
C-  PRINT CONTROL
PRI SUM = NONE, PRI STR = NONE
C-  ****************************************
C-  A CASESTUDY FACILITY HAS BEEN ADDED TO THE
C-  DESIGN II PROGRAM.  THIS EXAMPLE SHOWS
C-  BOTH STEP AND CHANGE COMMANDS. SEE THE
C-  CASEHXER.IN AND .OUT FILES FOR FURTHER
C-  INFORMATION.
C- ****************************************
CASESTUDY
CASEBLOCK,NAME=PRESTA
CHANGE,COMPRE76,EFF=.75 STE,COMPRE76,PRESTA2=325,INC=25,LIM=425
CASE, NAME = TEMINT1
CHANGE, COMPRE76, TEM INT=120,RESTORE TABLE,STREAM 11,TEMPERATURE TABLE,COMPRE76,PRESTA2
TABLE,COMPRE76,CALWOR
TABLE, COMPRESSOR 76, CAL ELE DRIVE
TABLE, COMPRESSOR 76, CAL POL PLOTX,COMPRE76,PRESTA2
PLOTY,STR11,TEM
```

```
PLOTX, COMPRE76,PRESTA2
PLOTY, COMPRE76, CAL WOR
END
```

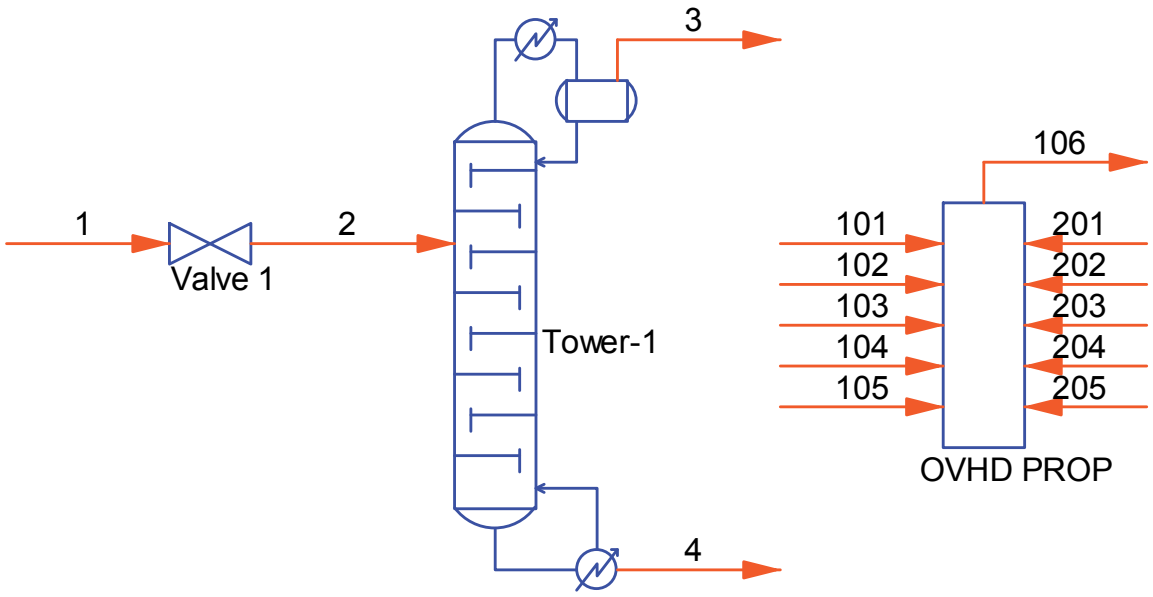
Case2.psd – Case Study Using Step Option



This sample flowsheet for case study is for varying the temperature of Flash-2 from -30 F to -103 F with steps of -7.5 F. One simulation runs all the cases (25 cases limit per simulation). All cases are tabulated and plotted.

```
AB123.
*
FLASH 2 = SEPR, 1,-2
        TEM OUT = 90
        PRE OUT = 500
        ISOTHERMAL
FLASH 3 = SPLIT, 2, -3, -4
GENERAL
COM=46,49,2,3,4,5,6,7,8,10,
FLO 1 = 1393.28, 500.47, 6131.42, 419.69, 133.4, 40.36 46.95, 19.77, 15.65, 31.29
TP1=90,500, CON TOL=.01, APISOAVEK,APISOAVEH,COPED PRI STR=NONE
C-
C-  A CASESTUDY FACILITY HAS BEEN ADDED TO DESIGN II.
C-  A COMPLETE LIST OF THE MODULES AND SPECIFICATIONS,
C-  AND CALCULATED VALUES WHICH ARE AVAILABLE FOR USE
C-  WITH CASESTUDY IS CONTAINED IN THE NEW FEATURES
C-  BULLETIN - VERSION 5.0.
C-
C-  THIS EXAMPLE SHOWS THE USE OF THE STEP OPTION AND
C-  PLOT COMMANDS.
C-
CASE
CASEBLOCK,NAME=CONDENSING CURVE
STEP, FLASH2, TEM OUT=-30, INCREMENTS = -7.5, LIMIT=-103 C-
C-  ANY VARIABLE ADJUSTED IN A CASEBLOCK WILL BE INCLUDED
C-  AUTOMATICALLY IN THE TABLE GENERATED FOR THE CASESTUDY.
C-
TABLE,STREAM 2,VAP FRA
TABLE,STREAM 4,FLO COM 4
C-
C-  TABLE DATA MAY ALSO BE PRESENTED GRAPHICALLY THROUGH
C-  USE OF THE PLOT X AND PLOT Y COMMANDS.  THE PLOTX
C-  VARIABLE IS THE INDEPENDENT VARIABLE AND THE PLOTY
C-  VARIABLES (UP TO 4) ARE THE DEPENDENT VARIABLES. ALL
C-  PLOTY VARIABLES MUST HAVE THE SAME DIMENSIONAL UNITS.
C-
C-  MULTIPLE SETS OF PLOTX AND PLOTY COMBINIATIONS ARE
C-  ALLOWED PER RUN, BUT ONLY ONE CASESTUDY SECTION MAY
C-  BE DEFINED.
C-
PLOTX,FLASH 2,TEM OUT
PLOTY,STREAM 2,VAP FRA,LABEL=VAPOR FRACTION PLOTX,STREAM 2,VAP FRA,LABEL=VAPOR FRACTION
PLOTY, STREAM 4, FLOW COMPONENT 4, LABEL= MOLES OF C2 LIQ.
END
```

Case3.psd – Case Study With Inline Fortran

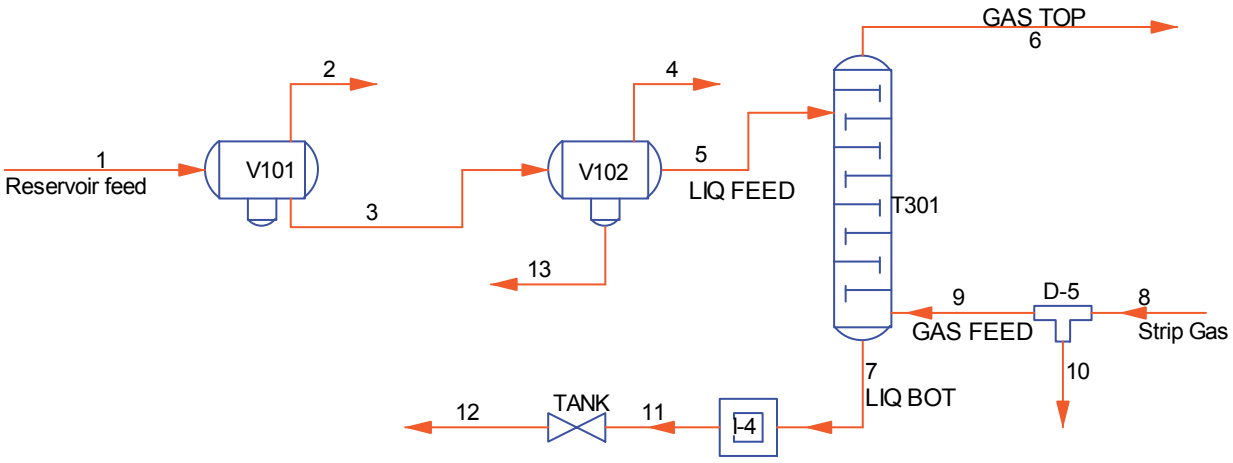


This example performs a case study by using inline fortran and the repeat Command. The objective is to run distillation column T1 at five sets of conditions for mass purity of ethane in the bottoms and tower Pressure.

```
AB123.
*
C-
C-************* CASE STUDY ********************
C- THIS EXAMPLE PERFORMS A CASE STUDY BY USING INLINE FORTRAN AND THE REPEAT=
C- COMMAND. THE OBJECTIVE IS TO RUN DISTILLATION COLUMN T1 AT FIVE
C- SETS OF CONDITIONS FOR MASS PURITY OF ETHANE IN THE BOTTOMS AND TOWER
C- PRESSURE. THE CASES ARE:
C-
C-      CASE #    PURITY OF C2 IN BOTTOMS     TOWER PRESSURE, PSIA
C-
C-         1         .001             345.0
C-         2         .003             345.0
C-         3         .003             334.7
C-         4         .005             334.7
C-         5         .007             300.0
C-
C- MASS PURITY IS THE RATIO OF THE MASS OF C2 IN THE BOTTOMS TO THE TOTAL MASS
C- IN THE BOTTOMS.
C-*******************************************************
VAL1= LETDOWN,1,-2, PRE OUT= 340
C-********************** FORTRAN ***************************
C- THE FOLLOWING FORTRAN STATEMENTS ARE EXECUTED BEFORE EQUIPMENT VAL1 IS CAL
C- CULATED. IT ESTABLISHES THE ONE DIMENSIONAL ARRAYS FOR THE PURITY SPEC
C- (SPECS) AND THE TOWER PRESSURES (PROPS). IT SETS UP TWO COUNTERS, ISTR AND
C- ISP. FINALLY, IT "SETS" THE OUTLET PRESSURE OF THE VALVE AND THE TOP
C- PRESSURE OF THE COLUMN.
C-*****************************************************
F-              PRE
F-              DIMENSION SPECS(5)
F-              DIMENSION PROPS(5)
F-              COMMON /MYVARS/ ISP,ISTR,SPEC
F-              DATA SPECS /.001, .003, .003, .005, .007/
F-              DATA PROPS /345.0, 345.0, 334.7, 334.7, 300.0/
F-              ISTR = ISTR + 1
F-              ISP = ISP + 1
F-              SPEC = SPECS(ISP)
F-              SET (VAL(1),PRE OUT) = PROPS(ISP)
F-              SET (DIS(2),PRE TOP) = PROPS(ISP) - 5.0
DIS2= T1,2,-3,-4,
        PARTIAL
        LOC FEE= 10, PRE TOP= 340
        DEL= 8, TEM TOP= 30, TEM BOT= 200
        TRA= 20, PUR BOT(MAS) 3= .001
```

```
        PRO= 0,260, REFLUX= 1.5
        SAV TEM, SAV VAP
        PRI SWI= 0
C-********************** FORTRAN **********************
C-THESE FORTRAN STATEMENTS EXECUTE "DURING" THE CALCULATION OF EQUIPMENT MOD
C-ULE T1. THE VARIABLE NAMED 'SUM' IS THE TOTAL MASS FLOW FROM THE BOTTOM OF
C-THE TOWER. THE VARIABLE 'C2FLOW' IS THE MASS FLOW OF ETHANE FROM THE BOTTOMS
C-. THE DEFINED PURITY SPEC EPS(1) NOW OVERRIDES THE PUR BOT(MAS)3= SPEC SET
C-ABOVE.
C-*****************************************************
F-              DURING
F-              COMMON /MYVARS/ ISP,ISTR,SPEC
F-              SUM=0.0
F-              DO 25 I=1,9
F-      SUM=FLOWL(I,1)*GET(BANK(AMW),POS(I))+SUM
F-  25 CONTINUE
F-      C2FLOW=FLOWL(3,1)*GET(BANK(AMW),POS(3))
F-      C2PUR=C2FLOW/SUM
F-      EPS(1)=(C2PUR -SPEC)/SPEC
MIX3=OVHD PROP,101,102,103,104,105,201,202,203,204,205,-106
C-********************** FORTRAN **********************
C-      THIS FORTRAN COPIES THE TOP AND BOTTOM PRODUCT STREAMS FOR EACH CASE:
C-      VAPOR PRODUCT STREAM RESULTS ARE STORED IN STREAMS 101 - 105
C-      LIQUID PRODUCT STREAM RESULTS ARE STORED IN STREAMS 201 - 205
C-*****************************************************
F-      PRE
F-      COMMON /MYVARS/ ISP,ISTR,SPEC
F-      JSTR = 100 + ISTR
F-      KSTR = 200 + ISTR
F-      CALL COPSTR(3,JSTR)
F-      CALL COPSTR(4,KSTR)
C-*****************************************************
C- GENERAL SECTION COMMANDS
C-*****************************************************
GENERAL,
COM= 49,2,3,4,5,6,7,8,10
FLOW 1= .2, 3.2, 156.7, 113.8, 38.9, 46.4, 20, 16, 30
TP (PSIG) 1 = 121.77, 350
C-****************************
C-  THERMODYNAMIC DATA
C-****************************
SOAVEK, SOAVEH, STDD
C-****************************
C-  CASE STUDY INFORMATION C-****************************
CAL SEQ = 1,2,3, REPEAT = 5
C-****************************
C-  PRINT RESTRICTIONS C-****************************
PRI STR= 1,101,102,103,104,105,201,202,203,204,205
END
```

Case4.psd – Case Study To Identify Optimum Liquid Feed Rate With Changing Flash Temperature



This flowsheet illustrates a case study analysis by changing reservoir feed rate as a change is made in flash V101 temperature. In all there are 12 cases to be run and all are tabulated.

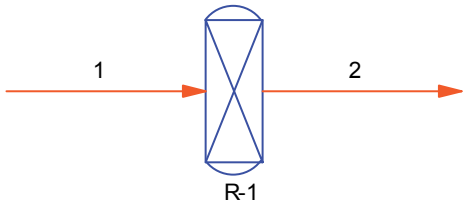
Chapter 8: CEP Magazine October 2005 Samples - c:\designii\samples\cep_oct2005

Layer 1 – Reactor modeling.psd – onion model layer 1

This series of simulation files is to aid the novice in process simulation in developing a reasonable good flowsheet based on process synthesis approach. Process synthesis approach utilized here is the Onion Model introduced by Linnhoff et al. (1982). Full description of the concept can be found in "Foo, Manan, Selvan and McGuire (2005), Integrate Process Simulation and Process Synthesis, Chemical Engineering Progress, October 2005, pp. 25-29".

The heart of the process - the reactor

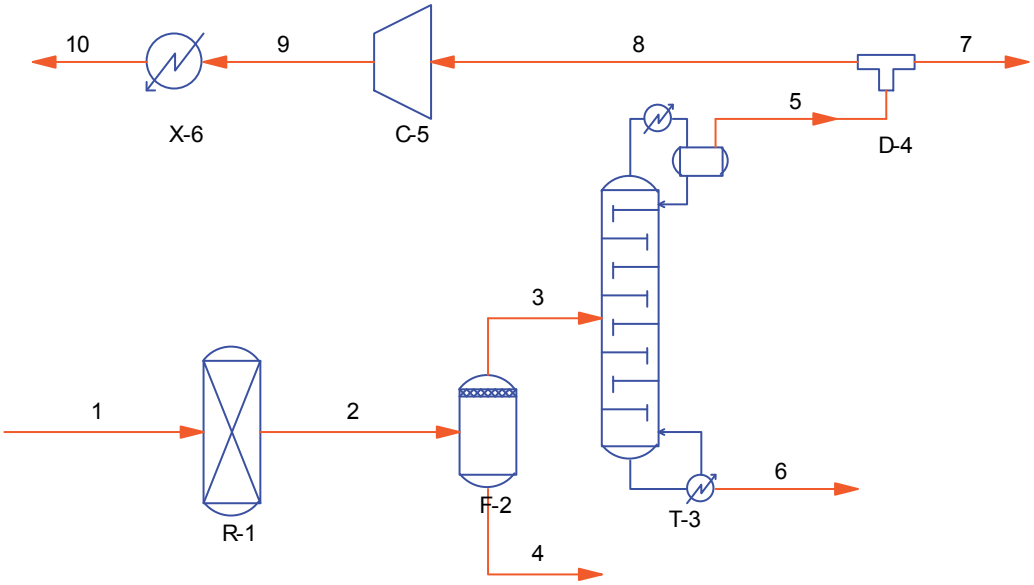
Case study: Production of n-Octane
Reaction path: 2 C2H4 + C4H10 ---> C8H18
Feed stream condition: 30C; 20 psia
Feed flowrates (k-mol/h): 20 C2H4, 10 n-C4H10, 0.5 i-C4H10, 0.1 N2
Reactor condition: operating T - 93C, 98% conversion, P drop 5 psi



Layer 2a – Without recycle.psd – onion model layer 2

Separation and recycle (before tear stream is applied).

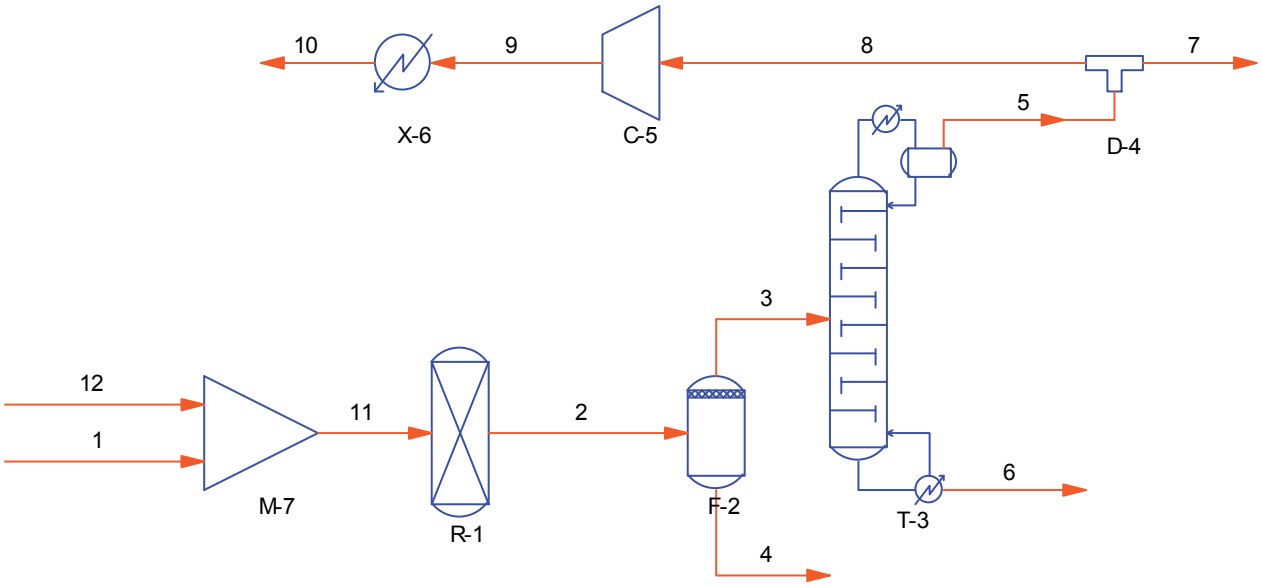
Separation units are added to segregate unreacted raw material from the product of n-octane.
Flash unit: P drop - 2 psi, operating T - 93C
Distillation: 10 trays, 15 psia, n-octane recovered at bottom stream
Compressor: P out - 20 psia, heat exchanger: outlet T 93C



Layer 2b – Tear stream.psd – onion model layer 2

Separation and recycle (with tear stream).

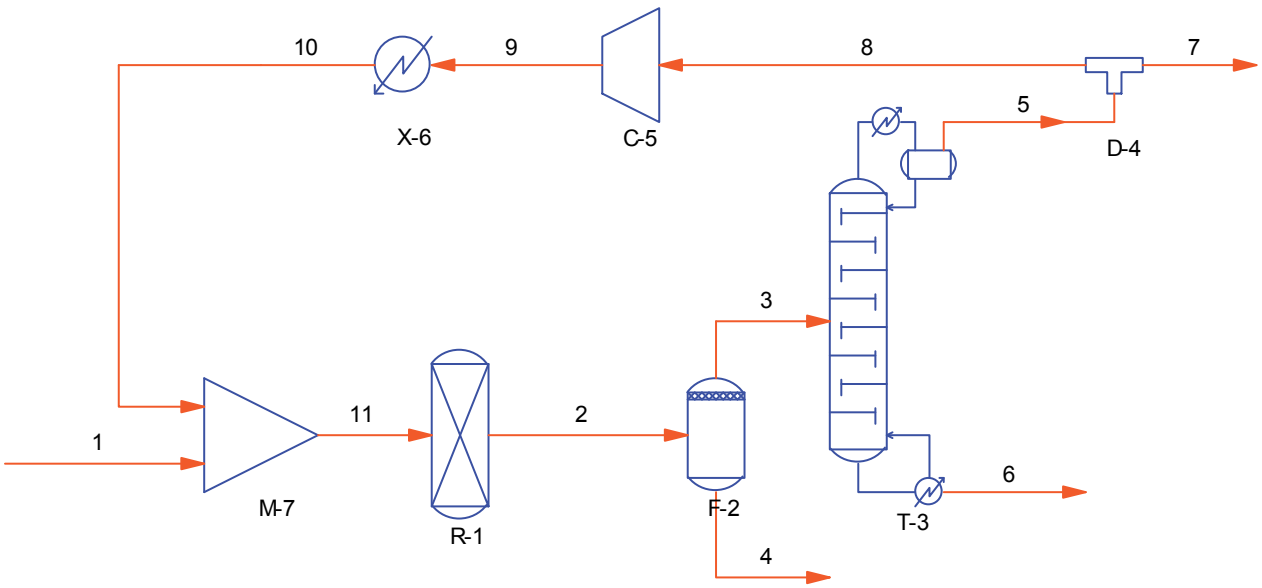
Tear stream is used to facilitate the simulation of recycle system. In this case, Stream 12 is the "tear". User can enter the data for this tear stream based on the simulation result of the stream to tbe recycle, i.e. Stream 10 in this case. The original feed stream (Stream 1) values are adjusted according to take care of the mass balance. The simulation is executed again.



Layer 2c - Recycle.psd – onion model layer 2

Separation and recycle (tear stream removed).

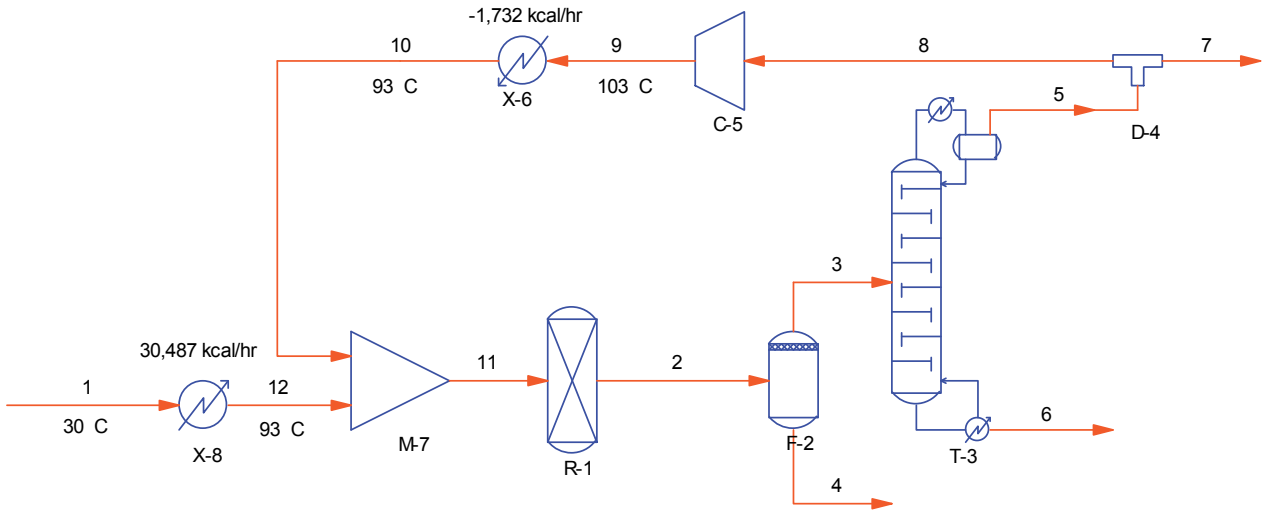
Tear stream is removed and recycle stream is connected to the system. Simulation is executed.



Layer 3a – Identify heat sink and source.psd – onion model layer 3

Heat exchanger network.

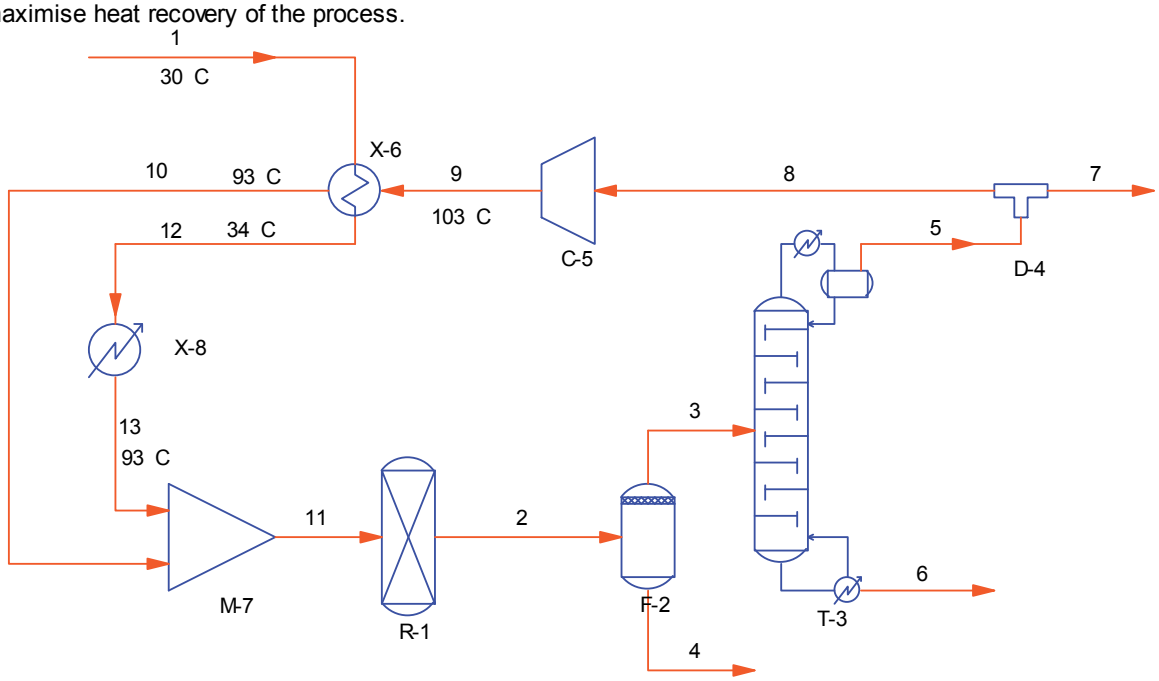
Determine the heat load required to heat or cool a process stream. A new heat exchanger is added to heat the feed stream from 30C. Hence there is 1 hot stream (stream to be cooled by X-6) & 1 cold stream (stream to be heated by X-8) in this process.



Layer 3b – Heat integration.psd – onion model layer 3

Heat exchanger network.

Using pinch analysis (Linnhoff et al, 1982; Seider et al., 2003; Smith, 2005), the heat exchanger network may be designed to maximise heat recovery of the process.



Chapter 9: ChemTran Samples - c:\designii\samples\chemtran

Acwtrct.in - Acetone-Water-1,1,2 TCE VLE Data Regression

VLE date regression example for acetone, water, and 1,1,2 tri chloro ethylene.

Chemtran.psd – Sample Flowsheet Showing ChemTran Data in a Flowsheet

A sample flowsheet showing how to add new components to a flowsheet.

Chemtran1.psd – Adding Components to Flowsheet Using Point Properties

A sample flowsheet showing how to add new components using point properties to a flowsheet.

Chemtrn1.in - Guidelines for Regressing Equilibria Data

Regressing equilibria data with ChemTran is often necessary in order to model chemical systems that exhibit non-ideal behavior. The following guidelines are presented to help you obtain the best possible data regressions.

Step 1 Find Data
Locate equilibria data for all binary pairs in your mixture. If data is unavailable for some of the binary pairs, you may consider using UNIFAC or Regular Solution Theory to estimate infinite dilution activity coefficients.

Step 2 Select Input Commands for Data
Decide what ChemTran Mixture Commands should be used to enter the data into your input file. See ***ChemTran: Correlating Mixture Data*** for further details.

Step 3 Select Objective Function
Select the objective function for your regression. The default objective function is the sum of the K-values. This is good for most situations. See ***ChemTran: Table 6.2. Objective Functions for reduction of VLE data***.

Step 4 Select Thermodynamic Correlation
Decide which thermodynamic correlation the data will be regressed to. The available choices are presented in ***ChemTran: K Value Options***. A general guideline to follow is that an equation of state should be selected for non-condensable gases and an activity coefficient method is suitable for liquid phase systems at low pressures. I your data includes two liquid phases your only choices are RENON or UNIQUAC.

Often times you will have several correlations that are suitable. If you are seeking the highest degree of accuracy you can consider regressing the data to several correlations in order to see which correlation provides the best data fit.

Step 5 Do preliminary Regression
Regress the data using the default interaction parameter calculations. The default calculations fit the data by calculating the 'A' interaction parameters such that the objective function is minimized. The 'B' parameters are set to zero. See the following input file and selected output.

ChemTran Input Default - Fit A interaction Parameters, B parameters Fixed to 0

```
AB123.
*VLE/LLE DATA REGRESION
COM=71,62

T-X(MMHG,C) 62,71=760,
       152.30, 0.000,
       146.30, 0.0100,
       96.40,  0.0550,
       95.00,  0.0810,
       95.20,  0.9800,
       96.70,  0.9850,
       98.00,  0.9960,
```

```
               99.30,  0.9980
               100.00, 1.0000,
        UNIQUACK
        END

Selected Portions of ChemTran Output

* * * *REDUCTION OF PHASE EQUILIBRIUM DATA FOR SYSTEM 1 * * * *
               COMPONENTS IN SYSTEM 1
        NO.     CODE NO.       NAME
-----------------------------------------------------------------------
         1           71 I-C3-BENZENE
         2           62 WATER

EQUATION OPTION IS UNIQUAC NUMBER OF ITERATIONS= 20
SUM OF SQUARES OF THE OBJECTIVE FUNCTION= 0.1055281E 00
STANDARD ERROR OF ESTIMATE= 0.8121273E-01
ROOT MEAN SQUARE OF ERROR= 0.7656810E-01



        CHEMSHARE CORRELATION CONSTANTS FOR ALL BINARY PAIRS
        FROM UNIQUAC EQUATIONS (BTU/LBMOL)
               DATA                    AVE TEMP
BINARY PAIR     ORIGIN  CONSTANT       VALUE         (F)
-----------------------------------------------------------------------
I-C3-BENZENE(71)+
WATER (62)      DATA    A(71,62)       3195.31 227.84
                              A(62,71)       -586.948
```

Step 6 Evaluate the Quality of the Regression

Once you have executed ChemTran evaluate the quality of your data fit by examining the output value labeled 'STANDARD ERROR OF ESTIMATE'. This is the standard deviation of the error of the objective function and is a measure of how well the correlation was fit to the data. It is desirable to have a standard error of estimate near 5% (.05), however, a greater error may be acceptable. In the case shown in Step 5 the standard error of estimate is 8.13%.

Step 7 Improve Data Fit

If you require a better data fit than what was obtained in the preliminary regression, you will want to take steps to improve the data fit. The options you can consider are:

- Regress the data to another thermodynamic correlation
- Fit the temperature dependent 'B' parameters
- Select a different data set to regress
- Discard data points that are inconsistent with the regression

The following ChemTran input file and selected output demonstrates how to fit the B parameters to the data.

```
        AB123.
        *VLE/LLE DATA REGRESSION
        COM=71,62
        C-
        C- VARY B PARAMETERS
        VARY 62,71=B12,B21
        C-ENTER STARTING GUESSES FOR A PARAMETERS FROM PRELIMINARY REGRESSION
        UNIQUACK (BTU/LBMOL,F) 71,62=227.84,3195.31,-586.958
        T-X(MMHG,C)62,71=760,
               152.30, 0.0000,
               146.30, 0.0100,
                 96.40,       0.0550,
                 95.00,       0.0810,
                 95.20,       0.9800,
                 96.70,       0.9850,
                 98.00,       0.9960,
                 99.30,       0.9980,
               100.00, 1.0000,
        UNIQUACK
        END

Selected portions of ChemTran output.
* * * *REDUCTION OF PHASE EQUILIBRIUM DATA FOR SYSTEM 1 * * * *
               COMPONENTS IN SYSTEM 1
        NO.     CODE NO.       NAME
-----------------------------------------------------------------------------------------------------------
         1           71 I-C3-BENZENE
         2           62 WATER

EQUATION OPTION IS UNIQUAC
NUMBER OF ITERATIONS=
```

```
47 SUM OF SQUARES OF THE OBJECTIVE FUNCTION= 0.270569E-02
STANDARD ERROR OF ESTIMATE= 0.1390193E-01
ROOT MEAN SQUARE OF ERROR= 0.1226035E-01


       CHEMSHARE CORRELATION CONSTANTS FOR ALL BINARY PAIRS
       FROM UNIQUAC EQUATIONS (BTU/LBMOL)
              DATA                  AVE TEMP
BINARY PAIR    ORIGIN  CONSTANT      VALUE        (F)
-------------------------------------------------------------------------------------------------------
I-C3-BENZENE(71)+
WATER (62)     DATA    A(71,62)      24061.3 227.84
                       A(62,71)      -1655.94
                       B(71,62)      -14.6924
                       B(62,71)      0.573410
```

Fitting the B parameters in addition to the A parameters resulted in the data fit improving from 8.12% to 1.39%.

Chemtrn2.in - Entering Properties to Define a Custom Component

The following input file illustrates how ChemTran can be used to define properties for a component not included in the DESIGN II or ChemTran component database. There are 7 commands required to define a custom component. If the following commands are entered, all other thermodynamic and physical properties can be calculated. Although the following input file defines the minimum set of commands necessary to define a custom component, you may enter data for any of the properties listed in ***ChemTran: Pure Component Command Details***.

```
AB123.
*CHEMTRAN DEMONSTRATION
COM=200
C-NON-DATABASE COMPONENT NAMES
C-
NAM200=2.3.4-TRI-C1-C5
C-
C-PURE COMPONENT DATA FOR COMPONENT 200
C-ENTER MOLECULAR WEIGHT, NORMAL BOILING POINT,
C-CRITICAL TEMPERATURE, CRITICAL PRESSURE,
C-VAPOR PRESSURE DATA, CRITICAL VOLUME,
C-AND IDEAL GAS HEAT CAPACITY DATA
MOL 200=114.22
TB(C) 200=113.47
TC(F) 200=559.47
PC(ATM) 200=26.94
P-T (C,MMHG) 200=36.57,300.0,47.71,36.57,760,113.47,
       20473.9,293.04
VC (FT3/LBMOL) 200=7.38
CP-T (R,DAT,T**3) 200=500,2800,45.35,540,57.52,720,68.22,900,
       77.56,1080,85.65,1260,92.6,1440,98.52,1620,
       103.51,1800,107.7,1980,111.1,2160,114.07,
       2340,116.47,2520,118.5,2700
FILE NEW = 234T,CHEM,DATA
       END
```

Chemtrn3.in - Entering Data for Pseudo Components (100) and Chemical Components (200)

Input file showing data entry for Pet-100 (pseudo components) and Chem-200 (chemical) level components for inclusion in DESIGN II database.

```
AB123.
*CHEMICAL COMPONENTS--CHARCOAL, ASH,& CARBON (200 LEVEL)
*AND PSEUDOCOMPONENT(100-104)

C-  CHAR, ASH, AND CARBON
COM=1,2,3,22,40,46,47,48,49,50,62,63,200,201,202
    100,101,102,103,104
STDK
LAT
C-FILE NEW=ASH,CHAR,DATA

C-*************************CHAR*****************************
MOL 200=12
NAM 200=CHAR
```

```
STRU 200=C1=C2
TB(F)200=-200
TC(F)200=-100
PC(PSIA)200=125
VC 200=
SOL 200=1
CP-T(K) 200=200,2000,1.8,298,2.7,400,3.15,500,3.68,600,
     3.96,700,4.27,800,4.74,1000,5.04,1200,5.34,1400
     5.80,1600
K-T(CON,R) 200=50,5000,.1E-6,0.0,0.0
L-T(BTU/LBMOL,F) 200=0,2500,0.0001,100,0.0001,1000,0.0001,2500
V-T(LBMOL/FT3,F) 200=0,2500,8.3333,0.,8.3333,2500,8.3333,-200

C-**************************ASH*****************************
MOL 201=116
NAM 201=ASH
STRU 201=SI1-C2,C2=O6,SI1-C3,SI1-C4,C4=O7,SI1-C5
TB(F)201=-200.
TC(F)201=-100.
PC(PSIA)201=1622.8
VC 201=
SOL 201=1.0
CP-T(K)201=200,2000,21.37,298,23.23,400,24.85,500,26.36,600,
     28.11,700,29.73,800,32.98,1000,36.24,1200,39.49,1400,
     41.50,1600
K-T(CON,R)201=50,5000,0.1E-6,0.0,0.0,
L-T(BTU/LBMOL,F)201=0,2500,0.0001,0,0.0001,1000,0.0001,2500
V-T(LBMOL/FT3,F)201=0,2500,1.5603,0,1.5603,1000,1.5603,-200

C-**************************CARBON**************************
NAM 202=CARBON
MOL 202=12.01
STRU 202=C1=C2
TB(F) 202=-200.
TC(F) 202=-100.
PC(PSIA)202=1216.7
VC (FT3/LBMOL) 202=1.25
SOL(CAL/CM3)202=1.0
K-T (CON,R)202=50,5000,0.1E-6,0.0,0.0,
L-T (BTU/LBMOL,F)202=0,2500,0.0001,0,0.0001,1000,0.0001,2500
V-T (LBMOL/FT3,F)202=0,2500,11.698,0,11.698,1000,11.698,-200
CP-T(K)202=200,2000,2.038,298,2.85,400,3.5,500,4.04,600,
     4.44,700,4.74,800,5.15,1000,5.43,1200,5.61,1400,
     5.72,1600

C-**********************LIGHT OIL 1**************************
API(API)100=3.0
AMW 100=112
AMB (F) 100=350.
NAM 100 =LIGHT OIL1

C-**********************LIGHT OIL 2**************************
API(API)101=-10.5
AMW 101=175
AMB(F)101=555.
NAM 101=LIGHT OIL2

C-**********************LIGHT OIL 3**************************
API(API)102=-14.4
AMW 102=215
AMB (F) 102=680.
NAM 102 =LIGHT OIL3

C-**********************LIGHT OIL 4**************************
API(API)103=-17.9
AMW 103=305
AMB(F)103=880.
NAM 103=LIGHT OIL3

C-**********************LIGHT OIL 5**************************
API(API)104=-19.3
AMW 104=410
AMB(F)104=170.
NAM 104=LIGHT OIL5

END
```

Chemtrn4.in - Entering Data for Solid Components (300)

Input file showing data entry for Chem-300 level (solid-carbon) component for inclusion in DESIGN II database.

```
AB123.
*DATA FOR CARBON as a solid

C-  Components numbered 300-310 are allocated as solids by DESIGN II
COM=300

C-  Name the component
NAM 300=CARBON

C-  Volume of Solid data
VS-T (DATA) 300=0,2500,0.0854847,0,0.0854847,1000,0.0854847,-200

C-  Molecular Weight data
MOL 300=12.01

C- Enthalpy of Solid data if available
HS-T(BTU/LBMOL,K)300=200,2000,2.038,298,2.85,400,3.5,500,4.04,600,
    4.44,700,4.74,800,5.15,1000,5.43,1200,5.61,1400,5.72,1600

C- Include the following line for DESIGN II accessible data file
C- A file named 'carbonxx.dat' will be saved in the same sub-directory.

FILE NEW=CARB,ONXX,DATA
END
```

Chemtrn5.in - Entering Experimental Or Literature Values For Binary Interaction Parameters

Input file showing binary interaction parameter data entry for a component system using available correlations. No VLE/LEE data regression is needed.

```
AB123.
C-  TRANS 2 BUTENE, DIMETHYL FORMAMIDE
COM =   26,       4136
RENON
RENON (BTU/LBMOL,F) 26,4136=122,2604.34,-180.706,0,0
FILE NEW = T2BT,DMFD,DATA
END
```

Chemtrn6.in - Calculating Binary Interaction Parameters by Regressing VLE Data

Input file showing VLE data entry to back out binary interaction parameters based on a chosen non-ideal thermodynamic (Wilson) expression.

```
AB123.
*ACETONE-WATER
FILE NEW=DATA,ACET,H20
WILSON K
C- ACETONE, WATER
COMP=1002,   62
C-
T-X-Y(MMHG,C)1002,62=760,  76.25, .0490, .6050
        72.40, .0690, .6780
        66.45, .1270, .7530
        62.75, .2000, .7900
        62.34, .2760, .8040
        60.90, .3100, .8260
        60.17, .4950, .8320
        58.80, .6500, .8550
        58.47, .6940, .8620
        57.95, .7500, .8760
        57.25, .8290, .9030
        56.75, .9000, .9360
END
```

Chemtrn7.in - Calculating Binary Interaction Parameters from Infinite Dilution Activity Coefficients

Input file showing a method to back out binary interaction parameters (any correlations) using infinite dilution activity coefficients of the binary. No VLE/LLE data required.

```
AB123.
C- ETHANOL, DIETHEYL ETHER
COM= 1012,   1005
FILE NEW = ETHO,DETE,DATA
WILSON
ACT(C) 1005,1012 = 40,2.63,3.92
END
```

Chemtrn8.in - Calculating Binary Interaction Parameters from Infinite Dilution Activity Coefficients Estimated from UNIFAC

Input file to back out binary interaction parameters using infinite dilution activity coefficients of the binary using UNIFAC (estimation method). No VLE/LLE data required.

```
AB123.
C- ETHANOL, ACETONE
COM= 1012, 1002
FILE NEW = ETHO,ACET,DATA
WILSON
ACT(UNIFAC,C) 1002,1012 = 40
END
```

Chemtrn9.in - Calculating Binary Interaction Parameters from Infinite Dilution Activity Coefficients using Regular Solution Theory

Input file to back out binary interaction parameters using infinite dilution activity coefficients of the binary using Regular Solution Theory. No VLE/LLE data required.

```
AB123.
C- 1,3 BUTADIENE, 1 BUTENE
COM = 28,24
WILSON
SCA
FILE NEW = 13BD,1BUT,DATA
END
```

Chemtrn10.in - Using the Structure Command to Define a Custom Component.

When part or all of the physical property data required to define a custom component is unavailable, you may wish to use ChemTran's group contribution methods to predict these properties. ChemTran's universal structure command is used to enter the molecular structure of the custom component and ChemTran will automatically determine what groups are present and predict all of the required physical properties. The following shows how the structure command is used to enter the component.

```
AB123.
*CHEMTRAN DEMONSTRATION
COM=204
C-NON-DATABASE COMPONENT NAMES
C-
NAM 204=2.3.3-TRI-C1-C4
C-C-STRUCTURE FOR COMPONENT 204
STRU 204=C1-C2,C2-C3,C3-C4,C2-C5,C2-C6,C3-C7
FILE NEW=233T,CHEM,DATA
END
```

Chemtrn11.in - Using ChemTran to Output Component Physical Properties

Input file showing use of PRInt PROperties command to print pure component properties in the DESIGN II/ChemTran database.

```
AB123.
*PRINT PROPERTIES OF BENZENE
COM = 40
PRInt PROperties
END
```

Chemtrn12.in - Using the Modified Peng-Robinson Equation of State

Input file to fit data using Modified Peng-Robinson equation of state. This new method is applicable to non-ideal chemical systems at high pressure.

```
AB123.
*WATER/CO2 REGRESSION WITH MODIFIED PENG ROBINSON
*METHOD
C-
C-     WATER CO2
COMPONENTS = 62, 49
C-
C- SPECIFY MODIFIED PENG ROBINSON EQUATION OF STATE
C- WITH MARGULES
C- MIX RULE
MODPENK (MAR)
C- SPECIFY REGRESSION OF KAPPA PARAMETERS FROM
C- LIBRARY VAPOR PRESSURE DATA
KMODPEN (VAP) 62
KMODPEN (VAP) 49
C-
C- REGRESS VLE DATA
C-
P-X-Y(BAR,K) 49,62=383.15, 99.999,.0140,.9560
        199.997,.0210,.9580
        299.995,.0240,.9480
        399.994,.0260,.9320
        499.992,.0280,.9140
        599.991,.0300,.8930
        699.990,.0315,.8720
        799.988,.0330,.8540
        899.986,.0345,.8400
FILE NEW = CO2,WATR,DATA
END
```

Chemtrn13.in - Entering Weak Electrolyte Data

Example to include weak electrolyte data (ionic formula & charge, ionic reaction information, Henry's constant information and model binary interaction parameters for ion-ion, molecule-ion, molecule-molecule pairs) using EDWARD's model.

```
AB123.
* DEFINE USER SPECIFIED REACTIONS
C- IDENTIFY LIBRARY AND ION COMPONENTS.
C-WATER, AMMONIA, HYDROGEN SULFIDE.
COMPONENT = 62, 63,50
C-      H+, OH-, HS-, NH4+,S=
ION COMPONENT = 9401, 9402, 9632, 9701,9631

C-CHOOSE THERMODYNAMIC METHOD.
EDWARDK

C- ENTER USER DEFINED LIBRARY REACTIONS.
LIBRARY REACTION = 101,104,105,109

C- DEFINE REACTION STOICHIOMETRY
C- REACTION 1  H20 <=> H+  +OH-
REACTION STOI 101=(1*62=1*9401+1*9402)
REACTION STOI 109=(1*63+1*62=1*9701+1*9402)
REACTION STOI 104=(1*50 = 1*9401 + 1*9632)
REACTION STOI 105=(1*9632 = 1*9401 + 1*9631)
```

```
C- ENTER EQUILIBRIUM CONSTANTS C1,C2,C3, and C4
EQU CON (R) 101=500,800,154.144,-24202.62,0.0,-22.4773,
EQU CON (R) 109=500,800,106.8259,-10645.3476,-0.006115561,-15.06399
EQU CON (R) 104=500,800,507.6058,-32462.496,0.0511013,-78.07186
EQU CON (R) 105=500,800,-234.4787,-730.8063,-0.03006157,33.88898

C- ENTER EDWARDS INTERACTION PARAMETERS FOR MOLECULE-MOLECULE PAIRS
EDA 63,63=-0.0260,-.0260,0.0,0.0,22.122,22.122
EDA 63,50=-0.1186,-0.1886,0.0,0.0,66.465,66.465
EDA 50,50=-0.2106,-0.2106,0.0,0.0,110.808,110.808

C- ENTER EDWARDS INTERACTION PARAMETERS FOR ION-ION PAIRS.
C- NOT NECESSARY FOR IONS OF SAME CHARGE.
EDA 9401,9402=0.208,0.208
EDA 9401,9632=0.194,0.194
EDA 9401,9631=.127,.127

EDA 9701,9402=0.060,0.060
EDA 9701,9632=0.046,0.046
EDA 9701,9631=-0.021,-0.021

EDB 9401,9402=0.65448,0.65448
EDB 9401,9632=0.61164,0.61164
EDB 9401,9631=0.40662,0.40662

EDB 9701,9402=0.20160,0.20160
EDB 9701,9632=0.15876,0.15876
EDB 9701,9631=-0.04626, -0.04626

C-ENTER EDWARDS INTERACTION PARAMETERS FOR ION-MOLECULE PAIRS.
EDA 63,9632=-0.06423,-0.06423,0.0001010,0.000101,0.0,0.0
EDA 63,9401=0.015,0.015,0.0,0.0,0.0,0.0
EDA 63,9402=-0.03801,-0.03801,0.0001169,0.0001169,0.0,0.0
EDA 50,9701=0.07932,0.07932,0.0000066,0.0000066,0.0,0.0
EDA 50,9401=0.017,0.017,0.0,0.0,0.0,0.0
EDA 50,9402=-0.05290,-0.05290,0.0001467,0.0001467,0.0,0.0
EDA 63,9631 = 0.032, 0.032,0.,0.,0.,0.

C- IDENTIFY CHEMICAL FILE NAME.
FILE NEW = AAAA,BBBB,CCCC
END
```

Chemtran_Pure CompDialog.psd – Entering Properties to Define a Custom Component

The following input file illustrates how ChemTran can be used to define properties for a component not included in the DESIGN II or ChemTran component database. There are 7 commands required to define a custom component. If the following commands are entered, all other thermodynamic and physical properties can be calculated. Although the following input file defines the minimum set of commands necessary to define a custom component, you may enter data for any of the properties listed in ***ChemTran: Pure Component Command Details***.

```
AB123.
*CHEMTRAN DEMONSTRATION
COM=200
C-NON-DATABASE COMPONENT NAMES
C-
NAM200=2.3.4-TRI-C1-C5
C-
C-PURE COMPONENT DATA FOR COMPONENT 200
C-ENTER MOLECULAR WEIGHT, NORMAL BOILING POINT,
C-CRITICAL TEMPERATURE, CRITICAL PRESSURE,
C-VAPOR PRESSURE DATA, CRITICAL VOLUME,
C-AND IDEAL GAS HEAT CAPACITY DATA
MOL 200=114.22
TB(C) 200=113.47
TC(F) 200=559.47
PC(ATM) 200=26.94
P-T (C,MMHG) 200=36.57,300.0,47.71,36.57,760,113.47,
       20473.9,293.04
VC (FT3/LBMOL) 200=7.38
CP-T (R,DAT,T**3) 200=500,2800,45.35,540,57.52,720,68.22,900,
       77.56,1080,85.65,1260,92.6,1440,98.52,1620,
```

```
        103.51,1800,107.7,1980,111.1,2160,114.07,
        2340,116.47,2520,118.5,2700
FILE NEW = 234T,CHEM,DATA
        END
```

Chemtran_Pure Comp & Mixture Dialog.psd – Entering Properties to Define a Custom Component and/ or Mixture

A sample flowsheet showing how to add new components and mixture data via their dialogs to a flowsheet.

Ctrefrig.in - Sample To Enter New Refrigerants Data

Sample for pure component data addition of Refrigerants R134A, R123 and SF6.

Democt.in - Demonstration To Add New Components

Demonstration to add new components not in DESIGN II database by using structure and point properties commands.

EgWater.in - Ethylene-Water VLE Data Using Peng-Robinson

VLE data regression using Peng-Robinson for ethylene-water system.

Eth2oct.in - Ethanol-Water VLE Data Using NRTL

VLE data regression using Non-Random Two Liquid theory (NRTL-RENON) for ethanol-water system.

Llect.in - Data Regression For LLE Column Using UNIQUAC

LLE data regression using UNIQUAC.

Meohct.in - Methanol-Water VLE Data

VLE data regression using Non-Random Two Liquid theory (NRTL-RENON) for methanol-water system.

Solidsct.in - Using Solids As Chem-200 Level Components

File to include solids data as Chem-200 level component to perform simulations of ash, char, and carbon.

Sulfract.in - Add Sulfuric Acid To DESIGN II Database

Sample for pure component data addition of Sulfuric Acid (H2SO4) to DESIGN II database.

Sulfrct.in - Add Sulfur To Database

Sample for pure component data of Sulfur (S) to DESIGN II database.

Unifct.in - Estimating Bips From Infinite Dilution Activity Coefficients

Input file to back out binary interaction parameters using infinite dilution activity coefficients of the binary using UNIFAC (estimation method).

Uniqct.in - VLE Data Regression Using UNIQUAC

VLE data regression using UNIQUAC for ethanol-water system.

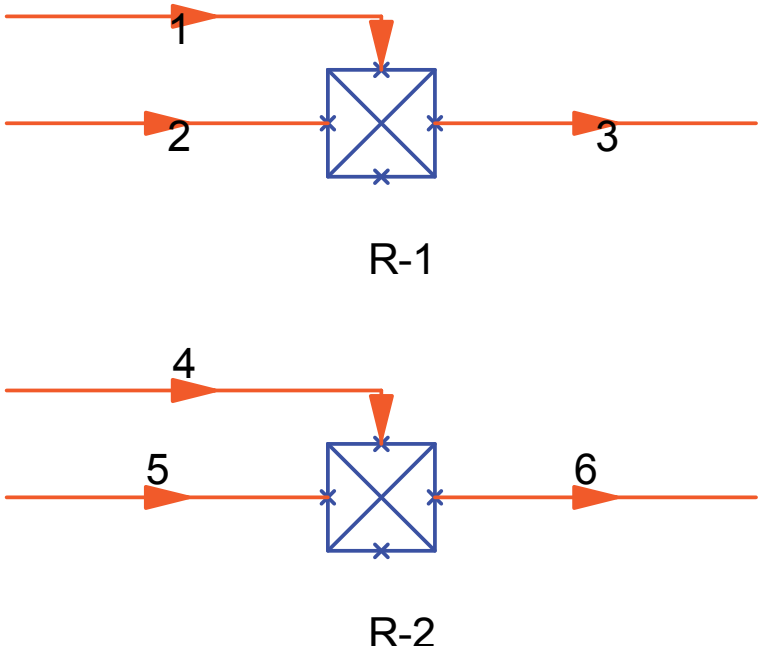
Ureact.in - Add UREA To Database

Sample for pure component of UREA [(NH2)2CO] to DESIGN II database.

Chapter 10: Coal Samples - c:\designii\samples\coal\carbon_equ.psd

C:\designii\samples\coal\carbon_equ.psd – Coal Gasification

Sample file illustrates coal gasification using equilibrium reactor.



Chapter 11: Add Module Samples - c:\designii\samples\equipmnt\add

Addmod.psd - Example Coding

This is a user written equipment module, or ADD block, of a simple air cooled exchanger.


```
AB123.
*ADD MODULE EXAMPLE CODING

ADD 1 = AIRCOOLER, 1,-2,
USE BLOCK 3
NEQP=10
EQP=8
F-          COMMON /AIRCLR/ TOUT,TINAIR,UX
F-          TOUT=150.
F-          TINAIR=100.
F-          UX  =4.2

C- GENERAL SECTION
GENERAL
  COMPONENTS = 100,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 1
  CON TOL = 1.E-003
  FLO(LB/HR)1 = 27300.,
  TP (F,PSIA) 1 = 250.,50.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
 AMB 100=350,
 API 100=37,
 AMW 100=175,
 NAM 100=PETFRAC

FORTRAN
   BLOCK 3=1,AIRCLR

C-*********************
C-  PLEASE REMOVE THE 'C-' IN THE FOLLOWING LINE TO STORE THE
C-  DATA IN THE FORTRAN FILE CALLED "AIRCOOL2.DAT"
C-*********************
C-   NEW FORTRAN FILE = AIRC,OOL2,DATA

START LIBRARY
      SUBROUTINE AIRCLR ( NIN,    NOUT,   NCP,    NEQP,   NDSP,
     &          SIVPFR, SITEMP, SIPRES, SIENTH, SIMOLE, SICOMP, SIKV,
     &          SOVPFR, SOTEMP, SOPRES, SOENTH, SOMOLE, SOCOMP,
     &          SOKV,   EQPAR,  DESPAR, AMW,    ATB,    IDCOMP  ,
     *          ISTOP,  KTRACE, NCP)
      DIMENSION SIVPFR(NIN), SITEMP(NIN), SIPRES(NIN), SIENTH(NIN),
     &          SIMOLE(NIN), SICOMP(NCP,NIN), SIKV(NCP,NIN)
      DIMENSION SOVPFR(NOUT), SOTEMP(NOUT), SOPRES(NOUT), SOENTH(NOUT),
     &          SOMOLE(NOUT), SOCOMP(NCP,NOUT), SOKV(NCP,NOUT)
      DIMENSION EQPAR(NEQP), DESPAR(NDSP), AMW(NCP), ATB(NCP),
     &          IDCOMP(NCP)
      DIMENSION DECANT(2)
      COMMON /AIRCLR/ TOUT,TINAIR,UX
C--
C--   THIS IS A SIMPLE AIR COOLED EXCHANGER ADAPTED FROM
C--   THE GPSA'S ENGINEERING DATA BOOK.
C--
C--   ONE INLET STREAM AND 1, 2, OR 3 OUTLET STREAMS ARE ALLOWED.
C--   IF 2 OR 3 OUTLET STREAMS ARE CODED, THE PHASES ARE SEPARATED
C--   AS IN THE VALVE MODULE.
C--   INPUT PARAMETERS
```

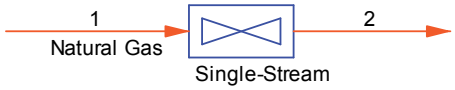
```
C--   EQPAR(1) =NE      ...EQUIPMENT NUMBER IN FLOWSHEET
C--   EQPAR(2) =DUMMY   ...SET TO ZERO, USED LATER FOR ISOTHERMAL FLASH
C--   EQPAR(3) =TOUT    ...TEM OUT SPEC ON PROCESS STREAM (F)
C--   EQPAR(4) =TINAIR  ...TEMPERATURE OF AIR (F), DEFAULT IS 90 F
C--   EQPAR(5) =UX      ...EXTENDED SURFACE HEAT TRANSFER COEFFICIENT
C--                     ...DEFAULT IS 4.0 BTU/HR*FT2*F
C--   OUTPUT PARAMETERS
C--   -----------------
C--   EQPAR(6) =Q       ...DUTY (BTU/HR)
C--   EQPAR(7) =DELAIR  ...AIR TEMPERATURE CHANGE (F)
C--   EQPAR(8) =TOUTA   ...TEM OUT OF AIR (F)
C--   EQPAR(9) =WTAIR   ...WEIGHT OF AIR FLOW (LB/HR)
C--   EQPAR(10)=ACFM    ...ACTUAL FLOW OF AIR (FT3/MIN)
C--
C--      COLLECT INPUT DATA AND SET DEFAULTS
      IF(TOUT.NE.0.)   EQPAR(3)=TOUT
      IF(TINAIR.NE.0.) EQPAR(4)=TINAIR
      IF(UX.NE.0.)     EQPAR(5)=UX
      TIN   =SITEMP(1) - 459.67
      TOUT  =EQPAR(3)
      HIN   =SIENTH(1)
      TINAIR=EQPAR(4)
      UX    =EQPAR(5)
      IF(TOUT.EQ.0.) WRITE(6,5)
5     FORMAT(//,5X,'WARNING- OUTLET TEMPERATURE NOT SET',/,
     &       5X,'OUTLET TEMPERATURE SET EQUAL TO INLET TEMPERATURE',/)
      IF(TOUT.EQ.0.) TOUT=TIN
      IF(TINAIR.EQ.0.)TINAIR=90.
      IF(UX.EQ.0.)UX=4.
C--
C--      ISOTHERMAL FLASH TO CALCULATE OUTLET STREAMS
      EQPAR(2) = 1.
      SITEMP(1)=TOUT + 459.67
      CALL FLASH1     ( NIN,    NOUT,   NCP,    NEQP,   NDSP,
     &          SIVPFR, SITEMP, SIPRES, SIENTH, SIMOLE, SICOMP, SIKV,
     &          SOVPFR, SOTEMP, SOPRES, SOENTH, SOMOLE, SOCOMP,
     &          SOKV,   DECANT, EQPAR,  DESPAR)
C--
C--      CALCULATE DUTY, AIR TEMP OUT, AND AIR FLOW
      HOUT   =0.
      DO 10 I=1,NOUT
10    HOUT   =HOUT + SOENTH(I)
      Q      =HOUT - HIN
      F1     =(UX-1.)/10.
      F2     =((TIN+TOUT)/2)-TINAIR
      DELAIR =F1*F2
      TOUTA  =TINAIR+DELAIR
      WTAIR  =ABS(Q)/(0.24*DELAIR)
      ACFM   =0.222*WTAIR/0.94
C--
C--      PUT RESULTS IN EQPAR VECTOR
      EQPAR(3) =TOUT
      EQPAR(4) =TINAIR
      EQPAR(5) =UX
      EQPAR(6) =Q
      EQPAR(7) =DELAIR
      EQPAR(8) =TOUTA
      EQPAR(9) =WTAIR
      EQPAR(10)=ACFM
C--
C--      OPTIONALLY, WRITE RESULTS TO OUTPUT FILE IN USER DEFINED FORMAT
      WRITE(6,100)
      WRITE(6,101)TOUT
      WRITE(6,102)TINAIR
      WRITE(6,103)UX
      WRITE(6,104)Q
      WRITE(6,105)DELAIR
      WRITE(6,106)TOUTA
      WRITE(6,107)WTAIR
      WRITE(6,108)ACFM
100   FORMAT(//,5X,'AIR COOLED HEAT EXCHANGER DATA',/,
     &       5X,'================================================',/)
101   FORMAT(5X,'TEM OUT SPEC ON PROCESS STREAM (F)  =',F15.2)
102   FORMAT(5X,'TEMPERATURE OF AIR (F)              =',F15.2)
103   FORMAT(5X,'EXTENDED SURFACE HEAT TRANSFER COEF.=',F15.2)
104   FORMAT(5X,'DUTY (BTU/HR)                       =',F15.2)
105   FORMAT(5X,'AIR TEMPERATURE CHANGE (F)          =',F15.2)
```

```
106   FORMAT(5X,'TEM OUT OF AIR (F)                  =',F15.2)
107   FORMAT(5X,'WEIGHT OF AIR FLOW (LB/HR)          =',F15.2)
108   FORMAT(5X,'ACTUAL FLOW OF AIR (FT3/MIN)        =',F15.2,/)
      RETURN
      END
STOP LIBRARY
END
```

Chapter 12: Air Cooled Exchanger Samples - c:\designii\samples\equipmnt\aircoolr

Airc1.psd – Single-Stream Natural Gas Cooler (rating off)

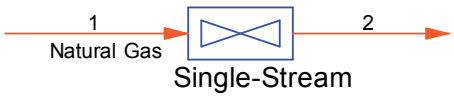
This is a simple case of a natural gas being cooled by air contact using air-cooled exchanger. The cold side air flow is used to cool the incoming hot side feed natural gas. In addition to heat exchange calculations, the air-cooled exchanger predicts the required air-flow for the specified air and process specifications. No rating is performed.




```
AB123.
*Courtesy Hudson Air Products

C- EQUIPMENT MODULES

AIRC 1 = Single-Stream , 1,-2,
  TEM OUT HOT(F) = 140.
  TEM OUT AIR(F) = 140.8
  TEM AIR IN(F) = 104.
  Q-T
C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,99,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 1.E-002,6.E-003,0.9,3.3E-002,2.1E-002,5.E-003,2.5E-002,0.,
  TOT FLO (LB/HR) 1 = 926715.
  TP (F,PSIA) 1 = 167.,881.85
  NAM STR 1 = Natural Gas
  NAM STR 2 = Strm 2
END
```

Airc2.psd – Single-Stream Natural Gas Cooler (rating on)

This is a simple case of a natural gas being cooled by air contact using air-cooled exchanger. The cold side air flow is used to cool the incoming hot side inlet feed natural gas. In addition to heat exchange calculations, a rating calculation is performed for the air-cooled exchanger based on the geometry specifications (fin, fan, and tube dimensions).and predicts the required air-flow rate for the specified fan power. Air-Cooled Exchanger rating results are also provided.



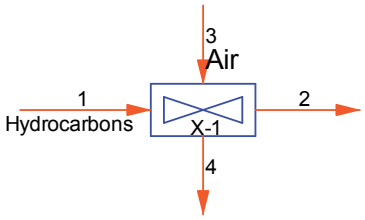

```
AB123.
*Courtesy-Hudson Air Products...Natural Gas Cooler (rating on)

C- EQUIPMENT MODULES

AIRC 1 = Single-Stream, 1,-2,
  FAN POW(HP) = 6.9
  DEL PRE HOT(PSIA) = 4.9782
  TEM AIR IN(F) = 104.
  Q-T
  RAT
  NUM PAR = 3
```

```
  NUM BUN BAY = 1
  UFA = 1.
  TYP = 1
  DRA TYP = 1
  U-M = 4
  FIN MAT = 4
  FIN HEI(IN) = 0.625
  FIN THI(IN) = 1.6E-002
  FIN SHA = EMB
  NUM FIN PER(1/IN) = 10.
  FAN DIA(FT) = 9.
  HUB DIA(FT) = 0.
  FAN SPE(RPM) = 345.
  FAN EFF(FRA) = 0.7
  FAN RED EFF(FRA) = 0.92
  NUM FAN = 2
  FOU AIR(1/BTU/FT2/HR/F) = 1.E-003
  TUB MAT = 1
  TUB OUT DIA(IN) = 1.
  TUB BWG = 16
  TUB WAL(IN) = 9.5E-002
  TUB LEN(FT) = 30.
  TUB PIT(IN) = 2.5
  TUB PAS = 1.
  TUB LAY = 60
  FOU TUB(1/BTU/FT2/HR/F) = 1.E-003
  NUM TUB = 264
  NUM ROW = 6

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,99,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 =1.E-002,6.E-003,0.9,3.3E-002,2.1E-002,
  5.E-003,2.5E-002,0.,
  TOT FLO (LB/HR) 1 = 926715.
  TP (F,PSIA) 1 = 167.,881.85
  NAM STR 1 = Natural Gas
  NAM STR 2 = Strm 2
END

Courtesy-Hudson Air Products...Natural Gas Cooler (rating o           PAGE    6
AIR        1 = Sing

*** PROCESS RATING - AIR COOLED EXCHANGER    1 = Single-Stream
=====================================================================
 NUMBER OF: PARALLEL BUNDLES  3   TUBES/BUNDLE  264         FANS/BAY  2
            BUNDLES/BAY  1
 ---------------------------------------------------------------------------
                               I     TUBE SIDE       I      AIR SIDE
 ---------------------------------------------------------------------------
 FEED STREAMS                  I    1                I   UTILITY STREAM
 PRODUCT STREAMS               I    2                I   UTILITY STREAM
 TEMPERATURE (IN/OUT)  DEG F   I  167.00 /  138.22   I  104.00 /  145.52
 PRESSURE    (IN/OUT)  PSIA    I  881.85 /  876.87   I   14.71 /   14.70
 VELOCITY    (IN/OUT)  FT/SEC  I   38.17 /   36.11   I    7.93 /    8.53
 HEAT EXCHANGED        BTU/HR  I    -1.64741E+07     I     1.64766E+07
 FOULING FACTOR   HR-FT2-F/BTU I     1.00000E-03     I     1.00000E-03
 HEAT TRANS. COEF BTU/HR/FT2/F I     16.714          I     7.547
 PRES DROP (SPEC/CALC) PSI     I     4.978 / 3.720   I     0.000 / 0.013
 TOTAL FLUID FLOWRATE  LB/HR   I     9.26715E+05     I     1.66009E+06
       VAPOR         (IN/OUT)  I9.2672E+05/9.2672E+05I     1.66009E+06
 PROPERTIES                    I                     I
 ----------                    I                     I
 SP. GRAVITY @ 60 F   DEG API  I       321.102       I
 MOLECULAR WEIGHT              I                     I
       VAPOR         (IN/OUT)  I  17.041 / 17.041    I     28.964
 VISCOSITY                CP   I                     I
       VAPOR         (IN/OUT)  I  0.0140 / 0.0136    I   0.0196 / 0.0196
 DENSITY               LB/FT3   I                     I
       VAPOR         (IN/OUT)  I   2.380 /  2.515    I   0.0704 / 0.0655
```

```
 SPECIFIC HEAT       BTU/LB/F  I                      I
       VAPOR         (IN/OUT)  I  0.6229 / 0.6219    I   0.2390 / 0.2390
 THE. CONDUCTIVITY BTU/FT/HR/F I                      I
       VAPOR         (IN/OUT)  I  0.0272 / 0.0260    I   0.0167 / 0.0167
 ------------------------------------------------------------------------------
 HEAT EXCHANGED       BTU/HR   =    0.16474E+08
 OVERALL U COEF.  BTU/FT/HR/F  =    4.4217
 CORRECTED LMTD       DEG F    =    27.849
 LMTD CORRECTION FACTOR        =    1.0000
       SURFACE AREA USED      FT2    =    1.33815E+05
 DELTA T CALCULATION METHOD    =    ARITHMETIC
 FAN DRAFT TYPE                =    FORCED
 FAN SPEED              RPM    =    345.0
 FAN DIAMETER           FT     =     9.0
 ACTUAL AIR RATE/FAN   FT3/MIN =    65503.
 FAN POWER/FAN          HP     =    6.898
 FAN EFFICIENCY                =    0.7000
 REDUCER EFFICIENCY            =    0.9200
 TUBE THERMAL COND BTU/FT/HR/F =    26.00
 ------------------------------------------------------------------------------
       EXCHANGER DETAILS:
 ------------------------------------------------------------------------------
 TUBE SIDE                          AIR SIDE
 =========                          ========
 OUTSIDE DIAMETER     IN =   1.000  NUMBER FINS PER IN =   10.0
 INSIDE DIAMETER      IN =   0.810  FIN HEIGHT      IN =   0.625
 TUBE PITCH (TRIANG.) IN =   2.50   FIN THICKNESS   IN =   0.016
 TUBE LENGTH          FT =   30.0   FIN SHAPE          =   EMBEDDED
 NUMBER TUBE PASSES      =   1.
 NUMBER OF ROWS          =   6.
 MATERIAL: TUBE = CARBON STEEL        FIN               = ALUMINUM
 ------------------------------------------------------------------------------
```

Airc3.psd – Two-Stream Hydrocarbon Gas Condenser (rating off)

In this sample simulation, hydrocarbons are condensed by air using air-cooled exchanger. The cold side air flow is used to condense as much as the incoming hot side process hydrocarbons feed. In addition to heat exchange calculations, the specified duty for air-cooled exchanger, the outlet process and air-conditions are also determined. No rating is performed.

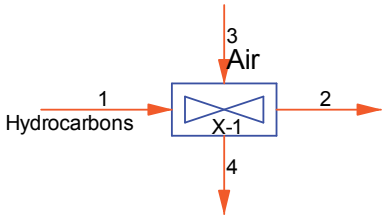



```
AB123.
*Courtesy-Hudson Air Products Hydrocarbon Condenser

C- EQUIPMENT MODULES

AIRC 1 = X-1, 1,3,-2,-4,
  DUT(BTU/HR) = 35700000.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 6,7,8,9,10,11,37,99,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 2.E-002,0.1,0.16,0.2,0.32,2*0.1,0.,
  TOT FLO (LB/HR) 1 = 248355.
  TP (F,PSIG) 1 = 179.6,14.2
  NAM STR 1 = Hydrocarbons
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 7*0.,1.,
  TOT FLO (SCF/MIN) 3 = 1288980.
  TP (F,PSIA) 3 = 109.4,14.7
```

```
  NAM STR 3 = Air
  NAM STR 4 = Strm 4
END
```

Airc4.psd – Two-Stream Hydrocarbon Gas Condenser (rating on)

In this sample simulation, hydrocarbons are condensed by air using air-cooled exchanger. The cold side air flow is used to condense as much as the incoming hot side process hydrocarbons feed. In addition to heat exchange calculations, rating is performed for the air-cooled exchanger based ONLY on the geometry specifications (fin, fan, and tube dimensions). Air-Cooled Exchanger rating results are also provided.




```
AB123.
*Courtesy-Hudson Air Products-Hydrocarbons Condenser (rating on)

C- EQUIPMENT MODULES

AIRC 1 = X-1, 1,3,-2,-4,
  DEL PRE HOT(PSIA) = 4.1
  Q-T
  RAT
  NUM PAR = 4
  NUM BUN BAY = 1
  UFA = 1.
  TYP = 2
  DRA TYP = 1
  U-M = 4
  FIN MAT = 4
  FIN HEI(IN) = 0.625
  FIN THI(IN) = 0.164
  FIN SHA = EXT
  NUM FIN PER(1/IN) = 10.
  FAN DIA(FT) = 13.
  HUB DIA(FT) = 0.
  FAN SPE(RPM) = 242.
  FAN EFF(FRA) = 0.7
  FAN RED EFF(FRA) = 0.92
  NUM FAN = 2
  FOU AIR(1/BTU/FT2/HR/F) = 1.95E-003
  TUB MAT = 1
  TUB OUT DIA(IN) = 1.
  TUB WAL(IN) = 0.109
  TUB LEN(FT) = 30.
  TUB PIT(IN) = 2.5
  TUB PAS = 2.
  TUB LAY = 60
  FOU TUB(1/BTU/FT2/HR/F) = 1.46E-003
  NUM TUB = 375
  NUM ROW = 5

C- GENERAL SECTION
GENERAL
  COMPONENTS = 6,7,8,9,10,11,37,99,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 2.E-002,0.1,0.16,0.2,0.32,2*0.1,0.,
  TOT FLO (LB/HR) 1 = 248355.
  TP (F,PSIG) 1 = 179.6,14.2
  NAM STR 1 = Hydrocarbons
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 7*0.,1.,
```

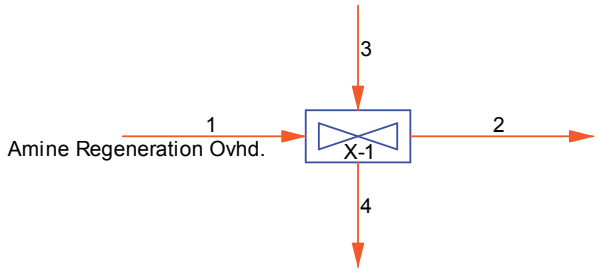
```
  TOT FLO (SCF/MIN) 3 = 1288980.
  TP (F,PSIA) 3 = 109.4,14.7
  NAM STR 3 = Air
  NAM STR 4 = Strm 4
END

Courtesy-Hudson Air Products-Hydrocarbons Condenser (rating           PAGE    7
AIR        1 = X-1

*** PROCESS RATING - AIR COOLED EXCHANGER    1 = X-1
==================================================
 NUMBER OF: PARALLEL BUNDLES  4   TUBES/BUNDLE  375         FANS/BAY  2
            BUNDLES/BAY  1
 ------------------------------------------------------------------------
                                I     TUBE SIDE       I      AIR SIDE
 ------------------------------------------------------------------------
 FEED STREAMS                   I    1                I    3
 PRODUCT STREAMS                I    2                I    4
 TEMPERATURE (IN/OUT)  DEG F    I  179.60 /  134.85   I  109.40 /  132.61
 PRESSURE    (IN/OUT)  PSIA     I   28.90 /   24.80   I   14.72 /   14.70
 VELOCITY    (IN/OUT)  FT/SEC   I   78.34 /   19.50   I   12.52 /   13.06
 HEAT EXCHANGED        BTU/HR   I    -3.28351E+07     I     3.28036E+07
 FOULING FACTOR   HR-FT2-F/BTU I     1.46000E-03     I     1.95000E-03
 HEAT TRANS. COEF BTU/HR/FT2/F I      9.892          I     8.838
 PRES DROP (SPEC/CALC) PSI      I     4.100 / 3.870   I     0.000 / 0.024
 TOTAL FLUID FLOWRATE  LB/HR    I     2.48355E+05     I     5.90373E+06
       VAPOR          (IN/OUT)  I2.4836E+05/5.5076E+04I     5.90373E+06
       LIQUID         (IN/OUT)  I0.0000E+00/1.9328E+05I     0.0000
 PROPERTIES                     I                     I
 ----------                     I                     I
 SP. GRAVITY @ 60 F   DEG API   I        86.826       I
 MOLECULAR WEIGHT               I                     I
       VAPOR          (IN/OUT)  I  79.240 / 74.306    I     28.964
       LIQUID         (IN/OUT)  I   0.000 / 80.646    I
 VISCOSITY                 CP   I                     I
       VAPOR          (IN/OUT)  I  0.0077 / 0.0075    I   0.0198 / 0.0204
       LIQUID         (IN/OUT)  I  0.000 /  0.194     I
 DENSITY                LB/FT3  I                     I
       VAPOR          (IN/OUT)  I   0.352 /  0.303    I   0.0698 / 0.0670
       LIQUID         (IN/OUT)  I   0.000 / 38.392    I
 SPECIFIC HEAT         BTU/LB/F  I                     I
       VAPOR          (IN/OUT)  I  0.4578 / 0.4379    I   0.2394 / 0.2393
       LIQUID         (IN/OUT)  I  0.0000 / 0.5581    I
 THE. CONDUCTIVITY BTU/FT/HR/F I                      I
       VAPOR          (IN/OUT)  I  0.0115 / 0.0107    I   0.0159 / 0.0164
      LIQUID         (IN/OUT)  I  0.0000 / 0.0583    I
 ------------------------------------------------------------------------
 HEAT EXCHANGED       BTU/HR   =    0.32835E+08
 OVERALL U COEF.  BTU/FT/HR/F  =    3.8040
 CORRECTED LMTD       DEG F    =    34.040
 LMTD CORRECTION FACTOR        =    0.9690
 SURFACE AREA USED      FT2    =    2.53438E+05
 DELTA T CALCULATION METHOD    =    LOG MEAN
 FAN DRAFT TYPE                =    FORCED
 FAN SPEED              RPM    =    242.0
 FAN DIAMETER           FT     =    13.0
 ACTUAL AIR RATE/FAN   FT3/MIN =   1.76246E+05
 FAN POWER/FAN          HP     =    33.06
 FAN EFFICIENCY                =    0.7000
 REDUCER EFFICIENCY            =    0.9200
 TUBE THERMAL COND BTU/FT/HR/F =    26.00
 ------------------------------------------------------------------------
       EXCHANGER DETAILS:
 ------------------------------------------------------------------------
 TUBE SIDE                      AIR SIDE
 =========                      ========
 OUTSIDE DIAMETER     IN =   1.000   NUMBER FINS PER IN =   10.0
 INSIDE DIAMETER      IN =   0.782   FIN HEIGHT      IN =   0.625
 TUBE PITCH (TRIANG.) IN =   2.50    FIN THICKNESS   IN =   0.016
 TUBE LENGTH          FT =   30.0    FIN SHAPE          =   EXTRUDED
 NUMBER TUBE PASSES      =   2.
 NUMBER OF ROWS          =   5.
 MATERIAL: TUBE = CARBON STEEL        FIN               = ALUMINUM
 ------------------------------------------------------------------------

       TABLE OF HEAT CONTENT-TEMPERATURE DIAGRAM
```

NO	TEMPERATURE DEG F HOT	COLD	HEAT FLOW(Q) BTU/HR	DELTA T DEG F	UA=DELTA Q/LMTD BTU/HR /F	
1	179.60	132.61	0.00000E+00	46.99	0.00000E+00	
2	177.55	132.45	-2.30714E+05	45.10	5011.2	DH
3	175.74	131.39	-1.72651E+06	44.35	33445.	
4	173.64	130.17	-3.45302E+06	43.48	39319.	
5	171.49	128.95	-5.17953E+06	42.55	40143.	
6	169.33	127.72	-6.90604E+06	41.61	41033.	
7	167.16	126.50	-8.63255E+06	40.65	41977.	
8	164.98	125.28	-1.03591E+07	39.70	42977.	
9	162.77	124.06	-1.20856E+07	38.71	44041.	
10	160.56	122.84	-1.38121E+07	37.72	45179.	
11	158.32	121.61	-1.55386E+07	36.70	46399.	
12	156.06	120.39	-1.72651E+07	35.67	47716.	
13	153.78	119.17	-1.89916E+07	34.61	49137.	
14	151.48	117.95	-2.07181E+07	33.53	50679.	
15	149.16	116.73	-2.24446E+07	32.43	52353.	
16	146.82	115.51	-2.41711E+07	31.31	54177.	
17	144.46	114.29	-2.58976E+07	30.18	56167.	
18	142.08	113.06	-2.76242E+07	29.02	58341.	
19	139.70	111.84	-2.93507E+07	27.86	60722.	
20	137.30	110.62	-3.10772E+07	26.68	63330.	
21	134.85	109.40	-3.28037E+07	25.45	66252.	MN
				TOTAL UA =	9.38399E+05	

```
 WHERE
  DH IS DEW POINT OF HOT STREAMS
  MN IS MIN. TEMP.   APPROACH
```

Airc5.psd – Two-Stream Amine Overheads Cooler (rating off)

In this simulation, mixed amine overheads are cooled by air using air-cooled exchanger. The cold side air flow is used to partially condense the incoming hot side process mixed amine feed. In addition to heat exchange calculations, the specified air flowrate determines the process (user specified) outlet and air outlet temperatures along with process outlet phase condition. No rating is performed.



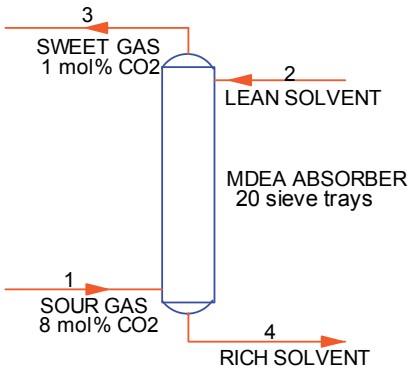

```
AB123.
*Courtesy Hudson Air Products-Mixed Amine Overheads Cooler

C- EQUIPMENT MODULES

AIRC 1 = X-1, 1,3,-2,-4,
  TEM OUT HOT(F) = 129.2

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,49,50,2,3,4,5,6,4051,7051,47,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 0.254,1.E-005,0.73,1.E-002,9.E-004,5.E-004,2.9E-004,2*3.E-005,
  3*0.,
  TOT FLO (LB/HR) 1 = 31892.
```

```
  TP (F,PSIG) 1 = 228.2,16.
  NAM STR 1 = Amine Regeneration Ovhd.
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 0.,0.79,9*0.,0.21,
  TOT FLO (SCF/MIN) 3 = 281055.
  TP (F,PSIA) 3 = 104.,14.7
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
END
```

Chapter 13: Amine Column Samples - c:\designii\samples\equipmnt\aminecol

Mxabs.psd - MEA/MDEA Absorber

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.



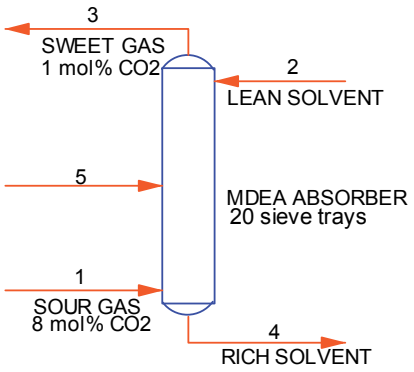
Assumptions
Use 39 weight % solution of MDEA.Amine (MEA 0.1 wt%).
Objectives
1 mol % of CO2 in the sweet gas, with top temperature of 130 F
Specification:
Tower diameter 70 inches. Active area 75%.
Tower internals - sieve trays, with 15% open area, single pass, 2 ft tray spacing.

```
AB123.
*CO2 BULK REMOVAL

C- EQUIPMENT MODULES
AMCL 1 = MDEA ABSORBER, 2,1,-3,-4,
  ABS1
  TRA = 20
  TRA TYP = 20*SIE,
  TRA PAS = 20*1,
  TOW DIA(IN) = 20*70.,
  TRA WEI(IN) = 20*2.5,
  TRA SPA(FT) = 20*2.,
  TRA ARE PER TOW = 20*75.,
  LOC FEE = 1,20
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004
  TRAy PERcent OPEn = 20*15

C- GENERAL SECTION

GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO(FRA)1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
```

```
  TOT FLO(MMSCF/DAY)1 = 60.
  TP (F,PSIA) 1 = 90.,900.
  NAM STR 1 = SOUR GAS
  NAM STR 3 = SWEET GAS
  FLO(FRA)2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO(GAL/MIN)2 = 500.
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 4 = RICH SOLVENT
END
```

Mxabsf.psd - Amine Absorber with Side Feed

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). A side feed stream is also processed in this absorber. Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.

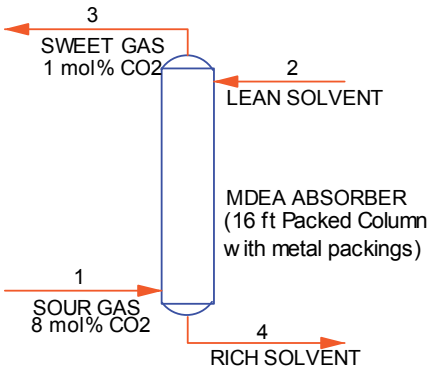



```
AB123.
*CO2 BULK REMOVAL -Side Feed

C- EQUIPMENT MODULES

AMCL 1 = MDEA ABSORBER, 2,5,1,-3,-4,
  ABS2
  TRA = 20
  TRA TYP = 20*SIE,
  TRA PAS = 20*1,
  TOW DIA(IN) = 20*70.,
  TRA WEI(IN) = 20*2.5,
  TRA SPA(FT) = 20*2.,
  TRA ARE PER TOW = 20*75.,
  LOC FEE = 1,10,20
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004
  TRAy PERcent OPEn = 20*15

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO(FRA)1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
  TOT FLO(MMSCF/DAY)1 = 60.
  TP (F,PSIA) 1 = 90.,900.
  NAM STR 1 = SOUR GAS
  NAM STR 3 = SWEET GAS
  FLO(FRA)2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO(GAL/MIN)2 = 500.
```

```
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 4 = RICH SOLVENT
  FLO(LBMOL/HR)5 = 3*1.,0.5,1.E-003,3*0.,
  TP (F,PSIG) 5 = 100.,880.
  NAM STR 5 = Strm 5
END
```

Mxpkabsf.psd - Amine Packed Column Absorber with Side Feed

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). A side feed stream is also processed in this packed column (2 ft ID, 16 ft in depth, 5/8 " metal packings, dry area 109 ft2/ft3, 75 % packing factor) absorber. Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.

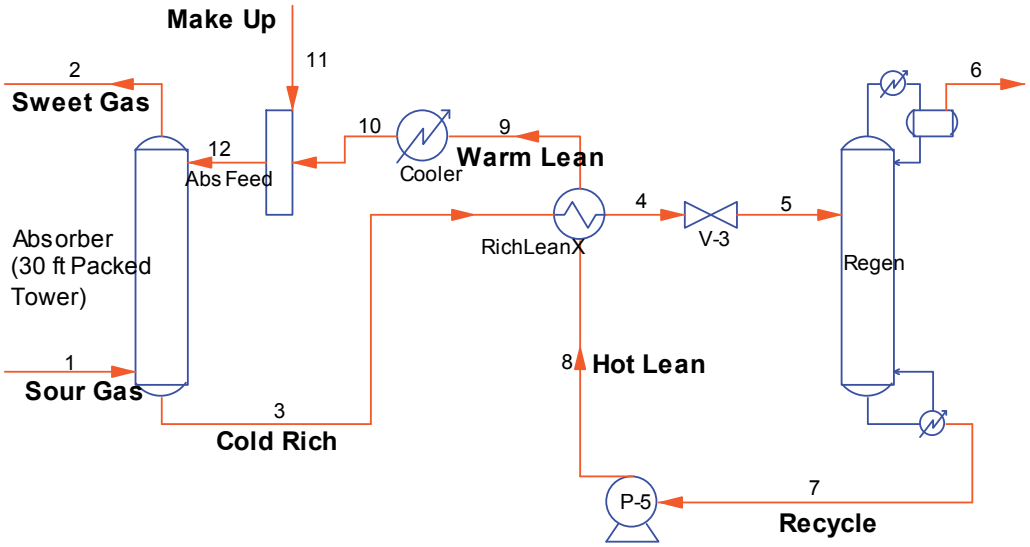



```
AB123.
*CO2 BULK REMOVAL -Packed Column with Side Feed

C- EQUIPMENT MODULES

AMCL 1 = MDEA ABSORBER, 2,5,1,-3,-4,
  ABS2
  PACK DEPTH PER TYPE(FT) = 2*8.,
  PACK MAT = 2*METAL,
  PACK TYP = 2*DUMPED,
  PACK SIZE(IN) = 2*0.625,
  PACK FAC = 2*75.,
  PACK DRY AREA(1/FT) = 2*109.,
  TOW DIA(IN) = 2*24.,
  REDIST DEPTH(FT) = 8.,
  LOC FEE (FT) = 1,10.,16.
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO (FRA) 1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
  TOT FLO (MMSCF/DAY) 1 = 60.
  TP (F,PSIA) 1 = 98.,900.
  NAM STR 1 = SOUR GAS
  FLO (FRA) 2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO (GAL/MIN) 2 = 400.
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 3 = SWEET GAS
```

```
  NAM STR 4 = RICH SOLVENT
  FLO(LBMOL/HR)5 = 3*1.,0.5,1.E-003,3*0.,
  TP (F,PSIA) 5 = 100.,900.
  NAM STR 5 = Strm 5
END
```

Mxpkplnt.psd - Amine Packed Column Absorber with Regenerator

The objective of this simulation is to contact CO2 (7 mol%) and H2S (2 mol %) in the natural gas stream with an aqueous stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .015 mol % of H2S and 0.50 mole % of CO2 per 25 MMSCFD of gas processed.



Use a packed column absorber with 30 ft of packing, 75 % tower flooding, 2/3" metal packings, dry area 105 ft2/ft3, 70 % packing factor and a redistributor at mid point of the two packing sections

Assumptions

Use 10 weight % solution of DEA blended with 39 wt % of MDEA to make about 240 gal/min of aqueous recirculating amine. Make-up water and the amine losses are estimated and made-up automatically by the Mass Balance unit module.

Objectives

To achieve 0.50 mol % of CO2 in the sweet gas, with top temperature of about 115 F

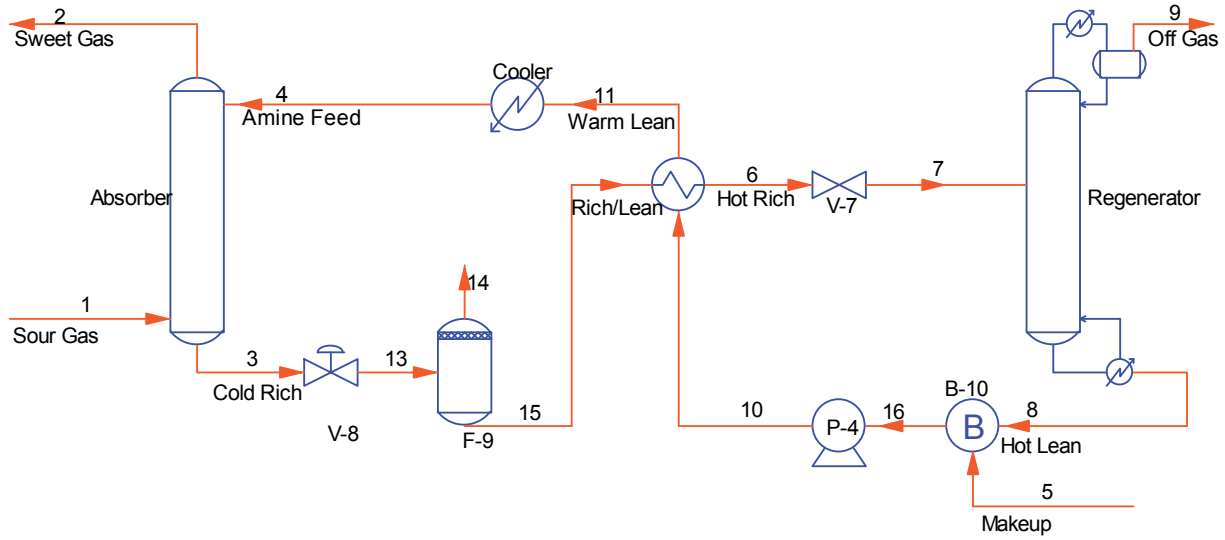
```
AB123.
*Mixed Amine Example MEA/MDEA using Packed Absorber

C- EQUIPMENT MODULES

MIX 1 = M-1, 11,10,-12,

HEA EXC 2 = Cooler, 9,-10,
  U(BTU/HR/FT2/F) = 50.
  SHE = 1
  TEM OUT(F) = 120.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 0.

VAL 3 = V-3, 4,-5,
  PRE OUT(PSIG) = 35.3

HEA EXC 4 = RichLeanX, 8,3,-9,-4,
  U(BTU/HR/FT2/F) = 50.
  TEM APP(F) = 25.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,

PUM 5 = P-5, 7,-8,
  PRE OUT(PSIG) = 1000.
```

```
  EFF = 0.9
  ELE DRI

AMCL 6 = Absorber, 12,1,-2,-3,
  ABS1
  PACK DEPTH PER TYPE(FT) = 2*15.,
  PACK MAT = 2*METAL,
  PACK TYP = 2*DUMPED,
  PACK SIZE(IN) = 2*0.675,
  PACK FAC = 2*70.,
  PACK DRY AREA(1/FT) = 2*105.,
  TOW DIA FLOOD = 2*75.,
  REDIST DEPTH(FT) = 15.,
  LOC FEE (FT) = 1,30.
  PRE TOP(PSIG) = 1000.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 0,4000.
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004

AMCL 7 = Regen, 5,-6,-7,
  REG1
  TRA = 10
  TRA TYP = 10*SIE,
  TRA PAS = 2*2,8*4,
  TOW DIA(FT) = 10*3.,
  TRA WEI(IN) = 10*2.,
  TRA SPA(IN) = 10*30.,
  TRA ARE PER TOW = 10*80.,
  TRA PER OPE = 10*15.,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIG) = 8.
  PRE REB(PSIG) = 15.
  PRO (LBMOL/HR) = 242.07,0
  TEM CON(F) = 110.
  REB DUT(BTU/HR) = 12500000.
  SYS FAC = 0.85
  MAX ITER = 150
  TOL = 1.E-003
F-      POST
F-      F11 = GET (STR(11), FLO COM(1))
F-      F1 = GET (STR(1), FLO COM(1))
F-      F6 = GET (STR(6), FLO COM(1))
F-      F2 = GET (STR(2), FLO COM(1))
F-      FIN = F1 + F11
F-      FOUT = F2 + F6
F-      DELFLO = (FOUT-FIN) * 0.8
F-      FLO11 = F11 + DELFLO
F-      WRITE (1, 111) FIN, F2, F6
F-      WRITE (6, 111) FIN, F2, F6
F-  111 FORMAT (' WATER IN = ', F8.2, ' F2= ', F8.2, ' F6 =', F8.2)
F-      WRITE (1, 121) F11, FLO11
F-      WRITE (6, 121) F11, FLO11
F-  121 FORMAT (' WATER DELTA = ', F8.2, '   NEW WATER MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (1)) = FLO11
F-      FM11 = GET (STR(11), FLO COM(2))
F-      FM6 = GET (STR(6), FLO COM(2))
F-      FM2 = GET (STR(2), FLO COM(2))
F-      FMOUT = FM6 + FM10
F-      DELFM = (FMOUT-FM11) * 0.8
F-      FLOM11 = FM11 + DELFM
F-      WRITE (1, 112) FM11, FM2, FM6
F-      WRITE (6, 112) FM11, FM2, FM6
F-  112 FORMAT (' DEA IN = ', F8.2, ' FM2= ', F8.2, ' FM6 =', F8.2)
F-      WRITE (1, 122) FM11, FLOM11
F-      WRITE (6, 122) FM11, FLOM11
F-  122 FORMAT (' DEA DELTA = ', F8.2, '   NEW DEA MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (2)) = FLOM11
F-      FMA11 = GET (STR(11), FLO COM(3))
F-      FMA6 = GET (STR(6), FLO COM(3))
F-      FMA2 = GET (STR(2), FLO COM(3))
F-      FMAOUT = FMA6 + FMA2
F-      DELFMA = (FMAOUT-FMA11) * 0.8
F-      FLOA11 = FMA11 + DELFMA
```

```
F-      WRITE (1, 113) FMA11, FMA2, FMA6
F-      WRITE (6, 113) FMA11, FMA2, FMA6
F-  113 FORMAT (' MDEA IN = ', F8.2, ' FMA2= ', F8.2, ' FMA6 =', F8.2)
F-      WRITE (1, 123) FMA11, FLOA11
F-      WRITE (6, 123) FMA11, FLOA11
F-  123 FORMAT (' MDEA DELTA = ', F8.2, '   NEW MDEA MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (3)) = FLOA11
F-      CALL INISTR (11)

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 20
  CON TOL = 1.E-003
  REC STR = 7,12,
  SAT FEE WAT = 1,
  FLO(LBMOL/HR)1 = 2.565,2*0.,54.851,191.978,2495.709,
  TP (F,PSIG) 1 = 100.,1000.
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO (FRA) 3 = 0.79984,2.74842E-002,9.69907E-002,2.35E-002,5.138E-002,
  8.0298E-004,
  TOT FLO (LBMOL/HR) 3 = 4156.6623
  TP (F,PSIA) 3 = 179.19,1014.696
  NAM STR 3 = Cold Rich
  NAM STR 4 = Strm 8
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 12
  FLO(LBMOL/HR)7 = 3363.,114.,403.,4.,6.,0.,
  TP (F,PSIG) 7 = 256.9,15.4
  NAM STR 7 = Recycle
  NAM STR 8 = Hot Lean
  NAM STR 9 = Warm Lean
  NAM STR 10 = Strm 5
  FLO (FRA) 11 = 1.,5*0.,
  TOT FLO (LBMOL/HR) 11 = 1.E-002
  TP (F,PSIG) 11 = 120.,1000.
  NAM STR 11 = Make Up
  FLO(LBMOL/HR)12 = 3386.,114.,403.,5.,6.,0.,
  TP (F,PSIA) 12 = 125.,1000.
  NAM STR 12 = Abs Feed
REC SEQ = 5, 6, 4, 2, 1, 3, 7, 5, 4, 2, 1
END
```

MxpInt1.psd - Mixed Amine Plant for CO2 / H2S Removal

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



Assumptions
Use 16 weight % solution of DEA.blended with 28 wt % of MDEA to make about 100 gal/min of aqueous recirculating amine. Make-up water is estimated automatically by the Inline Fortran code already set-up in the regenerator.

Objectives
To achieve about 1.7 mol % of CO2 in the sweet gas, with top temperature of 150 F

Specifications
Feed Gas
Temperature = 90 F
Pressure = 880 psig

Component	Mole Percent (Dry Basis)
Water	0.000
Nitrogen	0.170
CO2	5.600
H2S	0.150
Methane	75.119
Ethane	11.895
Propane	4.799
i-Butane	0.716
n-Butane	1.551

Equipment
Absorber:
Diameter = 4 ft
Number of Trays=10
Tray Parameters = Valve Trays, 2 ft spacing, 2 inch weirs, 63.6 % Active Area
Rich/Lean Exchanger: Temperature Approach = 10 F
Valve: Pressure Out = 25.2 psia
Lean Cooler: Temperature Out = 127 F
Pump: Pressure Out = 880 psig
Mixer: None
Regenerator:
Diameter = 3 ft
Number of Trays=10
Tray Parameters = 1 pass Valve Trays, 2 ft spacing, 2 inch weirs,
64 % Active Area
Overhead Condenser Temperature = 165 F
Overhead Condenser Pressure = 11.5 psig
Reboiler Pressure = 15 psig

Process Parameters:
Lean Solution Rate =100 USgpm
DEA Concentration = 16 wt percent
MDEA Concentration = 28 wt percent

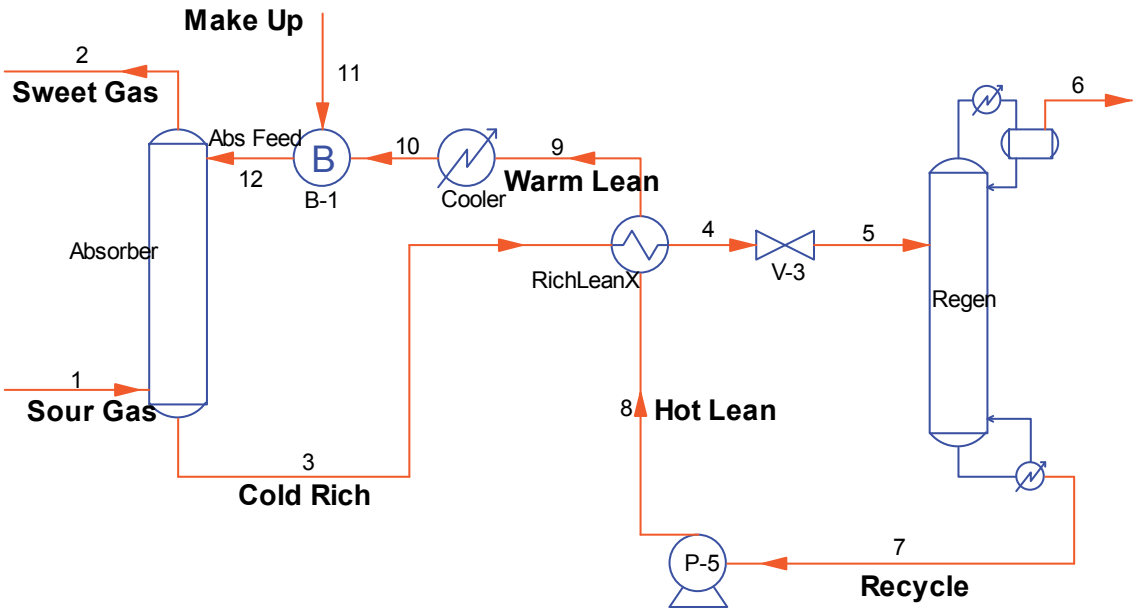
```
AB123.
*Sour Gas treatment with Mixed Amine

C- EQUIPMENT MODULES
```

```
AMCL 1 = Absorber, 4,1,-2,-3,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*4,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*63.6,
  LOC FEE = 1,10
  PRE TOP(PSIG) = 880
  PRE DEL = DPC
  PRO (LBMOL/HR) = 2000,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

AMCL 3 = Regenerator, 7,-9,-8,
  REG1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*64,
  LOC FEE = 2,
  PRE TOP(PSIG) = 11.5
  PRE DEL = DPC
  PRE CON(PSIG) = 11.5
  PRE REB(PSIG) = 15
  PRO (LBMOL/HR) = 100,0
  TEM CON(F) = 165
  REB DUT(BTU/HR) = 6700000
  SYS FAC = 0.85
  MAX ITER = 200
  TOL = 0.001

PUM 4 = P-4, 16,-10,
  PRE OUT(PSIG) = 880
  EFF = 0.9
  ELE DRI

HEA EXC 5 = Rich/Lean, 10,15,-11,-6,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

HEA EXC 6 = Cooler, 11,-4,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 7 = V-7, 6,-7,
  PRE OUT(PSIA) = 25.2

VAL 8 = V-8, 3,-13,
  PRE OUT(PSIG) = 75

FLA 9 = F-9, 13,-14,-15,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0
```

```
MAS BAL 10 = B-10, 5,8,-16,
  REC(GAL/MIN) = 105
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 5
  INP = -1
  OUT = -1

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,49,50,2,3,4,5,6,4051,7051,
  NAM 62 = 'WATER'
  NAM 46 = 'NITROGEN'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 5 = 'ISOBUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  AME UNI OUT
  MAX = 15
  CON TOL = 0.001
  REC STR = 8,4,
  ONE LOOP
  FLO (FRA) 1 = 0,0.0017,0.056,0.0015,0.75119,0.11895,0.04799,0.00716,0.01551,
  TOT FLO (MMSCF/DAY) 1 = 24.8
  TP (F,PSIG) 1 = 97,880
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  NAM STR 3 = Cold Rich
  FLO (FRA MAS) 4 = 0.51,8*0,0.16,0.33,
  TOT FLO (GAL/MIN) 4 = 105
  TP (F,PSIG) 4 = 127,880
  NAM STR 4 = Amine Feed
  FLO (FRA) 5 = 1,
  TOT FLO (GAL/MIN) 5 = 1
  TP (F,PSIA) 5 = 85,30
  NAM STR 5 = Makeup
  NAM STR 6 = Hot Rich
  NAM STR 7 = Strm 7
  FLO (FRA MAS) 8 = 0.51,8*0,0.16,0.33,
  TOT FLO (GAL/MIN) 8 = 105
  TP (F,PSIG) 8 = 248,15
  NAM STR 8 = Hot Lean
  NAM STR 9 = Off Gas
  NAM STR 10 = Strm
  NAM STR 11 = Warm Lean
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
END
```

MxpInt2.psd - Mixed Amine Plant 2

The objective of this simulation is to contact CO2 (7 mol%) and H2S (2 mol %) in the natural gas stream with an aqueous stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .005 mol % of H2S and 1.25 mole % of CO2 per 25 MMSCFD of gas processed.



Assumptions

Use 10 weight % solution of DEA.blended with 39 wt % of MDEA to make about 250 gal/min of aqueous recirculating amine. Make-up water and the amine losses are estimated and made-up automatically by the Inline Fortran code already set-up in the regenerator.

Objectives

To achieve 1.25 mol % of CO2 in the sweet gas, with top temperature of 137 F

Coding

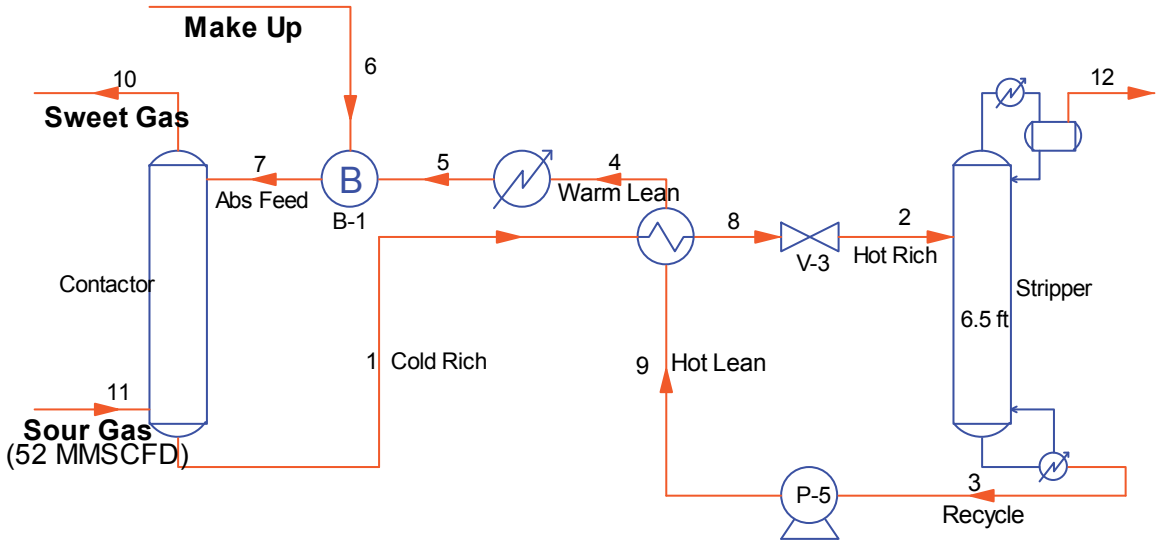
```
AB123.
*Mixed Amine Example MEA/MDEA

C- EQUIPMENT MODULES

MAS BAL 1 = B-1, 11,10,-12,
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 11
  INP = -1
  OUT = -1

HEA EXC 2 = Cooler, 9,-10,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 125
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 3 = V-3, 4,-5,
  PRE OUT(PSIG) = 35.3

HEA EXC 4 = RichLeanX, 8,3,-9,-4,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 25
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

PUM 5 = P-5, 7,-8,
```

```
  PRE OUT(PSIG) = 1000
  EFF = 0.9
  ELE DRI

AMCL 6 = Absorber, 12,1,-2,-3,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(IN) = 10*54,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 1,10
  PRE TOP(PSIG) = 1000
  PRE DEL = DPC
  PRO (LBMOL/HR) = 0,4000
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

AMCL 7 = Regen, 5,-6,-7,
  REG1
  TRA = 10
  TRA TYP = 10*SIE,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3.5,
  TRA WEI(IN) = 10*1.5,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*80,
  TRA PER OPE = 10*15,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28
  PRE DEL = DPC
  PRE CON(PSIG) = 12.6
  PRE REB(PSIG) = 15.4
  PRO (LBMOL/HR) = 253,0
  REF RAT = 0.87
  SYS FAC = 0.85
  MAX ITER = 300
  TOL = 0.007

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  NAM 62 = 'WATER'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 2 = 'METHANE'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  AME UNI OUT
  MAX = 20
  CON TOL = 0.003
  REC STR = 7,12,
  REC SEQ = 6,4,3,7,5,4,2,1,
  SAT FEE WAT = 1,
  FLO(LBMOL/HR)1 = 2.565,2*0,54.851,191.978,2495.709,
  TP (F,PSIG) 1 = 100,1000
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO (FRA) 3 = 0.79984,0.0274842,0.0969907,0.0235,0.05138,0.00080298,
  TOT FLO (LBMOL/HR) 3 = 4156.6623
  TP (F,PSIA) 3 = 179.19,1014.696
  NAM STR 3 = Cold Rich
  NAM STR 4 = Strm 8
```

```
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 12
  FLO(LBMOL/HR)7 = 3363,114,403,4,6,0,
  TP (F,PSIG) 7 = 256.9,15.4
  NAM STR 7 = Recycle
  NAM STR 8 = Hot Lean
  NAM STR 9 = Warm Lean
  NAM STR 10 = Strm 5
  FLO (FRA) 11 = 1,
  TOT FLO (LBMOL/HR) 11 = 0.01
  TP (F,PSIG) 11 = 125,1000
  NAM STR 11 = Make Up
  FLO(LBMOL/HR)12 = 3386,114,403,5,6,0,
  TP (F,PSIA) 12 = 125,1000
  NAM STR 12 = Abs Feed
END
```

MxpInt3.psd - CO2 / H2S Removal with only MEA

The objective of this simulation is to contact CO2 (2 mol%) and H2S (.5 mol %) in the natural gas stream with an aqueous stream with a one amine, MEA. Note that for using pure MEA, the other amine (DEA) can be a negligible amount but should be present in the flowsheet so that Mixed Amine thermo can be used. In the sweet gas stream the acid gases should be no greater than 500 ppm of H2s and 0.025 mole % of CO2 per 52 MMSCFD of gas processed.



Assumptions

Use 15 weight % solution of MEA.blended with 0 wt % of DEA to make about 370 gal/min of aqueous recirculating amine.
Make-up water and the amine losses are estimated and made-up automatically by the Mass Balance unit module.

Objectives

To achieve less than 500 ppm of H2S in the sweet gas, with top temperature of 107 F

Coding

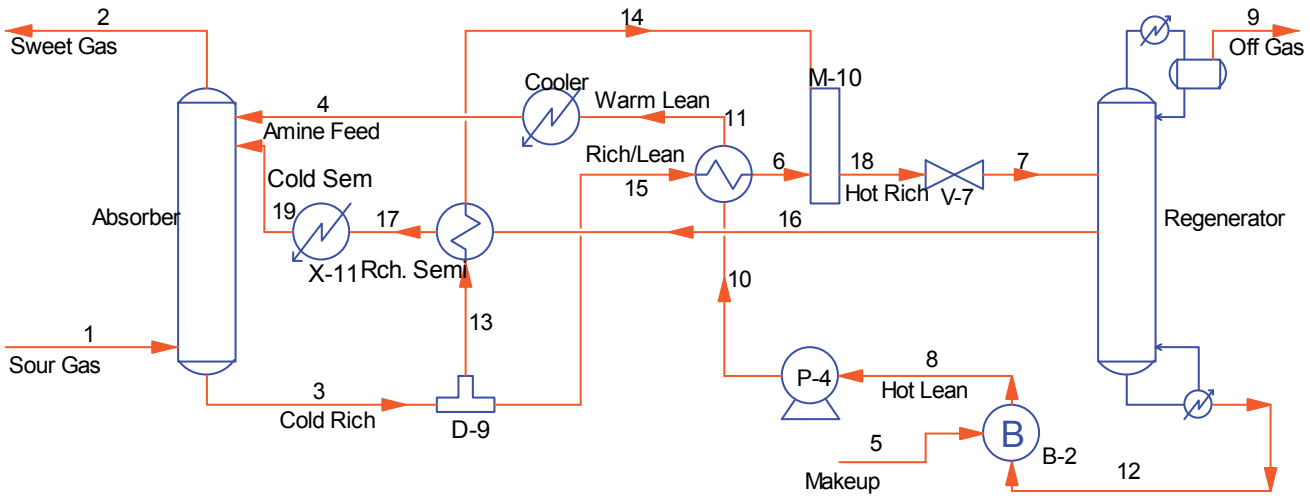
```
AB123.
*Mixed Amine Example MEA 15 % /MDEA- 0 %

C- EQUIPMENT MODULES

MAS BAL 1 = B-1, 6,5,-7,
  REC(GAL/MIN) = 350
  ACC = 0.8
  IDS = 62,
  VAR = 6
  INP = -1
  OUT = -1

HEA EXC 2 = X-2, 4,-5,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 110
```

```
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 3 = V-3, 8,-2,
  PRE OUT(PSIA) = 26

HEA EXC 4 = X-4, 9,1,-4,-8,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 25
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

PUM 5 = P-5, 3,-9,
  PRE OUT(PSIA) = 890
  EFF = 0.9
  ELE DRI

AMCL 6 = Contactor, 7,11,-10,-1,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(IN) = 10*55,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 1,10
  PRE TOP(PSIA) = 890
  PRE BOT(PSIA) = 900
  PRO (LBMOL/HR) = 5000,0
  SYS FAC = 0.75
  MAX ITER = 200
  TOL = 0.0005

AMCL 7 = Stripper, 2,-12,-3,
  REG1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*6.5,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 2,
  PRE TOP(PSIG) = 12
  PRE DEL = DPC
  PRE CON(PSIG) = 12
  PRE REB(PSIG) = 14.5
  PRO (LBMOL/HR) = 145,0
  TEM CON(F) = 119
  REB DUT(BTU/HR) = 14500000
  SYS FAC = 0.75
  MAX ITER = 200
  TOL = 0.001

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,50,62,4155,4051,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 62 = 'WATER'
  NAM 4155 = 'MEA'
```

```
  NAM 4051 = 'DEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  AME UNI OUT
  MAX = 15
  CON TOL = 0.005
  REC STR = 7,3,
  NAM STR 1 = Cold Rich
  FLO (FRA) 2 = 0.00080298,2*0,0.05138,0.0235,0.79984,0,0.0274842,
  TOT FLO (LBMOL/HR) 2 = 4156.6623
  TP (F,PSIA) 2 = 179.19,50
  NAM STR 2 = Hot Rich
  FLO(LBMOL/HR)3 = 3*0,60.567,1.052,8575.602,446.398,0,
  TP (F,PSIA) 3 = 230,890
  NAM STR 3 = Recycle
  NAM STR 4 = Warm Lean
  FLO(LBMOL/HR)5 = 1,2*0,86.9,5.3,9431.1,491,0,
  TP (F,PSIA) 5 = 110,914.7
  NAM STR 5 = Strm 5
  FLO (FRA) 6 = 5*0,1,
  TOT FLO (LBMOL/HR) 6 = 0.01
  TP (F,PSIA) 6 = 110,890
  NAM STR 6 = Make Up
  FLO(LBMOL/HR)7 = 1,2*0,60.567,1.052,8575.602,446.398,0,
  TP (F,PSIA) 7 = 110,890
  NAM STR 7 = Abs Feed
  NAM STR 8 = Strm 8
  NAM STR 9 = Hot Lean
  NAM STR 10 = Sweet Gas
  FLO(LBMOL/HR)11 = 5014,557.1,0,111.4,27.86,3*0,
  TP (F,PSIA) 11 = 90,914.7
  NAM STR 11 = Sour Gas
  NAM STR 12 = Strm 12
     CONverge on TEMperature = 3, 7
END
```

MxpIntsd.psd - Mixed Amine Plant with Side Draw and Side Feed

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



Assumptions

Use 16 weight % solution of DEA.blended with 28 wt % of MDEA to make about 100 gal/min of aqueous recirculating amine.

Make-up water is estimated automatically by the Inline Fortran code already set-up in the regenerator.

Objectives
To achieve 1.7 mol % of CO2 in the sweet gas, with top temperature of 150 F
Sidedraw from tray#5 is 2 lbmol/hr back to absorber as side feed.

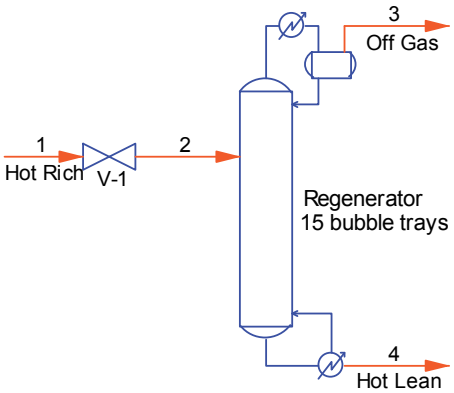
```
AB123.
*Sour Gas treatment with Mixed Amine-Side Feed & Side Draw

C- EQUIPMENT MODULES

AMCL 1 = Absorber, 4,19,1,-2,-3,
  ABS2
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*4,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*63.6,
  LOC FEE = 1,2,10
  PRE TOP(PSIG) = 880
  PRE DEL = DPC
  PRO (LBMOL/HR) = 2000,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

MAS BAL 2 = B-2, 5,12,-8,
  REC(GAL/MIN) = 105
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 5
  INP = -1
  OUT = -1

AMCL 3 = Regenerator, 7,-9,-16,-12,
  REG2
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*64,
  LOC FEE = 2,
  LOC SID = 5
  PRE TOP(PSIG) = 11.5
  PRE DEL = DPC
  PRE CON(PSIG) = 11.5
  PRE REB(PSIG) = 15
  PRO (LBMOL/HR) = 100,2,0
  TEM CON(F) = 165
  REB DUT(BTU/HR) = 6700000
  SYS FAC = 0.85
  MAX ITER = 200
  TOL = 0.001

PUM 4 = P-4, 8,-10,
  PRE OUT(PSIG) = 880
  EFF = 0.9
  ELE DRI

HEA EXC 5 = Rich/Lean, 10,15,-11,-6,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

HEA EXC 6 = Cooler, 11,-4,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
```

```
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 7 = V-7, 18,-7,
  PRE OUT(PSIA) = 25.2

HEA EXC 8 = Rch. Semi, 16,13,-17,-14,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

DIV 9 = D-9, 3,-13,-15,
  FLO(FRA) = 0.05

MIX 10 = M-10, 14,6,-18,

HEA EXC 11 = X-11, 17,-19,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,49,50,2,3,4,5,6,4051,7051,
  NAM 62 = 'WATER'
  NAM 46 = 'NITROGEN'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 5 = 'ISOBUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  AME UNI OUT
  MAX = 100
  CON TOL = 0.001
  REC STR = 8,4,3,16,
  FLO (FRA) 1 = 0,0.0017,0.056,0.0015,0.75119,0.11895,0.04799,0.00716,0.01551,
  TOT FLO (MMSCF/DAY) 1 = 24.8
  TP (F,PSIG) 1 = 97,880
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO(LBMOL/HR)3 = 1556.8,0,114.74,4.11,1.28,0.17,0.04,0.001,0.007,77.57,120,
  TP (F,PSIG) 3 = 142.4,880
  NAM STR 3 = Cold Rich
  FLO(LBMOL/HR)4 = 1572,0,3.4,0.001,5*0,77.4,119.6,
```

```
  TP (F,PSIG) 4 = 127,880
  NAM STR 4 = Amine Feed
  FLO (FRA) 5 = 1,
  TOT FLO (LBMOL/HR) 5 = 1
  TP (F,PSIA) 5 = 85,30
  NAM STR 5 = Makeup
  NAM STR 6 = Hot Rich
  NAM STR 7 = Strm 7
  FLO (FRA) 8 = 0.6,0,0.01,0.001,0,3*0.005,0.015,0.16,0.33,
  TOT FLO (GAL/MIN) 8 = 105
  TP (F,PSIG) 8 = 248,15
  NAM STR 8 = Hot Lean
  NAM STR 9 = Off Gas
  NAM STR 10 = Strm
  NAM STR 11 = Warm Lean
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  FLO(LBMOL/HR)16 = 2,
  TP (F,PSIA) 16 = 150,14.7
  NAM STR 16 = Strm 16
  NAM STR 17 = Strm 17
  NAM STR 18 = Strm 18
  NAM STR 19 = Cold Sem
C- REC SEQ=4,1,5,6,2,7,3,4,5,6,2
END
```

Mxregn.psd - Mixed Amine Stripper Example

The objective of this simulation is to strip the enriched mixed amine free of water and residual light ends by contacting with a mixture of two amines (DEA and MDEA). The rich hot lean amine exits at the bottom of stripper with no loss of amine at the top of the regenerator in the off gas.



Assumptions
Use 36 wt % MDEA and 9 wt% DEA

Objectives
1E-05 mol % of Amines in the off gas, with top temperature of 125 F

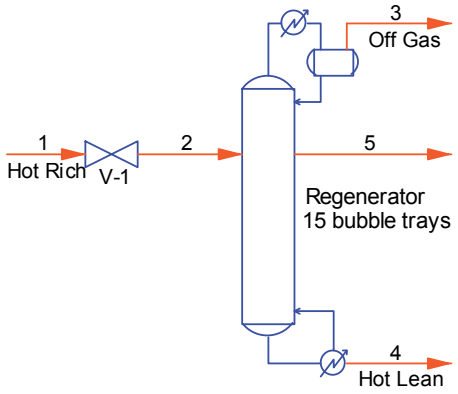
Specification:
Tower diameter 3 ft inches. Active area 80%.
Tower internals - 15 bubble trays, two pass, 2 ft tray spacing and 3 inches weir height.

```
AB123.
*Mixed Amine Stipper Example

C- EQUIPMENT MODULES

VAL 1 = V-1, 1,-2,
  PRE OUT(PSIA) = 50.

AMCL 2 = Regenerator, 2,-3,-4,
  REG1
  TRA = 15
  TRA TYP = 15*BUB,
  TRA PAS = 15*2,
  TOW DIA(FT) = 15*3.,
```

```
  TRA WEI(IN) = 15*3.,
  TRA SPA(FT) = 15*2.,
  TRA ARE PER TOW = 15*80.,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIA) = 27.3
  PRE REB(PSIA) = 29.66
  PRO (LBMOL/HR) = 250.,0
  TEM CON(F) = 125.
  REB DUT(BTU/HR) = 12005700.
  SYS FAC = 0.75
  MAX ITER = 150
  TOL = 1.E-003

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 10
  CON TOL = 1.E-003
  NAM STR 3 = Off Gas
  FLO(FRA)1 = 0.7998,2.748E-002,9.699E-002,2.35E-002,5.138E-002,8.029E-004,
  TOT FLO(LBMOL/HR)1 = 4156.662
  TP (F,PSIA) 1 = 230.,1000.
  NAM STR 1 = Hot Rich
  NAM STR 2 = Strm 2
  NAM STR 4 = Hot Lean
END
```

Mxregnsd.psd - Amine Stripper with Side Draw

The objective of this simulation is to strip the enriched mixed amine free of water and residual light ends by contacting with a mixture of two amines (DEA and MDEA). A side draw at tray #5 is also used. The rich hot lean amine exits at the bottom of stripper with no loss of amine at the top of the regenerator in the off gas.



Assumptions
Use 36 wt % MDEA and 9 wt% DEA

Objectives
1E-05 mol % of Amines in the off gas, with top temperature of 125 F.
Specification:
Tower diameter 3 ft inches. Active area 80%.
Tower internals - 15 bubble trays, two pass, 2 ft tray spacing and 3 inches weir height.
Side draw 2 lbmol/hr at tray 5.

```
AB123.
*Mixed Amine Stipper Example -Side Draw
```

```
C- EQUIPMENT MODULES

VAL 1 = V-1, 1,-2,
  PRE OUT(PSIA) = 50.

AMCL 2 = Regenerator, 2,-3,-5,-4,
  REG2
  TRA = 15
  TRA TYP = 15*BUB,
  TRA PAS = 15*2,
  TOW DIA(FT) = 15*3.,
  TRA WEI(IN) = 15*3.,
  TRA SPA(FT) = 15*2.,
  TRA ARE PER TOW = 15*80.,
  LOC FEE = 1,
  LOC SID = 5
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIA) = 27.3
  PRE REB(PSIA) = 29.66
  PRO (LBMOL/HR) = 250.,2.,0
  TEM CON(F) = 125.
  REB DUT(BTU/HR) = 12005700.
  SYS FAC = 0.75
  MAX ITER = 150
  TOL = 1.E-003

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 10
  CON TOL = 1.E-003
  NAM STR 3 = Off Gas
  FLO(FRA)1 = 0.7998,2.748E-002,9.699E-002,2.35E-002,5.138E-002,8.029E-004,
  TOT FLO(LBMOL/HR)1 = 4156.662
  TP (F,PSIA) 1 = 230.,1000.
  NAM STR 1 = Hot Rich
  NAM STR 2 = Strm 2
  NAM STR 4 = Hot Lean
  NAM STR 5 = Strm 5
END
```

Chapter 14: Batch Distillation Column Samples - c:\designii\samples\equipmnt\batch

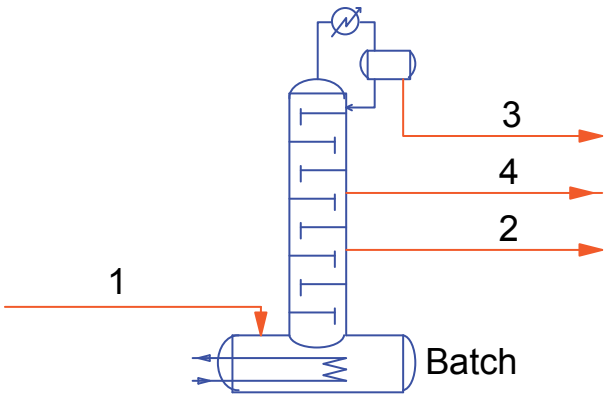
Actmohct.in – Acetone, Water, Methanol VLE Data Regression

It is a ChemTran VLE data regression input file for acetone, water and methanol mixture.

Acwtmeoh.dat – Acetone, Water, Methanol VLE Data Regression

It is a ChemTran created binary data file created by the input file above for later use by DESIGN II during runtime.

Batch1.psd – Acetone / Methanol / Water Example



In the following example, an initial charge of acetone, methanol and water is to be batch distilled and the resulting overhead product collected at different stages into separate tanks.

The first event models the initial boil-up period when the column operates at total reflux and without any product removal. This is required since the initial composition on every stage is the same as that entered for the initial column charge. Product draw off should not occur until steady state conditions have been reached within the column. When the column is operating at total reflux only one heat and material balance specification is required, which in this case is the reboiler duty. The initial charge composition is defined by stream number 1 in the GENeral section and the total charge in the column by the total holdup.

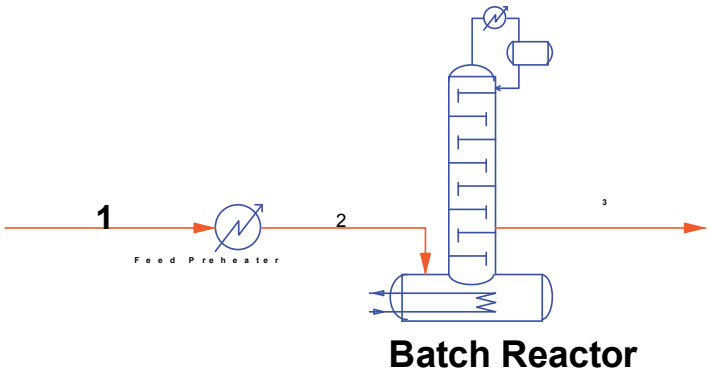
The second event starts immediately after the first event has terminated. Overhead product is collected instantly and continues until the tank composition of the most volatile component, acetone, drops to 73 mole %. In this case the column requires two heat and material balance specifications - reboiler duty and reflux ratio. The pressure and holdup information entered for the first event continues throughout the second event.

In the third event, overhead product is collected into two separate tanks with the changeover occurring after 0.35 hours from the start of this event. The latter of these tanks, represented by stream number 4, collects the highest concentration of the second most volatile component, methanol. The heat and material balance specifications, pressure and holdup information from the second event are repeated in this event. The event stops after 0.8 hours of operations

```
AB123.
* BATCH EXAMPLE.  ACETONE-METHANOL-WATER SYSTEM.

BATCH 1 = COL,1,-2,-3,-4,
   TOTAL
   STAGES = 7
   LOC SIDE = 1,1,1

   START EVENT 1 = BOIL-UP, STEADY STATE
      HOLDUP (KGMOL) 1,1 = 5.0
      HOLDUP (KGMOL) 2,6 = 1.0
```

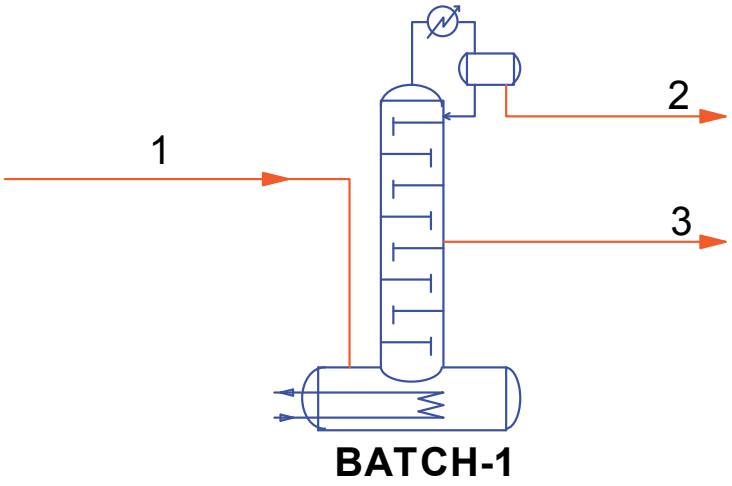
```
      HOLDUP (KGMOL) 7,7 = 90.0
      PRE (MMHG)     1,7 = 760
      REB DUTY (KCAL/HR) = 1.0E6
      HISTORY TIME (HR) LIQ COM = 0.05,1,7
      STEADY STATE TOLERANCE = .001
   STOP WHEN (TIME.GT.2.0 .OR. STEADY STATE)

   START EVENT 2 = PRODUCT, DRAW OFF
      REFLUX RATIO = 3.0
      REB DUTY (KCAL/HR) = 1.0E6
      SIDE = 1
   STOP WHEN (AVE PURITY SIDE 1 COM 1 .LT. 0.73)

   START EVENT 3 = SIDE, DRAW OFF
      REFLUX RATIO = 3.0
      REB DUTY (KCAL/HR) = 1.0E6
      SIDE = 2(TIME .LT. 0.35),3(TIME .GT. 0.35)
   STOP WHEN (TIME .GT. 0.8)

GENERAL
   COMPONENTS = 1002, 1021, 62
   FLOW (FRA) 1 = 0.3,0.3,0.4
   TOTAL FLOW (KGMOL/HR) 1 = 100
   TP (C,MMHG) 1 = 60,760
   UNIFAC
END
```

Batch2.psd - Batch Reactions For Ethyl Acetate



The batch distillation module can also be used to conduct batch tray reactions. This simulation shows the manufacture of ethyl acetate from acetic acid and ethanol.

Batch3.psd – Benzene / Toluene Separation



This example is for batch distillation of benzene and toluene with plots.

Bentolct.in – Benzene and Toluene VLE Data Regression

It is a ChemTran VLE data regression input file for benzene and toluene mixture.

Benztolu.dat – Benzene and Toluene VLE Data Regression

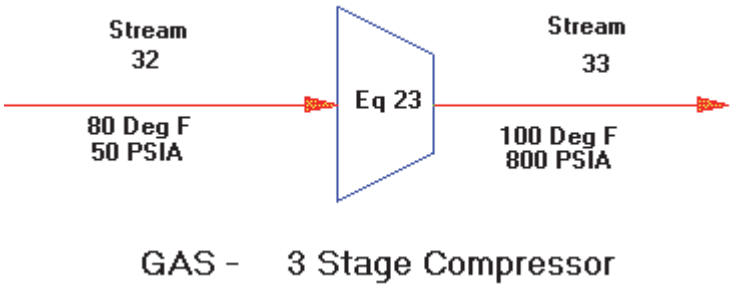
It is a ChemTran created binary data file created by the input file above for later use by DESIGN II during runtime.

Chapter 15: Compressor Samples - c:\designii\samples\equipmnt\compre

Compr1.psd - Multistage Compression

Compress an 80ºF gas stream from 50 PSIA to 800 PSIA in three stages (125, 325, and 800 PSIA discharge pressures); interstage cooling to 100 ºF with 72% adiabatic efficiency. Find the total work required, fuel for gas driver, and cooling water required for interstage cooling.

```
COMPREssor 23 = GAS, 32, -33,
   PREssure OUT = 800
   ENTropy calculation, EFFiciency = 0.72
   STAges = 3,
   PREssure of STAges = 125, 325, 800
   TEMperature of INTercooler = 100
```



Method of Calculation

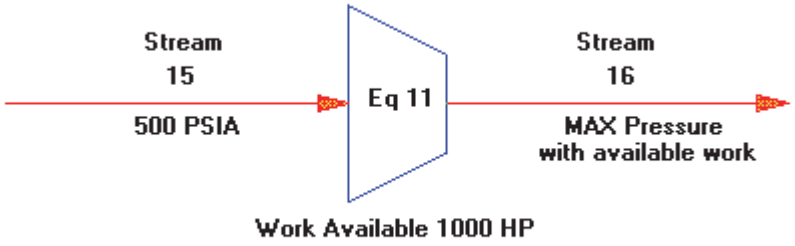
All properties of stream 32 are known. The stream is compressed to 125 PSIA at constant entropy, and then the adiabatic efficiency is applied to calculate the correct enthalpy and outlet temperature. The enthalpy required to cool the stream to 100 F is calculated. The stream (at 100 F and 125 PSIA) is then compressed to 325 PSIA using the same procedure.

The final compression from 100 F and 325 PSIA to 800 PSIA is performed **without** the cooling of the discharge stream to 100 F. The work required for each stage of compression is totaled. The fuel required is calculated assuming 980 BTU/SCF fuel and 8000 BTU/HR/HP. The cooling water required is calculated assuming a 15F temperature rise in the cooling water.

Compr2.psd - Specified Work Available

An existing 1000 HP compressor is to be used to compress a gas from 500 PSIA to the maximum obtainable pressure. A polytropic efficiency of 0.75 and a Cp/Cv ratio of 1.26 are assumed. Find the discharge pressure possible.

```
COMPRESsor 11 = COMP, 15, -16,
   PRESsure OUT = 2000
   WORk  AVAilable = 1000
   POLytropic COEfficient = 0.275,
   EFFiciency = 0.75
```



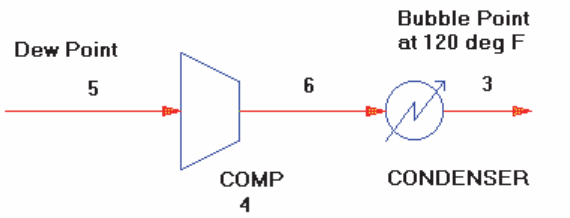
Method of Calculation

All properties of stream 15 are known. The stream is compressed to 2000 PSIA as specified. The work from this compression is calculated from the polytropic compressor equation and checked against the WORk AVAilable. The discharge pressure is recalculated until the WORk AVAilable is matched.

Note that a PRE OUT command was entered even though the desired specification was the WOR AVA. If the outlet pressure was not specified, the calculation would have proceeded as follows: stream 15 would be compressed to 1000 PSIA (default is twice the inlet pressure when no PRE OUT is specified). The compression work required to reach 1000 PSIA is calculated and compared to the specified WOR AVA of 1000 HP. If the calculation can reach 1000 PSIA without exceeding 1000 HP, the simulation will stop. Therefore, to achieve a desired work , specify a PRE OUT which is greater than can be obtained with the specified WORk AVAilable.

Compr3.psd - Bubble Point Specified

Design a compressor for a refrigeration cycle. The condenser exchange is an air cooler with a design outlet temperature of 120F. Use an adiabatic efficiency of 72%.



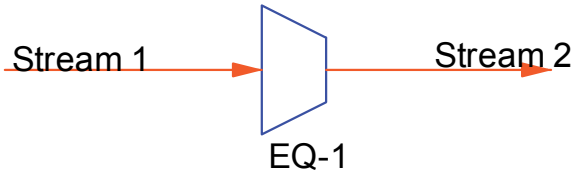

```
COMPREssor 4 = COMP, 5, -6,
   PREssure OUT is BUBble point,
   TEMperature of BUBble point = 120
   ENTropy calculation, EFFiciency = 0.72
```

Method of Calculation

Stream 5 is at its dew point coming from the refrigerant evaporator chiller. The stream is compressed at constant entropy to the bubble point pressure calculated at the specified temperature. The adiabatic efficiency is applied and the enthalpy and discharge temperature are calculated. The work required is calculated along with the amount of fuel gas (default drive) used for the driver.

Compr4.psd - Calculation of Polytropic Coefficient

The example shows a compressor module to calculate a polytropic coefficient, for a given compression pressure out.

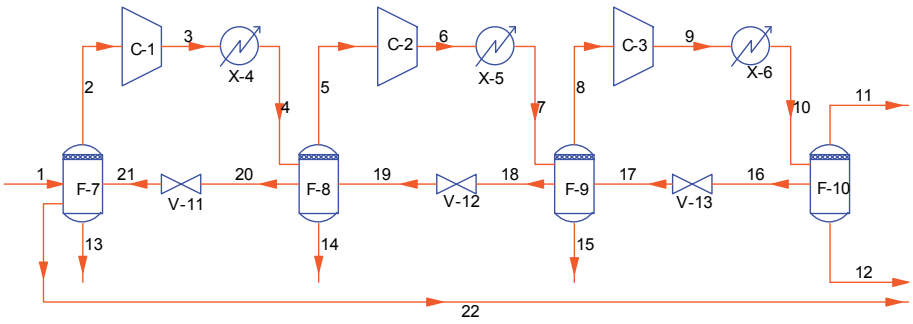



```
COMPREssor 1 = COM, 1, -2
   CALculate POLytropic coefficient
   PREssure OUT = 35
```

STAGES	1.000000
WORK CAPACITY HP	1000000.
CALCULATED PRESSURE OUT PSIA	35.00000
SPECIFIED PRESSURE OUT PSIA	35.00000
DRIVE TYPE	ELECTRIC
CALC TYPE	POLYTROP
POLY COEF (CALC)	0.2124675
EFF (FOR CALC TYPE)	0.7200000
POLYTROPIC HEAD FT-LBf/LBm	21300.06
KW USAGE	7.987961
REAL WORK HP	-10.72182
ISENTROPIC HEAD FT-LBf/LBm	20761.76
INLET T F	20.00000
INLET P PSIA	14.70000
INLET V FT3/MIN	126.9034
FLOWRATE LBMOL/HR	22.00000
FLOWRATE LB/HR	717.6049
DISCHARGE T F	112.9510
DISCHARGE V FT3/MIN	63.31862
CALCULATED Cp/Cv	1.178942

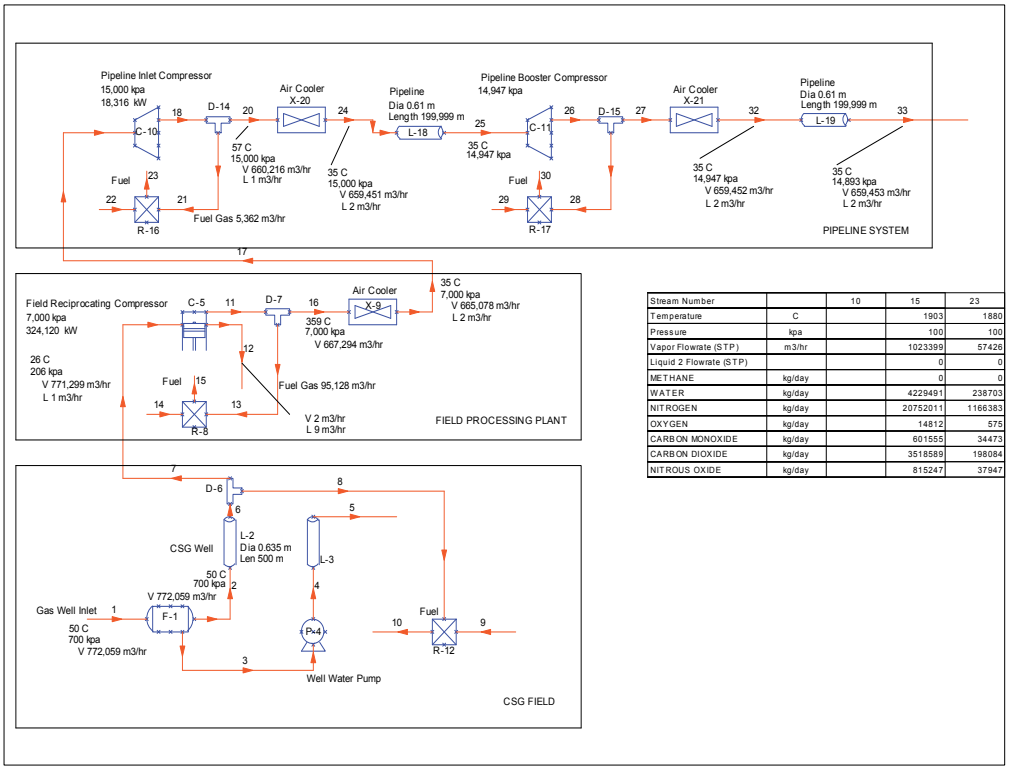
Compr5.psd - Cascaded Multi-Stage Compression

This flowsheet illustrates a three stage cascaded compressor station with intercoolers, re-flash, and valves.



Generic CSG-LNG-3.psd - Generic Coal-seam Feed Gas

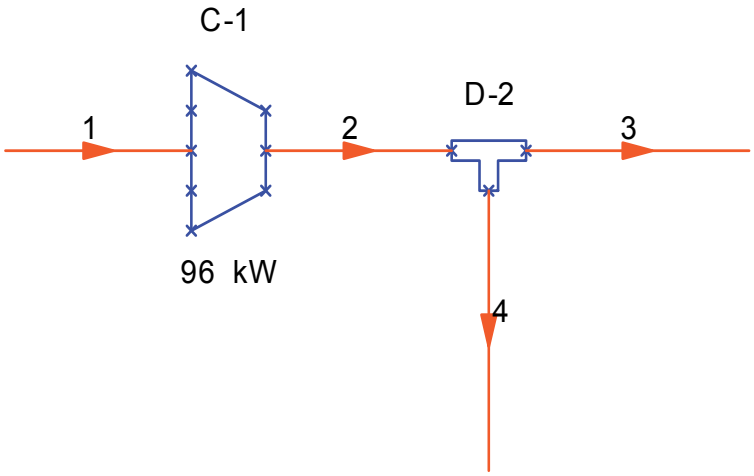
This is a generic coal-seam feed gas to a LNG plant.



Stream Number		10	15	23
Temperature	C		1903	1880
Pressure	kpa		100	100
Vapor Flowrate (STP)	m3/hr		1023399	57426
Liquid 2 Flowrate (STP)			0	0
METHANE	kg/day		0	0
WATER	kg/day		4229491	238703
NITROGEN	kg/day		20752011	1166383
OXYGEN	kg/day		14812	575
CARBON MONOXIDE	kg/day		601555	34473
CARBON DIOXIDE	kg/day		3518589	198084
NITROUS OXIDE	kg/day		815247	37947

Inline Fuel Compressor.psd - Inline Fuel Compressor Flowsheet

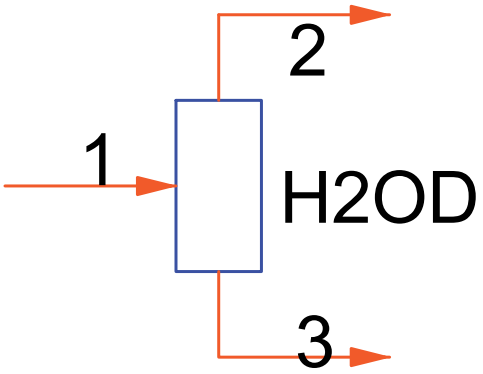
Shows a stream being compressed and divided.


C-1
D-2
1
2
3
4
96 kW

Chapter 16: Component Splitter Samples - c:\designii\samples\equipmnt\comspl

Comp1.psd - Water Decant

Component splitters are used to separate a known amount of a particular component from the feed mixture. This example is set to remove 98.5% of the water from the liquid feed stream to the bottoms.



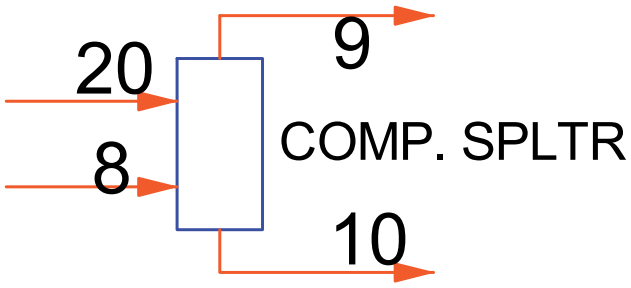

```
COMponent SPLitter 3 = H2OD, 4, -5, -6,
        RECovery TOP = 8*1, 0.015
GENeral,
        COMponents = 2,3,4,5,6,7,8,10,62
```

Method of Calculation

Since 98.5% of the water is to go to the BOTtoms stream, the RECovery to the TOP stream is 1.5%. The "8*1" specifies 100% of the first 8 components go to the top stream. Since product temperatures are not specified, the feed temperature is used by the program.

Comp2.psd - Shortcut Column Calculation

When columns are located inside a process loop (which means they must be calculated with each iteration), a good idea is to shortcut the column to minimize calculation time. Columns can be shortcut with the SHOrtcut fractionator module if they have both a condenser and reboiler and only a top and bottom product. Other columns can be shortcut with the COMponent SPLitter Module.



The rich oil demethanizer shown below can be shortcut with the COMponent SPLitter and some very valuable information obtained. Not only do you get an excellent guess for the top and bottom product temperature (needed for the rigorous column calculation DIStillation module), but also, the total reboiler requirements can be obtained by performing a heat balance around the column. Since you may be interested in a side heater (for which a duty must be specified in the DIStillation modules), this allows you to develop a realistic view based on the total required duty.

Assumed Recoveries,

Percent	Component
100.0	N2
50.0	CO2
99.5	C1

70.0	C2
25.0	C3
2.0	i-C4
2.0	n-C4
--	i-C5
--	n-C5
--	C6
--	C8

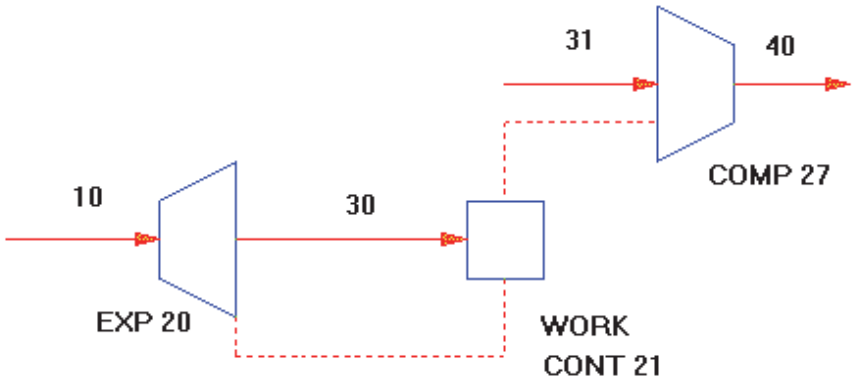
```
COMponent SPLitter 5 = DEC1, 20, 8, -9, -10
RECovery TOP(FRAction) = 1.0, 0.5, 0.995, 0.7, 0.25,.02, 0.02, 4*0.
TEMperature TOP DEW,
TEMperature BOTtom BUBble
GENeral,
COMponents = 46, 49, 2, 3, 4, 5, 6, 7, 8, 10, 100
END
```

Method of Calculation

Feed streams 20 and 8 are adiabatically mixed and then the recovery factors applied to obtain the top and bottom products. The dew point for the top product and bubble point for the bottom product are calculated.

Chapter 17: Controller Samples - c:\designii\samples\equipmnt\cont

Contr1.psd - Information Passing



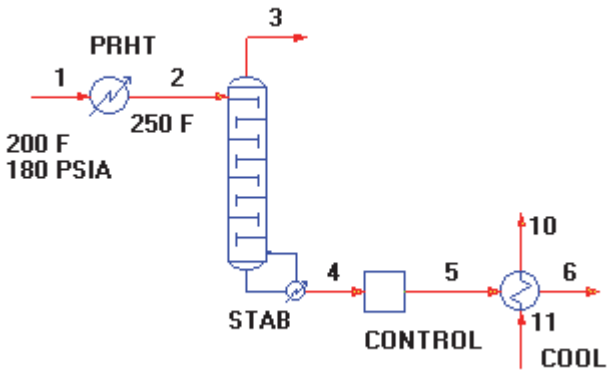
Use CONTroller 21 to send 98% of the calculated work produced by an expander (EXP 20) to be used as available work for a compressor (COM PRE 27). To simplify coding, all of the information required for the CONTroller command should be organized. For this example, the required information is as follows:

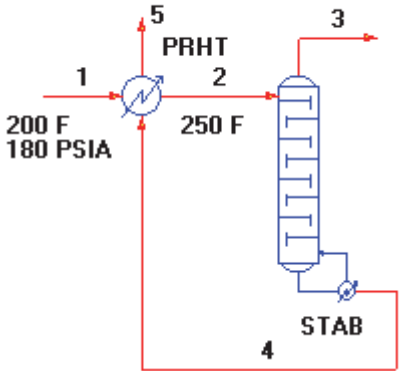
X1 is WORk AVAilable, an Equipment module command given in the COMPREssor section.
equipment is COMPREssor 27
X2 is CALculated WORk, a calculated equipment parameter given in Table 2.
SCAle is 0.98, the factor by which the CALculated WORk is to be multiplied (for shaft losses). It is positive because the calculated work for the expander is positive.
equipment or stream is EXPander 20

```
CONTroller 21 = WORK, 30, -31,
   SET WORk AVAilable OF COMPREssor 27 =
   CALculated WORk (SCAle = 0.98) FROm EXPander 20
```

Contr2.psd - Passing Information to Minimize Recycle Calculations

The top flowsheet would require at least 2 iterations to converge. The second flowsheet with a controller is not a recycle problem. The specified duty for HEAt EXChanger 4, -1.5E6 BTU/HR, is never actually used because it is overridden by the value passed to it by the CONTroller.



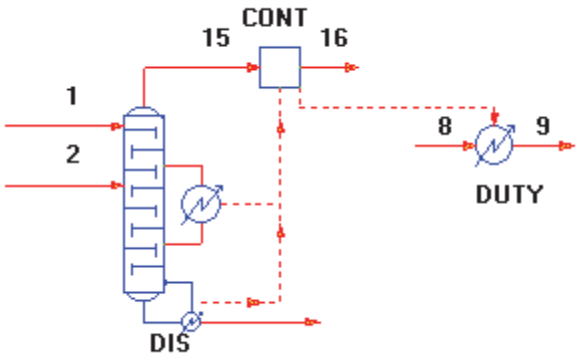



```
HEAt EXChanger 1 = PRHT, 1, -2,
       TEMperature OUT = 250,
DISTillation 2 = STAB, 2, -3, -4,
       ABSorber with REBoiler, ......
```

```
CONTroller 3 = DUTY, 4, -5,
       SET DUTy of HEAt EXChanger 4 =
       CALculated DUTy (SCAle = -1)FROm HEAt EXChanger 1
HEAt EXChanger 4 = COOL, 5, 11, -6, -10,
       DUTy = -1.5E6
GENeral,
       FLOw 11 = FLOw 1, TP 11 = TP 1
       ...
END
```

Contr3.psd - Passing Information-Two Parameters

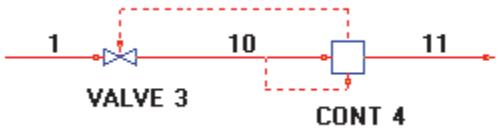
Data to be passed can also be calculated from two measured properties. For example, the second side heater and calculated reboiler duty of a column (DIS 10) can be added and used as the duty for the heating medium by using a dummy heat exchanger (HEA EXC 23). Remember to enter a DUT command with a negative value for HEA EXC 23 specification option.

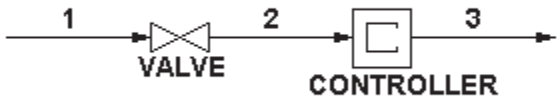



```
CONTroller 11 = DUTY, 15, -16,
   SET DUT of HEAEXC 23 =
   HEAT 2 (SCAle = - 1) FROm DIS 10 + CAL REB (SCA= -1) FRO DIS 10
```

Contr4.psd - Setpoint Convergence #1 Example

Vary the pressure out of a valve (Valve 3) until the temperature out (stream 10) is -10 C with a 1 C tolerance. The calculation of the valve module is repeated by changing the pressure of valve from a minimum to a maximum pressure value until the set point of -10 C is achieved at the outlet stream of the valve. Note that the streams 2 and 3 are of the same properties and no changes in mass or heat occur across the controller module.





Alternatively this may be done with one module.

```
FLAsh 3 = TOUT, 9, -11
   ADIbatic, TEMperature OUT (C) = -10
```

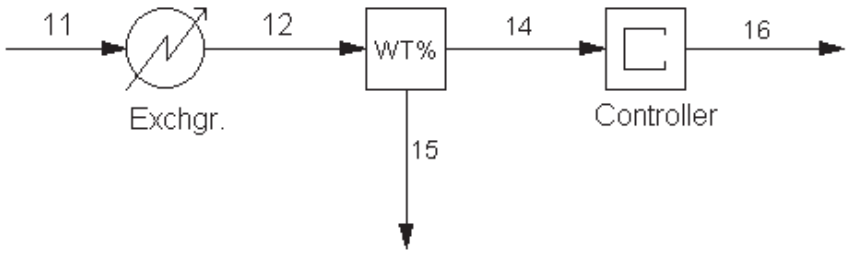
Coding

```
AB123.
*Setpoint Convergence Example

C- EQUIPMENT MODULES

VAL 1 = VALVE, 1,-2,
   PRE OUT(PSIA) = 100.
```

```
CONT 2 = CONTROLLER, 2,-3,
   VARY PRE OUT (MIN=100, MAX=300, STEP =20) OF VALve 1
   UNTIL TEMperature FROm STReam 2 = -10 (TOLerance=1,C)

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
  TEM UNI OUT = C
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 100.,20.,25.,
  TP (F,PSIA) 1 = 0.,40.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
END
```

Contr5.psd - Setpoint Convergence # 2 Example

In this example, the controller is set to vary the FACtor of the stream manipulator to achieve a flow of 37000 lb/hr of the component ID# 6 (n-Butane) in stream 14.

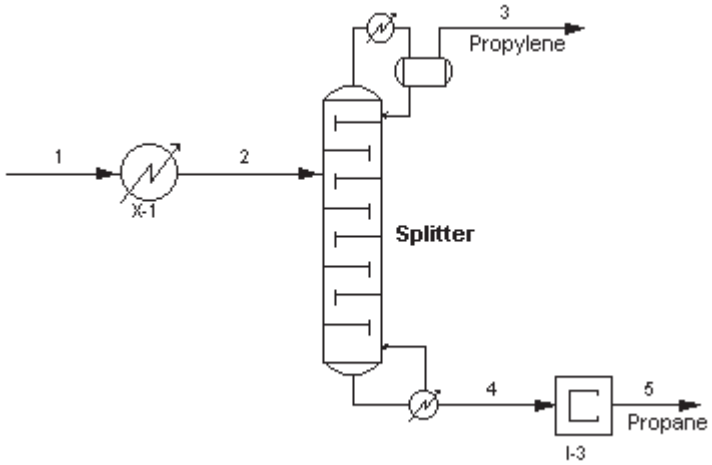



```
AB123.
*Mass flowrate for access by the CONTroller module

C- EQUIPMENT MODULES

STR 1 = WT%, 12,-15,-14,
  MUL
  SAM FAC, FAC = 1.

CONT 2 = Controller, 14,-16,
  VARY FAC (MIN=0.,MAX=100.,STE=5.) OF STR 1
  UNTIL FLO COM 6 (SCA = 1.) FROM STREAM 14 =
  37000. (BEG=1,LOO=30,LB/HR)

HEA EXC 3 = Exchgr., 11,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 80.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 2.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,6,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'N-BUTANE'
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
```

```
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)11 = 100.,26.,38.,11.,
  TP (F,PSIA) 11 = 70.,120.
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
END
```

Contr6.psd - Vary Reflux Ratio to achieve Product Purity

In this simulation the controller varies the reflux ratio of the distillation column, so that only 0.25 moles of propylene leave at the bottom of the distillation column. Note that the purity of the top or bottom stream is controlled by changing the specified reflux ratio.

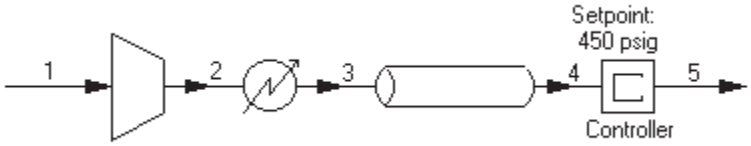



```
AB123.
*SETPOINT CONVERGENCE EXAMPLE 3

C- EQUIPMENT MODULES

HEA EXC 1 = X-1, 1,-2,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 80.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 0.

DIS 2 = Splitter, 2,-3,-4,
  PAR
  TRA = 155
  TEM TOP(F) = 112.5
  TEM BOT(F) = 131.
  PRE TOP(PSIA) = 269.
  DEL(PSI) = 10.
  LOC FEE = 100,
  PRO (LBMOL/HR) = 1480.,0.
  SUP
  MAX DIMEN = 157
  REF = 22.5
  PUR TOP (MOL, REL) 23 = 0.996
  LOC HEA = 101,
  HEA(BTU/HR) = 75000000.,

CONT 3 = I-3, 4,-5,
    VARY REF OF DIS 2
    UNTIL FLO COM 23 FROM STREAM 4 = .25

C- GENERAL SECTION
GENERAL
  COMPONENTS = 4,23,
  PENK
  PENH
```

```
  STDD
  MAX = 1
  CON TOL = 1.E-003
  FLO(LB/HR)1 = 7724.,82304.,
  TP (F,PSIA) 1 = 50.,275.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Propylene
  NAM STR 4 = Strm 4
  NAM STR 5 = Propane
END
```

Contr7.psd - Use Controller To Reach Desired Outlet Pressure

The controller in this simulation is set to vary the compressor pressure out to achieve a 450 psig at the outlet of the line module irrespective of the conditions of line module. All properties of stream 1 are known.



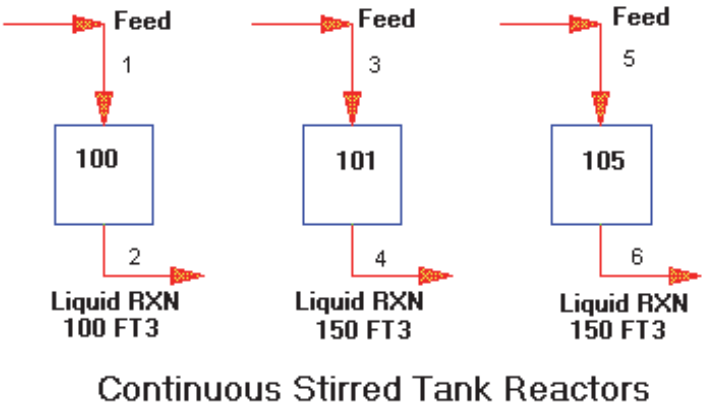

```
AB123.
*Line Module using Controller to reach outlet pressure desired.(LINE, PRE OUT)

C- EQUIPMENT MODULES

COMPRE 1 = C-1, 1,-2,
  PRE OUT(PSIA) = 800.
  EFF = 0.72
  CAL POL
  ELE DRI
  STA = 1

LINE 2 = L-2, 3,-4,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 12. MILE
  LEN (MILE) = 2.,2.,2.,2.,2.,2.,
  SEG LEN (MILE) = 1.2
  TEM SUR(F) = 80.,80.,80.,80.,80.,80.,
  U(BTU/HR/FT2/F) = 1.8,1.8,1.8,1.8,1.8,1.8,

HEA EXC 3 = X-3, 2,-3,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 120.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 0.

CONT 4 = Controller, 4,-5,
   VARY PRESSURE OUT OF COMPRESSOR 1
   UNTIL PRESSURE FROM STREAM 4 = 450 (PSIG)

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 1
  CON TOL = 1.E-003
  FLO(FRA)1 = 0.98,2.E-002,
  TOT FLO(MMSCF/DAY)1 = 50.
  TP (F,PSIA) 1 = 80.,400.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
END
```

Chapter 18: Continuous Stirred Tank Reactor (CSTR) Samples - c:\designii\samples\equipmnt\cstr

Cstr1.psd - Liquid Reactions - Continuous Stirred Tank Reactor Example

Each reactor model assumes a steady state flow of species for the reactor and that the concentrations of species and temperature are uniform throughout the reactor. Multiple liquid or gas phase reactions is specified, along with Frequency, Order, and Activation Energy values of each participating reaction.

```
AB123.
*CSTR EXAMPLE PROBLEM
CSTR 100 = LIQUID RXN,1,-2,VOLume = 150,
   TEMperature OUT = 220,
   R 1 = (1*1001 + 1*1022 = 1*2031 + 1*62),
   R 3 = (1*1001 + 1*1012 = 1*1031 + 1*62),
   R 2 = (1*1031 + 1*62 = 1*1001 + 1*1012),
   ORDer 1 = 1001,1.5,1022,.5,
   ORDer 2 = 1031,1,62,1,
   ORDer 3 = 1001,1,1012,1,
   FREquency (CM3,GMOL,MIN) =
        1,17.4,2,1.63,3,4.76,
   ACTivation energy = 1,0.0,2,0.0,3,0.0,
CSTR 101 = LIQUID RXN,3,-4,
   VOLume = 150,
   ADIabatic,
   HEAt added= -1.5E5,
   R 1 = (1*1001 + 1*1022 = 1*2031 + 1*62),
   R 3 = (1*1001 + 1*1012 = 1*1031 + 1*62),
   R 2 = (1*1031 + 1*62 = 1*1001 + 1*1012),
   ORDer 3 = 1001,1,1012,1,
   ORDer 2 = 1031,1,62,1,
   ORDer 1 = 1001,2,1022,0,
   FREquency (CM3,GMOL,MIN) = 1,17.4,2,1.63,3,4.76,
   ACTivation energy = 1,0.0,2,0.0,3,0.0,
CSTR 105 = LIQUID RXN,5,-6,
       VOLume = 150,TEMperature OUT = 220,
F-             DURING
F-             RATE1=16.723*CONC1001**2
F-             RATE2=1.5665*CONC1031*CONC62
F-             RATE3=4.5747*CONC1001*CONC1012
       R 1 = (1*1001 + 1*1022 = 1*2031 + 1*62),
       R 3 = (1*1001 + 1*1012 = 1*1031 + 1*62),
       R 2 = (1*1031 + 1*62 = 1*1001 + 1*1012),
GENeral
       COMponents = 1001,1022,2031,62,1012,1031
       FLOw 1 = 10,49.7,2*10,40,10
       TP (C) 1 = 100,50
       FLOw 3 = FLOw 1,TP 3 = TP 1
       FLOw 5 = FLOw 1,TP 5 = TP 1
       PENK,PENH
       HEAt FORmation (KCAL/GMOL) 1001 =     -103.93
       HEAt FORmation (KCAL/GMOL) 1022 =     -65.59
       HEAt FORmation (KCAL/GMOL) 2031 =     -126.1
       HEAt FORmation (KCAL/GMOL) 1012 =     -56.12
       HEAt FORmation (KCAL/GMOL) 1031 =     -105.86
END
```



Continuous Stirred Tank Reactors

Sample Output: Results from CSTR 105 are shown below.

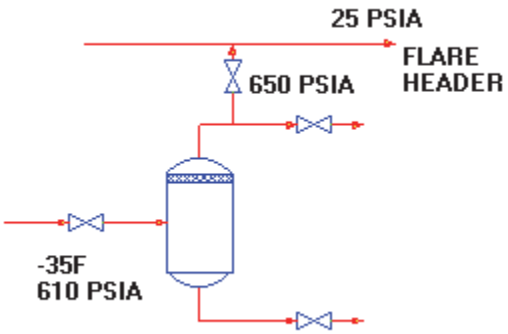
```
                    STREAM SUMMARY (CONTINUED)
                    --------------------------
STREAM NUMBER            5              6
STREAM NAME       Strm 5         Strm 6
EQUIP CONXION        FEED-LIQU      LIQU-PROD
                   (   0)-( 101)   ( 101)-(   0)
VAPOR FRACTION      0.00000E+00    0.00000E+00
TEMPERATURE F        212.00         220.00
PRESSURE PSIA        50.000         50.000
```

```
ENTHALPY BTU/HR    -0.15702E+07   -0.14836E+07
Cp  BTU/LBMOL/R      42.055         42.248
Cv  BTU/LBMOL/R      32.646         32.920
LB/FT3       T-P     48.064         47.800
S.G.(60F)    STP    0.84597        0.84573
GAL/MIN      STP     19.731         19.737
MMSCF/DAY    STP
MOLECULAR WT         64.378         64.378

FLOW RATES    LBMOL/HR

ACETIC ACID          10.000         5.0857
N-BUTANOL            49.700         47.573
N-BUTYL ACETATE      10.000         12.127
WATER                10.000         14.914
ETHANOL              40.000         37.212
ETHYL ACETATE        10.000         12.788
                    -----------    -----------
TOTAL    LBMOL/HR    129.70         129.70
TOTAL       LB/HR    8349.8         8349.8

EQUIPMENT NO.         105
EXTERNAL NAME        LIQUID RXN
TYPE OF CSTR =           DUTY SPEC.
VOLUME  FT3                 150.00
HEAT ADDED  BTU/HR       -1.50000E+05
PRES OUT    PSIA            50.000
CALC DUTY   BTU/HR       -1.50000E+05
NUMBER OF REACTIONS =   3
REACTION NO.   1
  1.00*ACETIC ACID      +  1.00*N-BUTANOL
                    ----------->
  1.00*N-BUTYL ACETATE  +  1.00*WATER
REACTION NO.   2
  1.00*ETHYL ACETATE    +  1.00*WATER
                    ----------->
  1.00*ACETIC ACID      +  1.00*ETHANOL
REACTION NO.   3
  1.00*ACETIC ACID      +  1.00*ETHANOL
                    ----------->
  1.00*ETHYL ACETATE    +  1.00*WATER

STREAM NUMBER    5     CONNECTING  (   0)  FEED
 NAME: Strm 5                   TO  ( 101)  LIQUID RXN
 P.R. K-VALUES    P.R. ENTHALPY    STD  DENSITY    FLOW RATES /HR

                      TOTAL   LIQUID    VAPOR    TOTAL    TOTAL
 COMPONENT NAME       LBMOL    LBMOL    LBMOL     LB    MOL PCT   KVALUE
 --------- ----     -------- -------- -------- -------- -------- --------
ACETIC ACID          10.0000  10.0000       0.  600.520   7.7101
N-BUTANOL            49.7000  49.7000       0.  3683.76   38.319
N-BUTYL ACETATE      10.0000  10.0000       0.  1161.61   7.7101
WATER                10.0000  10.0000       0.  180.153   7.7101
ETHANOL              40.0000  40.0000       0.  1842.72   30.840
ETHYL ACETATE        10.0000  10.0000       0.  881.040   7.7101
                    -------- -------- -------- -------- --------
TOTAL                129.700  129.700       0.  8349.81  100.000
TOTAL   LB           8349.81  8349.81       0.

STREAM NUMBER    5
TEMPERATURE DEG F      212.00
PRESSURE    PSIA       50.000
ENTHALPY    BTU/HR   -1.57018E+06
VAPOR FRACTION        0.00000E+00

PROPERTIES              TOTAL       LIQUID
----------------      ----------  ----------
FLOWRATE   LBMOL/HR     129.7000    129.7000
MOLECULAR WT.            64.3779     64.3779
ENTHALPY BTU/LBMOLE  -12106.2559 -12106.2559
ENTROPY  BTU/LBMO/R     -14.3012    -14.3012
CP       BTU/LBMOL/R                 42.0549
CV       BTU/LBMOL/R                 32.6463
DENSITY  LB/FT3                      48.0680
Z-FACTOR                           0.9292E-02
FLOWRATE GAL/MIN T-P                 21.6585
S.G.(TS/60) GPA  STP                0.845970
```

VISCOSITY CP	0.398383
TH.COND BTU/FT/HR/F	0.142756
SURFACE TENSION DYNE/CM	19.8278

Chapter 19: Depressuring Samples - c:\designii\samples\equipmnt\depres

Depr1.psd - Vertical Vessel Example with Pressure Specifications



The depressurizer module is set to relieve pressure of a process stream based on the vessel geometry and conditions of the incoming stream.

```
DEPressure vessel 1 = V101, 1, - 2
        VERtical
        CROss sectional area(IN2) = 2.7648
        SET pressure = 650
        BACk pressure = 25
        DIAmeter(FT) = 13
        REDuced pressure = 200
GENeral
        COMponents = 46, 2, 3, 4
        FLOw(FRAction) 1 =.05, .4, .325, .225
        TOTal FLOw 1 = 1749.625
        TP 1 = -35, 610
        APISOAVEH, APISOAVEK
END
```

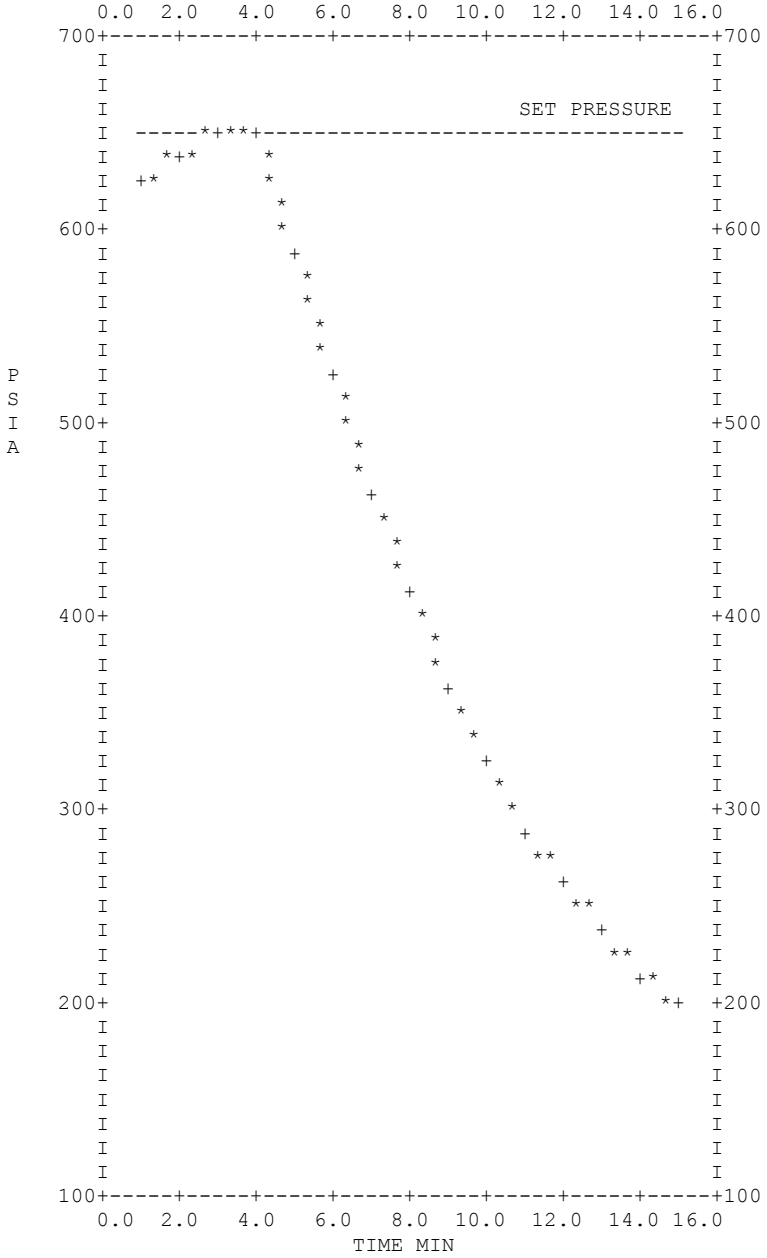
Results from this sample input are shown below and on the following pages.

```
VERTICAL    VESSEL FOR DEPRESSURING SYSTEM,
                   RADIUS =       6.5000     FT
                   LENGTH =       28.933     FT
                   LIQUID HEIGHT =  10.800    FT
                   VAPOR PHASE VENTING

INITIAL CONDITION: TEMPERATURE  = -35.000    F
                   PRESSURE  =     610.00    PSIA
                   SET PRESSURE =  650.00    PSIA
                   BACK PRESSURE=  25.000    PSIA
                   WETTED AREA FACTOR =    34500. BTU/HR/FT2
                   BACK PRESSURE FACTOR =   1.000
                   DISCHARGE COEFFICIENT =  1.000
                   RUPTURE DISK FACTOR =    1.000
                   BALANCED BEL. FACTOR =   1.000
```

INITIAL	VAPOR		LIQUID	
COMPOSITION	LBMOL	LB	LBMOL	LB
NITROGEN	60.228	1687.2	27.206	762.14
METHANE	288.10	4622.0	413.08	6627.1
ETHANE	57.884	1740.4	513.39	15437.
PROPANE	11.254	496.23	384.50	16954.
TOTAL	417.47	8545.9	1338.2	39780.

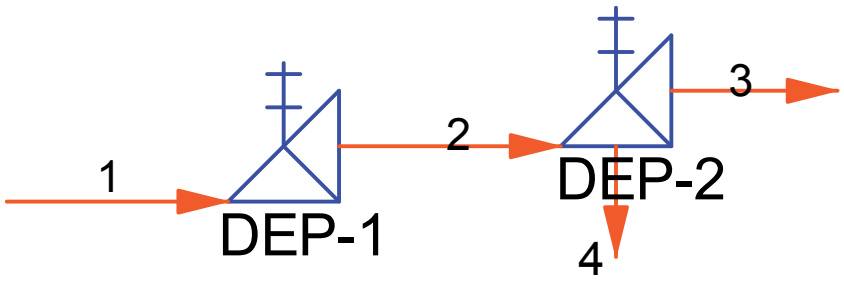
```
                     DEPRESSURING ANALYSIS TABLE
                     ------------ -------- -----
```

TIME MIN	TEMPERATURE DEG F	PRESSURE PSIA	VAPOR LB	LIQUID LB / TIME INTERVAL	VENT GAS LB	TEMP DEG F
1.00	-32.27	621.23	8709.	3.9616E+04	0.0000E+00	0.0000E+00
2.00	-29.56	632.51	8875.	3.9451E+04	0.0000E+00	0.0000E+00
3.00	-26.85	643.83	9043.	3.9282E+04	0.0000E+00	0.0000E+00
4.00	-24.16	655.20	9216.	3.9110E+04	0.0000E+00	0.0000E+00
5.00	-26.60	583.27	8243.	3.7418E+04	2665.	-95.78
6.00	-28.88	518.96	7340.	3.5956E+04	2365.	-90.51
7.00	-31.12	460.06	6585.	3.4651E+04	2060.	-85.94
8.00	-33.11	408.74	5923.	3.3513E+04	1800.	-82.08
9.00	-34.86	364.20	5344.	3.2511E+04	1581.	-78.78
10.00	-36.36	325.56	4839.	3.1618E+04	1398.	-75.93
11.00	-37.63	292.02	4399.	3.0815E+04	1243.	-73.44
12.00	-38.68	262.88	4017.	3.0086E+04	1111.	-71.23
13.00	-39.54	237.56	3686.	2.9419E+04	998.5	-69.25
14.00	-40.21	215.53	3399.	2.8804E+04	902.6	-67.45
15.00	-40.71	196.38	3150.	2.8232E+04	820.1	-65.80



Depr2.psd - Spherical Vessel With Specified Initial Volume

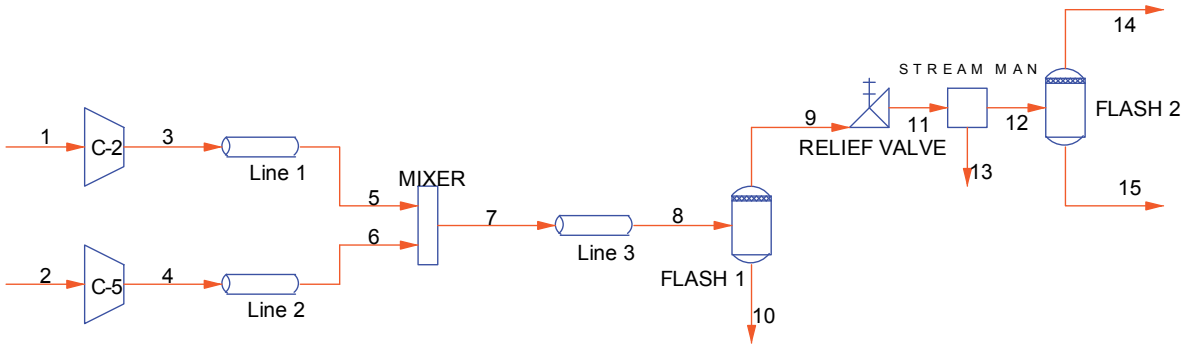
Two depressuring vessels are used with specified vessel DIAmeter, LIQuid HEIght, and LENgth.




```
AB123.
*DEPRESSURING VESSELS
DEPressure 1 = DEP1, 1, -2
        CROss sectional area (IN2) = 2.7648
        SET PREssure = 650
        BACk PREssure = 25
        DIAmeter =     2.5
        LIQuid HEIght = 9.5
        LENgth = 16
        REDuced pressure = 200
DEPressure 2 = DEP2, 2, -12, -13
        PRInt = 3
        TIMe STEp (MIN) = 2.0
        SPHere
        LIQuid HEIght = 6.5
        HEAt = 0
        CROss sectional area (IN2) = 2.7648
        SET PREssure = 600
        DIAmeter (M) = 3.9624
GENeral
        COMponents = 46, 2, 3, 4
        FLOw (FRA) 1 = .05, .4, .325, .225
        TOTal FLOw 1 = 1749.645
        TP 1 = -35, 610
        APISOAVEH, APISOAVEK, APISOAVED
END
```

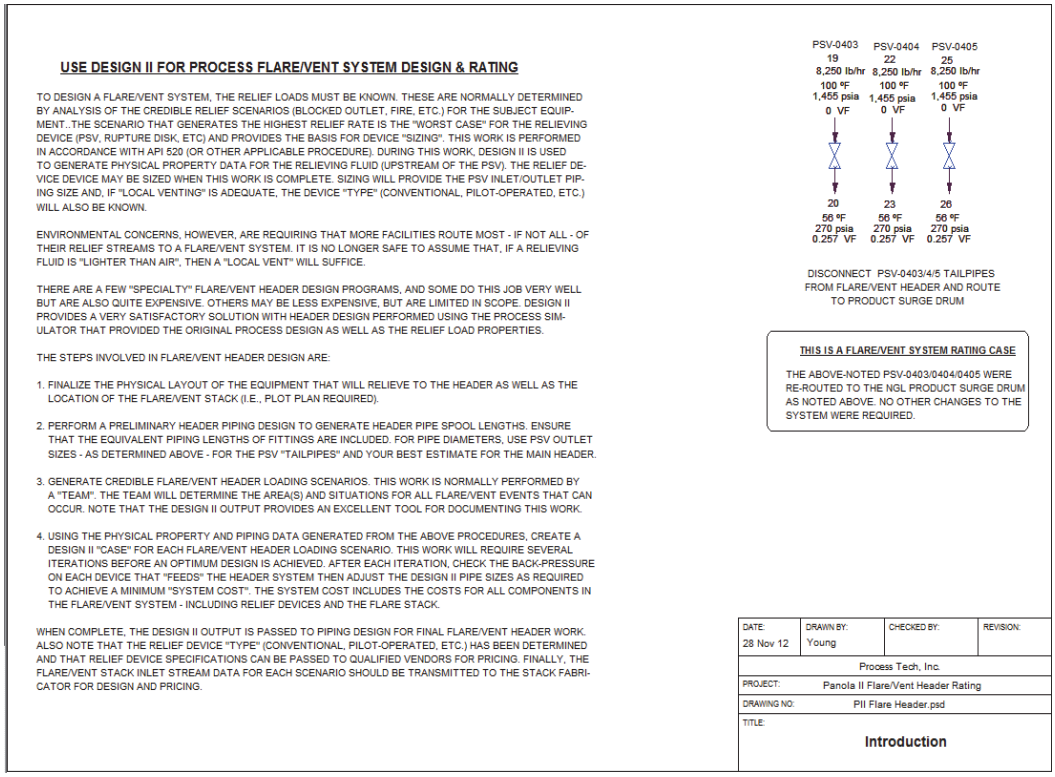
Flrehdr.psd - Flare Header Simulation

The sample simulation shows two different lines being compressed for a typical flare header relief valve sizing calculation. Methods such as API 520 and API521 relief valve sizing can be performed via the depressurizer module. The depressurizer module produces the vapor and liquid compositions relieved from the set pressure to a pre-defined reduced pressure. Ambient pressure is the default for the back pressure.



PII FIrehdr.psd - Flare/Vent System Design & Rating

Use DESIGN II for Process Flare/Vent System design rating.



USE DESIGN II FOR PROCESS FLARE/VENT SYSTEM DESIGN & RATING

TO DESIGN A FLARE/VENT SYSTEM, THE RELIEF LOADS MUST BE KNOWN. THESE ARE NORMALLY DETERMINED BY ANALYSIS OF THE CREDIBLE RELIEF SCENARIOS (BLOCKED OUTLET, FIRE, ETC.) FOR THE SUBJECT EQUIPMENT..THE SCENARIO THAT GENERATES THE HIGHEST RELIEF RATE IS THE "WORST CASE" FOR THE RELIEVING DEVICE (PSV, RUPTURE DISK, ETC) AND PROVIDES THE BASIS FOR DEVICE "SIZING". THIS WORK IS PERFORMED IN ACCORDANCE WITH API 520 (OR OTHER APPLICABLE PROCEDURE). DURING THIS WORK, DESIGN II IS USED TO GENERATE PHYSICAL PROPERTY DATA FOR THE RELIEVING FLUID (UPSTREAM OF THE PSV). THE RELIEF DEVICE DEVICE MAY BE SIZED WHEN THIS WORK IS COMPLETE. SIZING WILL PROVIDE THE PSV INLET/OUTLET PIPING SIZE AND, IF "LOCAL VENTING" IS ADEQUATE, THE DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) WILL ALSO BE KNOWN.

ENVIRONMENTAL CONCERNS, HOWEVER, ARE REQUIRING THAT MORE FACILITIES ROUTE MOST - IF NOT ALL - OF THEIR RELIEF STREAMS TO A FLARE/VENT SYSTEM. IT IS NO LONGER SAFE TO ASSUME THAT, IF A RELIEVING FLUID IS "LIGHTER THAN AIR", THEN A "LOCAL VENT" WILL SUFFICE.

THERE ARE A FEW "SPECIALTY" FLARE/VENT HEADER DESIGN PROGRAMS, AND SOME DO THIS JOB VERY WELL BUT ARE ALSO QUITE EXPENSIVE. OTHERS MAY BE LESS EXPENSIVE, BUT ARE LIMITED IN SCOPE. DESIGN II PROVIDES A VERY SATISFACTORY SOLUTION WITH HEADER DESIGN PERFORMED USING THE PROCESS SIMULATOR THAT PROVIDED THE ORIGINAL PROCESS DESIGN AS WELL AS THE RELIEF LOAD PROPERTIES.

THE STEPS INVOLVED IN FLARE/VENT HEADER DESIGN ARE:

1. FINALIZE THE PHYSICAL LAYOUT OF THE EQUIPMENT THAT WILL RELIEVE TO THE HEADER AS WELL AS THE LOCATION OF THE FLARE/VENT STACK (I.E., PLOT PLAN REQUIRED).
2. PERFORM A PRELIMINARY HEADER PIPING DESIGN TO GENERATE HEADER PIPE SPOOL LENGTHS. ENSURE THAT THE EQUIVALENT PIPING LENGTHS OF FITTINGS ARE INCLUDED. FOR PIPE DIAMETERS, USE PSV OUTLET SIZES - AS DETERMINED ABOVE - FOR THE PSV "TAILPIPES" AND YOUR BEST ESTIMATE FOR THE MAIN HEADER.
3. GENERATE CREDIBLE FLARE/VENT HEADER LOADING SCENARIOS. THIS WORK IS NORMALLY PERFORMED BY A "TEAM". THE TEAM WILL DETERMINE THE AREA(S) AND SITUATIONS FOR ALL FLARE/VENT EVENTS THAT CAN OCCUR. NOTE THAT THE DESIGN II OUTPUT PROVIDES AN EXCELLENT TOOL FOR DOCUMENTING THIS WORK.
4. USING THE PHYSICAL PROPERTY AND PIPING DATA GENERATED FROM THE ABOVE PROCEDURES, CREATE A DESIGN II "CASE" FOR EACH FLARE/VENT HEADER LOADING SCENARIO. THIS WORK WILL REQUIRE SEVERAL ITERATIONS BEFORE AN OPTIMUM DESIGN IS ACHIEVED. AFTER EACH ITERATION, CHECK THE BACK-PRESSURE ON EACH DEVICE THAT "FEEDS" THE HEADER SYSTEM THEN ADJUST THE DESIGN II PIPE SIZES AS REQUIRED TO ACHIEVE A MINIMUM "SYSTEM COST". THE SYSTEM COST INCLUDES THE COSTS FOR ALL COMPONENTS IN THE FLARE/VENT SYSTEM - INCLUDING RELIEF DEVICES AND THE FLARE STACK.

WHEN COMPLETE, THE DESIGN II OUTPUT IS PASSED TO PIPING DESIGN FOR FINAL FLARE/VENT HEADER WORK. ALSO NOTE THAT THE RELIEF DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) HAS BEEN DETERMINED AND THAT RELIEF DEVICE SPECIFICATIONS CAN BE PASSED TO QUALIFIED VENDORS FOR PRICING. FINALLY, THE FLARE/VENT STACK INLET STREAM DATA FOR EACH SCENARIO SHOULD BE TRANSMITTED TO THE STACK FABRICATOR FOR DESIGN AND PRICING.

PSV-0403 19 8,250 lb/hr 100 °F 1,455 psia 0 VF → 20 56 °F 270 psia 0.257 VF

PSV-0404 22 8,250 lb/hr 100 °F 1,455 psia 0 VF → 23 56 °F 270 psia 0.257 VF

PSV-0405 25 8,250 lb/hr 100 °F 1,455 psia 0 VF → 26 56 °F 270 psia 0.257 VF

DISCONNECT PSV-0403/4/5 TAILPIPES FROM FLARE/VENT HEADER AND ROUTE TO PRODUCT SURGE DRUM

THIS IS A FLARE/VENT SYSTEM RATING CASE

THE ABOVE-NOTED PSV-0403/0404/0405 WERE RE-ROUTED TO THE NGL PRODUCT SURGE DRUM AS NOTED ABOVE. NO OTHER CHANGES TO THE SYSTEM WERE REQUIRED.

DATE:	DRAWN BY:	CHECKED BY:	REVISION:
28 Nov 12	Young		

Process Tech, Inc.

PROJECT: Panola II Flare/Vent Header Rating

DRAWING NO: PII Flare Header.psd

TITLE: **Introduction**

```
AB123.
*Crude Maker -

C- EQUIPMENT MODULES

FLA 1 = F-1, 1,-2,-3,-4,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0
  VERT
  MIS ELI
  SUR (MIN) = 5

C- CRUDE FEED CHARACTERIZATION
CRUDE
  FEE BLEND  1 = 1
  FEE VOL  1 = 0,5,10,20,30,40,50,60,70,80,100,
  FEE TBP (F) 1 = 50,100,175,200,250,300,350,400,450,500,550,
  FEE (BBL/DAY) 1 = 600
  FEE WAT (BBL/DAY) 1 = 50
  FEE GRA (API) 1 = 22.5
  FEE MOL VOL  1 = 0,5,10,20,30,40,50,60,70,80,90,100,
  FEE NAME  1 = Crude TBP Feed
  FEE REA () 1 = 0.12,2,1.9,
  FEE VIS TEM(F) = 100,210,
  FEE BUL VIS SET 1 (CST) STR 1 = 86
  FEE VIS VOL SET 1  STR 1 = 0,32,38,50,60,70,75,78,100,
  FEE VIS SET 1 (CST) STR 1 = 0,3.8,6,16,52,216,1000,2600,39000,
  FEE BUL VIS SET 2 (CST) STR 1 = 7.8
  FEE VIS VOL SET 2  STR 1 = 0,32,38,50,60,70,75,78,100,
  FEE VIS SET 2 (CST) STR 1 = 0,1.4,1.8,3.2,5.9,14,30,50,245,
  FEE BULK PRO SET 1  STR 1 = 0.94
  FEE PRO VOL SET 1  STR 1 = 0,5,6.5,10,21,30,40,50,61,71,81.7,100,
  FEE PRO SET 1 (WGT) STR 1 = Sulfur,0,2*0.002,0.004,0.032,0.16,0.46,0.81,0.94,
           1.11,1.57,2.38,
  FEE BULK PRO SET 2  STR 1 = 80.6
  FEE PRO VOL SET 2  STR 1 = 0,10,30,55,85,100,
```

```
  FEE PRO SET 2 (PPT) STR 1 = Pour Point,-68.8,-56.2,-40,89.6,104,111.2,


C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,2,3,4
  NAM POS 1 = 'WATER'
  NAM POS 2 = 'METHANE'
  NAM POS 3 = 'ETHANE'
  NAM POS 4 = 'PROPANE'
C-  Thermo Comments
  MOD ESSO
  APIMOD
  STDD
  DEN LIQ = STDD
  VIS VAP = MODAPI
  VIS LIQ = LNAV
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  VIS UNI OUT = CST
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  REI PRO = 4,
  TP (F,PSIG) 1 = 80,1000
  NAM STR 1 = 'Crude TBP Feed'
  NAM STR 2 = 'Strm 2'
  NAM STR 3 = 'Strm 3'
  NAM STR 4 = 'Strm 4'
END
```

Chapter 20: Distillation Column Samples - c:\designii\samples\equipmnt\distill

Acwtrtcl.dat – ChemTran Data File for Runtime of DISTL5.PSD

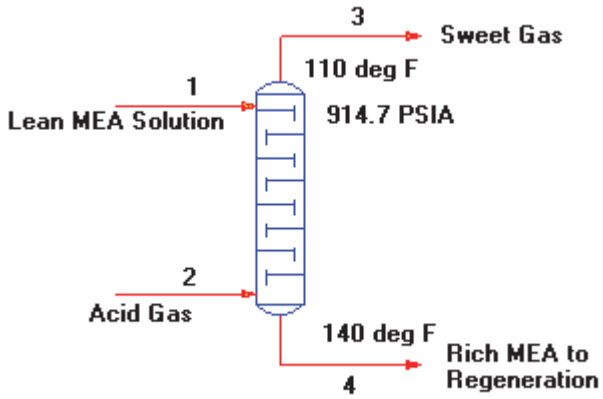
It is a ChemTran created binary data file created by the ChemTran input file DISTL5CT.IN for use by DESIGN II at runtime of DISTL5.PSD.

Distl1.psd - Absorber- MEA Treating

This is a non-ideal column calculation which involves chemical reactions between the monoethanolamine, CO2, and H2S. The heat of reaction is taken into account in the calculation.

Preliminary steps include:

1. Determine which amine to use - monoethanolamine or diethanolamine.
2. Determine the concentration of MEA solution to use (typically between 10 and 35 weight-percent).
3. Determine MEA circulation rate. Upper loading for rich amine is normally 0.6 moles acid gas/mole MEA (the MDEAK technique can handle much higher loadings).



Feeds	1	2
Temperature, F	110	90
Pressure, PSIA	914.7	914.7
Flowrate,	LBMOL/HR	
CH4	0	5351.87
CO2	62.4	125.76
H2S	1.21	31.32
H2O	9440.67	0.0
MEA	489.6	0.0

Guesses:Temperature of Top Product = 110 F, Temperature of Bottoms Product = 140 F, Bottoms Product = 10130 LBMOL/HR

No specifications are required for ABSorber.

```
AB123.
* MEA CONTACTOR
DIStillation 1 = CONT, 1, 2, -3, -4
       ABSorber
       TRAys = 7,
       LOCation of FEEds = 1, 8
       PREssure of TOP = 914.7
C- INITIAL GUESSES
       TEMperature of TOP GUEss = 110,
       TEMperature of BOTtom GUEss = 140
       PROduct rates = 0, 10130
GENeral
       COMponents = 2, 49, 50, 62, 4155
       FLOwrates 1 = 0, 62.4, 1.21, 9440.67, 489.6,
       TP 1 = 110,914.7
       FLOwrates 2 =5251.87,125.76, 31.32, 0, 0,
       TP 2 = 90,914.7
C-
C- THERMODYNAMICS CHOICES
C-   K-VALUES   ENTHALPIES
       MDEAK, MDEAH
END
```

Distl2.psd - Lean Oil Absorber with Presaturator

This example illustrates the simulation of a Lean Oil Absorber with a Presaturator.

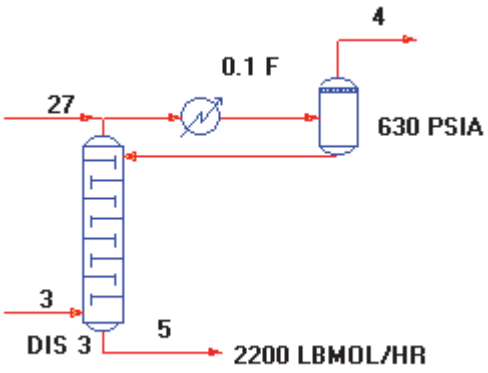
Feed Streams	27	3
Temperature, F	0	0

Pressure, PSIA	650	635
Flowrate		LBMOL/HR,
N2	0	30
CO2	0	50
C1	0	10100
C2	0	500
C3	0	150
iC4	0	40
nC4	0	30
iC5	0	15
nC5	0	10
C6	0	20
C7	0	5
Lean Oil	1200	0

(Lean Oil: API = 50, mean average boiling point = 325 F, Molecular Weight = 135)

Guesses: Bottoms product rate = 2200 LBMOL/HR, Temperature of top tray = 5 F
Temperature of bottom product = 9 F

```
AB123.
* LEAN OIL ABSORBER WITH PRESATURATOR
DIStillation 3 = LOAB, 27, 3, -4, -5
        ABSorber
        PREsaturator,
        TEMperature of CHIller = 0
        TRAys = 5,
        LOCation of FEEds = 0, 6
        PREssure of TOP = 635
        PREssure of CONdenser = 630
C- INITIAL GUESSES
        TEMperature of TOP = 5,
        TEMperature of BOTtom = 9
        PROduct rates = 0, 2200
        PRInt SWItch = 2
GENeral
        COMponents =46,49,2,3,4,5,6,7,8,10,11,100
C- DEFINE ATTRIBUTES OF COMPONENT 100
        NAMe 100 = LEAN OIL,
        API 100 = 50,
        AMB 100 = 325,
        AMW 100 = 135
        TP 27 = 0, 650,
        TP 3 = 0, 635
        FLOwrates 27 = 11 * 0 , 1200
        FLOwrates 3 = 30,50,10100,500,150,40,30,15,10,20,5,0
C- THERMODYNAMIC K-VALUE AND ENTHALPY OPTIONS
        SOAVEK, SOAVEH
END
```

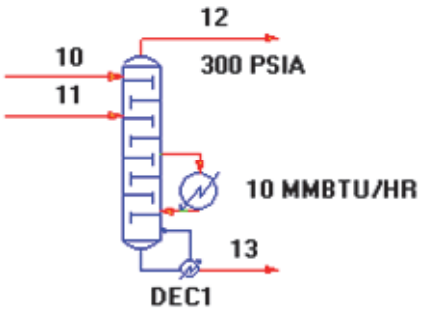


Distl3.psd - Cryogenic Demethanizer with Side Heater

A cryogenic demethanizer column is simulated using the distillation module. The distillation is also set to handle a side heater and the column is set to achieve 0.025 molar ratio of methane to propane in the bottoms.

Feeds	Stream 10	Stream 11
Temperature, F	-142.6	-140.4
Pressure,PSIA	300	300
Flowrate,LBMOL/HR		
N2	400	116
CO2	440	354
C1	23070	12295
C2	894	1154
C3	194	416
i-C4	21	67
n-C4	28	108
i-C5	4.1	32
n-C5	3.8	38
C6	6.1	121
C7+	0.05	40

(C7+ : API = 74.1, Mean Average Boiling Point = 209 F, Molecular Weight = 100)



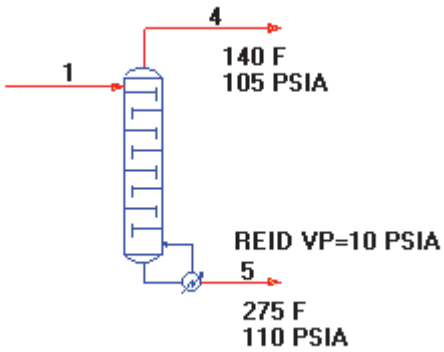
Guesses: Bottoms Product Rate = 2700 LBMOL/HR, Temperature of Top Product = -125 F
Temperature of Bottom Product = 50 F
Specification: .025 C1/C2 ratio in bottoms

```
AB123.
* CRYOGENIC DEMETHANIZER
C-
DIStillation 6 = DEC1, 10, 11, -12, -13
       ABSorber with REBoiler
       TRAys = 10,
       LOCation of FEEds = 1, 3
       PREssure of TOP tray = 300
C-
C- SPECIFICATION
       RATio in BOTtoms 2, 3 = 0.025
C-
C- INITIAL GUESSES
       TEMperature of TOP = -125,
       TEMperature of BOTtom = 50
       PROduct rates = 0, 2700
C-
       LOCation of HEAt = 7,
       HEAt = 10E6
       PRint SWItch = 2
GENeral
       COMponents = 46, 49, 2, 3, 4, 5, 6, 7, 8, 10, 100
C- DEFINE ATTRIBUTES OF COMPONENT 100
       NAMe 100 = C7 +,
       API 100 = 74.1,
       AMB 100 = 209,
       AMW 100 = 100
       TP 10 = -142.6, 300,
       TP 11 = -140.4 , 300
       FLO 10=400,440,23070,894,194,21,28,4.1,3.8,6.1,0.05
       FLO 11=116,354,12295,1154,416,67,108,32,38,121,40
C-
C- THERMODYNAMIC K-VALUE CHOICE
       APISOAVEK
END
```

Distl4.psd - Inline Fortran to meet Reid Vapor Pressure specification.

This example creates a Reid vapor pressure specification for the DIStillation column using Inline FORTRAN. It calls the Design II subroutine REIDVP to calculate the Reid vapor pressure of the bottom product. Note the use of the FLOWL (x,y) utility to retrieve component flow rates for liquid leaving the bottom stage of the column. A PURity Bottom specification is entered but will be overwritten with the Inline FORTRAN EPS (1) specification. A Reid vapor pressure of 10 PSI is desired for the bottom product. Note that a Reid vapor pressure command has been added to the GENeral section for the bottom product (Stream 5).

Feed	Stream 1
Temperature, F	85
Pressure, PSIA	310
Flowrate, LBMOL/HR	
Nitrogen	2.05
CO2	30.52
Methane	8.55
Ethane	17.7
Propane	48.75
i-Butane	15.015
n-Butane	64.35
i-Pentane	21.54
n-Pentane	26.4
n-Hexane	32.97
n-Heptane	20.11
n-Octane	6.81
n-Nonane	1.44
n-Decane	1.905
n-Undecane	11.7




```
AB123.
* SAMPLE INLINE FORTRAN PROBLEM
C-
C-  CREATE REID VAPOR PRESSURE SPECIFICATION FOR THE
```

```
C-  DISTILLATION COLUMN USING INLINE FORTRAN. CALL
C-  DESIGN II'S SUBROUTINE REIDVP TO CALCULATE VALUE
C-  FOR REID VAPOR PRESSURE OF THE BOTTOMS PRODUCT.
C-  USE FLOWL(X,Y) UTILITY TO GET COMPONENT FLOWRATES
C-  FOR LIQUID LEAVING THE BOTTOM STAGE OF THE COLUMN. C-
DIStillation 1 = STABILIZER, 1, -4, -5
        ABSorber with REBoiler,
        TRAys = 12,
        LOCation of FEEd = 1
        PREssure of TOP = 105,
        DELta pressure = 5
        PROduct rates = 0, 116
C-
C- INITIAL GUESSES
               TEMperature of TOP = 140,
               TEMperature of BOTtom = 275
C- SUPER CONVERGENCE TECHNIQUE IS REQUIRED FOR INLINE
C- FORTRAN SPEC SUPer
C-
C-  PURITY SPEC WILL BE REPLACED BY INLINE FORTRAN EPS(1) SPEC
C-  FOR REID VAPOR PRESSURE -- BOTTOM PRODUCT RVP = 10
C-  EPS(1) SPECIFICATION APPROACHES ZERO WHEN SPEC IS MET
C-
               PURity BOTtom 2 = 0.015
C-
F-      DURING
F-      CALL REIDVP(FLOWL(1,1),RVP,IC,ERR)
F-      EPS(1)=(RVP - 10.)/10.
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,11,12,13,14,15
        FLOw 1 = 2.05, 30.52, 8.55, 17.7, 48.75, 15.015,
        64.35, 21.54, 26.4,32.97, 20.11, 6.81, 1.44, 1.905,
        11.7
        TP 1 = 120, 115
C- THERMODYNAMICS CHOICES
C-  K-VALUES    ENTHALPIES    DENSITIES
   APISOAVEK,   APISOAVEH,        STDD   REID = 5
END
```

Distl5.psd - Liquid-Liquid Extraction

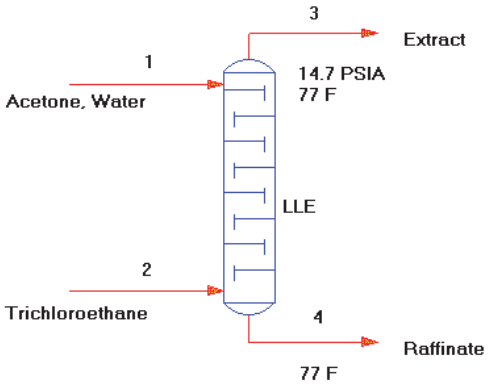
This example illustrates a liquid-liquid extraction using the ABS column type with the LLE option. In this problem, an acetone-water mixture is contacted counter-currently with trichloroethane.

Guesses: Bottoms Product = 2.6 LBMOL/HR

Feeds	Stream 1	Stream 2
Temperature, F	77	77
Pressure, PSIA	14.7	14.7
Flowrate, LB/HR		
C2H3CL3	0	30
C3H6O	50	0
H2O	50	0

No specifications are required for ABSorber.

```
AB123.
*Extraction column
DIStillation = LLE, 1, 2, -3, -4,
        LLE
        PREssure TOP = 14.7,
        TEMperature TOP = 77,
        TEMperature BOTtom = 77,
        TRAys = 6,
        LOCation of FEEds = 1, 6
        PROduct rates = 0, 2.6
GENeral
        COMponents = 3050, 1002, 62,
        FLOwrate (LB/HR) 1 = 0, 50, 50,
        FLOwrate (LB/HR) 2 = 30, 0, 0,
        TP 1 = 77, 14.7
        TP 2 = 77, 14.7
        CHEmtran FILe = LLE,UNQC,DATA
        UNIQUACK, LAT
END
```



Distl6.psd - Sour Water Stripper

This example illustrates the simulation of Sour water strippers using the Beychok method. NH3, H2S, and H2O interaction date is provided automatically. Because of the high degree of compositional dependence of the K-values for H2S, NH3, H2O systems the following steps are recommended.

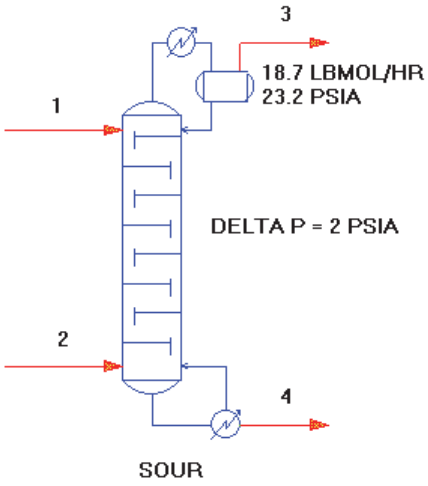
1. Use the Steam Tables* and an educated guess on NH3 and H2S effects to make an accurate guess for the TEM TOP and TEM BOT. If you do not enter a TEMperature PROfile, enter the value for the top tray as your guess for TEMperature of TOP rather than the product temperature. The column profile if fairly linear, except between trays 0 and 1. A change in the temperature guess of 5 F may be the difference between convergence and nonconvergence.
2. Use the Steam Tables* with the assumption that feeds are all water or steam, to estimate a VAPor PROfile.
3. Use COM CON = 10,2 (COM CON 12,4 and DVS = 0.4 with SUPER) to insure stability in K-values.
4. Tighter overall column balances can be obtained by changing the TOLerance to 1E-6 (Default is 1E-5) with negligible effect in the costs.
5. Use SOUR for K-values and YEN-Alexander heats for enthalpies. SOUR uses the Beychok technique for H2O, H2S and NH3 K-values. For other components Chao-Seader K-values are used unless overridden by entering tabular K-values via ChemTran.

* A preliminary run with VAPor PREssure K-values and LATent heat enthalpies can be used to generate this information.

Feeds	Stream 1	Stream 2
Temperature, F	180	309
Pressure, PSIA	23.2	24.7
Flowrate, LB/HR		
H2O	74500	11800
NH3	149.6	0
H2S	89.6	0

Guesses:

Reflux ratio = 32.7
Temp top product = 199 F
Temp bottoms product = 240 F
Vapor profile = 411, 695, 5*0, 725 Specification:
Top product rate = 18.7 LBMOL/HR



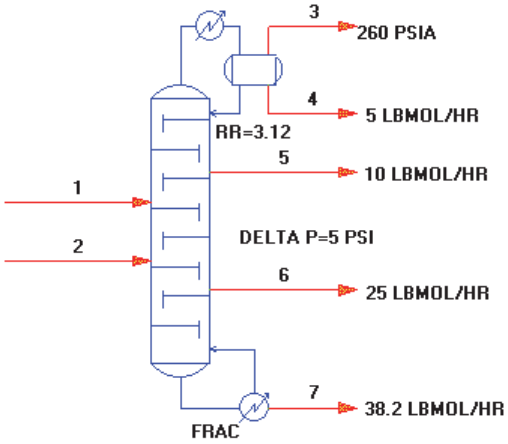

```
AB123.
* SOUR WATER STRIPPER
C-
DIStillation 1 = SOUR , 1, 2, -3, -4
        STRipper
        TRAys = 8,
        LOCation of FEEds = 1, 9
        PREssure of TOP = 23.2,
        DELta pressure = 2
        TEMperature of TOP  = 199,
        TEMperature of BOTtom  = 240
C-
C- SPECIFY TOP PRODUCT RATE
        PROduct rates = 18.7, 0
        REFlux GUEss = 32.7
        VAPor PROfile = 411, 695, 5 * 0, 725
        COMpositional CONvergence = 10, 2,
        TOLerance = 1E-6
        PRInt SWItch = 2
GENeral
        COMponents = 62, 63, 50
        TP 1 = 180, 23.2,
        TP 2 = 309, 24.7
        FLOwrates (LB/HR) 1 = 74500, 149.6, 89.6
        FLOwrates (LB/HR) 2 = 11800, 0, 0
C-
C- FORCE STREAM 2 TO BE ALL VAPOR
        VAPor = 2
C-
C- THERMODYNAMIC CHOICES
C-   K-VALUES    ENTHALPIES
        SOUR , YEN
C-
C- SET OUPUT UNITS OF MASS FLOWS TO POUNDS
        QUAntity UNIts OUT = LB
END
```

Distl7.psd - Partial Condenser with Three Side Draw Products

This example illustrates the simulation of a partial condenser with three liquid side draw products.

Guesses:
Top product temperature = 95F
Bottoms product temperature = 340 F Specifications:
Reflux Ratio = 3.12
Bottoms Product Rate = 38.2 LBMOL/HR

Feeds	Stream 1	Stream 2
Temperature, F	130	220
Pressure, PSIA	270	270
Component Flowrates, LBMOL/HR		
Methane	2	
Ethane	10	
Acetylene	6	1
Propane	12	7
Isobutane	1	4
Normal Butane	3	17
Normal Pentane	5	15.2
Normal Hexane		9
Normal Heptane		4.5
Normal Octane		4.3
Normal Nonane		3.5



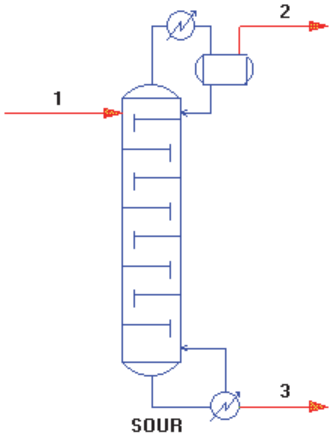

```
AB123.
* CONDENSATE FRACTIONATOR
DIStillation 1 = FRAC, 1, 2, -3, -4, -5, -6, -7
        PARtial CONdenser
        TRAys = 17,
        LOCation of FEEds = 7, 12
        PREssure of TOP = 260,
        DELta pressure = 5
        TEMperature of TOP = 95,
        TEMperature of BOTtom = 340
C- 2 SPECIFICATIONS PLUS SIDEDRAW RATES
        REFlux ratio = 3.12
        PROduct rates = 0, 5, 10, 25, 38.2
        LOCation of SIDedraws = 0, 3, 15
        PRInt SWItch = 2
GENeral
COMponents = 2, 3, 23, 4, 5, 6, 8, 10, 11, 12, 13
        TP 1 = 130, 270,
        TP 2 = 220, 270
        FLOwrates 1 = 2, 10, 6, 12, 1, 3, 0.5
        FLO 2 = 0, 0, 1, 7, 4, 17, 15.2, 9, 4.5, 4.3, 3.5
C- THERMODYNAMICS CHOICES
PENK, BINary PARameters = PENG1
END
```

Distl8.psd - Partial Condenser-Sour Water Stripper with CO2

This example has to consider the solubilities and vapor phase interactions of H2S, NH3 and CO2 in water. The effect of CO2 on the pH of the system is also taken into account.

Preliminary steps:

1. Use COM CON = 30,3 with regular convergence technique or COM CON = 12,4 and DVS = 0.4 with SUPer to insure stability of K-values.
2. Use APISOUR K-values and YEN-Alexander enthalpies. The APISOUR technique also handles the solubilities of H2, N2, methane, ethane, propane, isobutane, and normal butane in water. K-values for heavier hydrocarbon are calculated using the Chao-Seader correlation.
3. If you do not enter a TEMperature PROfile, use the top tray temperature rather than the condenser temperature as an initial guess.
4. The APISOUR technique can also model use of caustics or organic acids



to control pH. See Thermodynamics Section, for details.

Feed	Stream 1
Temperature, F	230
Pressure, PSIA	34.7
Flowrate, LB/HR	
NH3	155
H2S	234
CO2	40
H2O	49,426
CH4	800
C2H6	100

Guesses: Temperature of Top = 128 F
Temperature of Bottom = 239 F
Vapor Profile = 75, 825, 8 * 0, 708

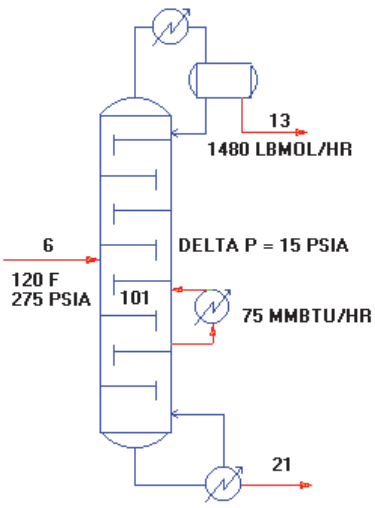
Specifications: Vapor Product rate = 75.0 LBMOL/HR, Reflux = 10.0

NOTE: If you are entering LIQuid COMposition profiles for CO2 or NH3, you do not need COM CON = 12,4 or SUPer.

```
AB123.
* SOUR WATER COLUMN
C-
DIStillation 1 = SOUR, 1, -2, -3
        PARtial CONdenser
        TRAys = 9,
        LOCation of FEEd = 3,
        PREssure of TOP = 180,
        DELta pressure = 3
        TEMperature of TOP  = 128,
        TEMperature of BOTtom  = 239
C-
C- SPECIFICATIONS
        PROduct rates = 75, 0,
        REFlux = 10.0
C-
C- OPTIONAL INPUT
        PREssure of CONdenser = 26.7
C-
C- INITIAL GUESSES
        VAPor PROfile = 825, 7 * 0, 722
        REBoiler VAPor GUEss = 708
C-
C- CONVERGENCE CONTROL
        SUPer, COMpositional CONvergence = 12, 4, DVS = 0.4
GENeral
        COMponents = 63, 50, 49, 62, 2, 3
        FLOw rates (LB/HR) 1 = 155, 234, 40, 49426, 800, 100
        TP 1 = 230, 34.7
C-
C- THERMODYNAMICS OPTIONS
C-  K-VALUES     ENTHALPY
            APISOUR,       YEN
END
```

Distl9.psd - Total Condenser-Split Propylene from Propane

This example illustrates the separation of propane and propylene in a distillation column.



Feed

Temperature, F	120
Pressure, PSIA	275
Component flowrates, LB/HR	
propane	7724
propylene	62304

Guesses: Top product temperature = 112.5 F
Bottom Product temperature = 131 F

Specifications: Reflux = 22.5
Purity of propylene in overhead product = 0.996

```
AB123.
* C3 SPLITTER
```

```
DIStillation 12 = C3TW, 6, -13, -21
        TOTal
        TRAys = 155,
        LOCation of FEEd = 100
        PREssure of TOP = 269,
        DELta pressure = 10
        TEMperature of TOP GUEss = 112.5,
        TEMperature of BOTtom GUEss = 131
        PROduct rates = 1480, 0
C-
C- SPECIFICATIONS
        REFlux ratio = 22.5,
        PURity of TOP 23 = 0.996
C-
C- SIDE HEAT EXCHANGER
        LOCation of HEAt = 101 , HEAT = 75E6
GENeral
        COMponents = 4, 23
        FLOwrates (LB/HR) 6 = 7724, 62304,
        TP 6 = 120, 275
C-
C- THERMODYNAMICS OPTIONS
C-   K-VALUES     ENTHALPIES
     PENK,          PENH, BINary PARameters = PENG1
END
```

DividedWallColumn2.psd – Separation of Three Products using One Column

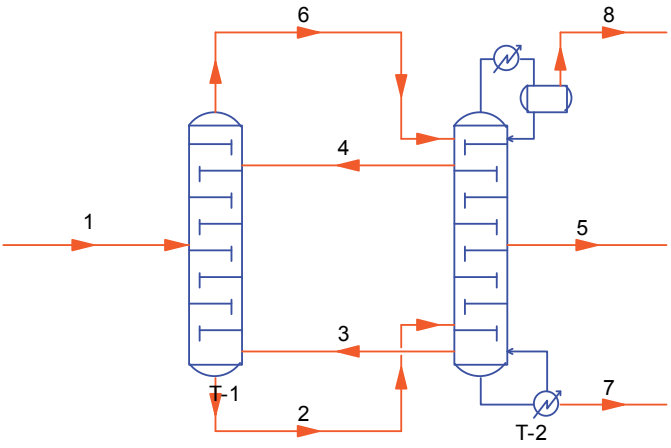
This example illustrates the separation of three products using a divided wall column. Since DESIGN II does not have a divided wall column model, the model uses a combination of an absorber column with a fractionator column.

Feed

Temperature	128 C
Pressure	5 atm
Component flowrates, lbmol/hr	
n-butane	280
n-pentane	300
n-hexane	140
n-heptane	530

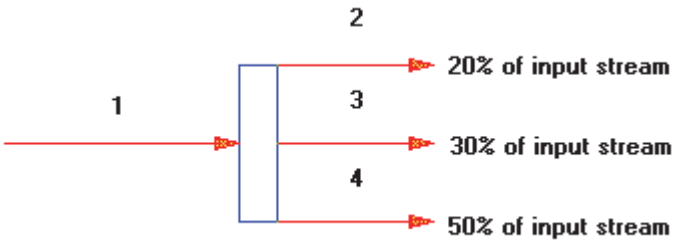
Specifications:

Condenser Temperature	200 F
Bottom Product Rate	500 lbmol/hr



Chapter 21: Divider Samples - c:\designii\samples\equipmnt\divider

Divdr1.psd - Division into Fractions



Given stream 1, divide it into three output streams equal to 20%, 30%, and 50% of stream 1.

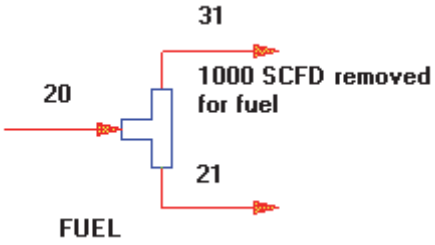
Coding

```
DIVider 1 = FRAC, 1, -2, -3, -4,
     FLOw RATe (FRA) = .2, .3, .5
```

Method of Calculation

All properties of stream 1 are known. The temperature and pressure of the feed are transferred to the output streams. The enthalpy and component flowrates are divided into the specified fractions.

Divdr2.psd - Flowrate Specification Example



Given stream 20, create a slip stream with a flow of 1000 SCF/DAY.

Coding

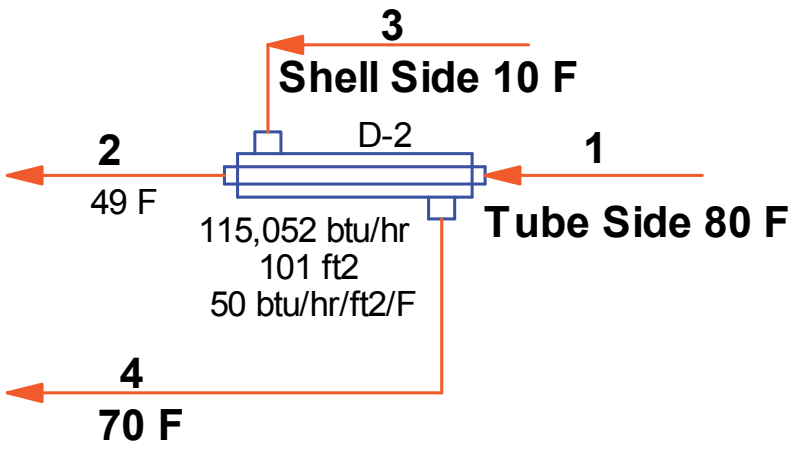
```
DIVider 14 = FUEL, 20, -31, -21,
     FLOw RATe (SCF/DAY) = 1000
```

Method of Calculation

All properties of stream 20 are known. The 1000 SCF/DAY flow is subtracted from the feed and put into the first output stream. The rest of the feed is put into the second output stream. The enthalpy and component flowrates are distributed properly and other properties are transferred from the feed.

Chapter 22: Double Pipe Exchanger Samples - c:\designii\samples\equipmnt\doublepipe

Doublepipe5.psd - Division into Fractions



Given stream 1, divide it into three output streams equal to 20%, 30%, and 50% of stream 1.

Coding

```
DIVider 1 = FRAC, 1, -2, -3, -4,
     FLOw RATe (FRA) = .2, .3, .5
```

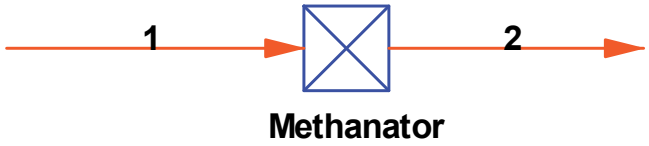
Method of Calculation

All properties of stream 1 are known. The temperature and pressure of the feed are transferred to the output streams. The enthalpy and component flowrates are divided into the specified fractions.

Chapter 23: Equilibrium Reactor Samples - c:\designii\samples\equipmnt\equireac

Equlb1.psd - Using Product Constraint

This simulation shows an equilibrium reactor to set the amount of any desired component wanted in the product (Product Constraint).




```
AB123.
*METHANATOR

EQU REA 1 = Methanator, 1,-2,
  MAX ITE = 30
  PRO (LBMOL/HR)2 = 1.514

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,49,48,1,62,46,
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 15.14,2*0,0.2,84.07,0.59,
  TP (K,PSIA) 1 = 1066.67,179.12
END
```

Equlb2.psd - Use Extent Of Reactions

A sample simulation showing an equilibrium reactor to set the extent of reaction for any defined reaction.


```
AB123.
*Extent of Reactions

EQU REA 1 = Methanator, 1,-2,
  MAX ITE = 30
  INE = 46,
  R1 = (1*2+1*62 = 1*48+3*1)
  TEM REA (F) 1 = 0
  R2 = (1*48+1*62 = 1*49+1*1)
  EXT (LBMOL/HR) 2 = 13.8

GENERAL
  COMPONENTS = 2,49,48,1,62,46,
  BWRK
  BWRH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  CON TOL = 0.001
```

```
  FLO(LBMOL/HR)1 = 15.14,2*0,0.2,84.07,0.59,
  TP (K,PSIA) 1 = 1066.67,179.12
END
```

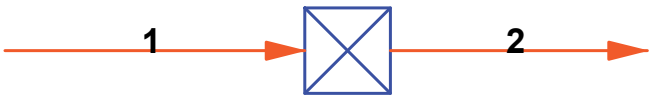
Equlb3.psd - Use Temperature Approach For Reaction 1

An example simulation showing an equilibrium reactor to set approach to equilibrium temperature of any defined reaction.


```
MYXXX.CGR
*Adiabatic Reaction

EQU REA 1 = Adiabatic Reactor, 1,-2,
  ADI
  MAX ITE = 30
  R1 = (1*2+1*62 = 1*48+3*1)
  TEM REA (F) 1 = 45
  R2 = (1*48+1*62 = 1*49+3*1)
  TEM REA (F) 2 = 0

GENERAL
  COMPONENTS = 2,49,48,1,62,
  PENK
  PENH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  BINPAR = PENG1
  AME UNI OUT
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 15.14,2*0,0.2,84.07,
  TP (K,PSIA) 1 = 1066.67,179.12
END
```

Equlb4.psd - Multi-Component System at Equilibrium

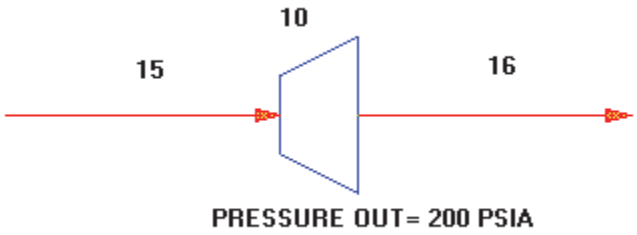
A simulation scheme for an equilibrium reactor to set the type of guessed products from a known feed of a reaction mechanism.




```
AB123.
*Multiple Reactions

EQU REA 1 = Equilibrium Reactor, 1,-2,
  MAX ITE = 30
  GUEssed products = 2,20,10,10,3*5,3*10,15

GENERAL
  COMPONENTS = 1,2,3,4,5,6,7,49,48,62,
  LKPK
  LKPH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  BINPAR = PENG1
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 2,20,2*10,3*5,2*10,15,
  TP (F,PSIA) 1 = 600,100
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
END
```

Chapter 24: Expander Samples - c:\designii\samples\equipmnt\expander

Expnd1.psd - Expansion of a Stream



Expand a stream from 800 PSIA to 200 PSIA with an efficiency of 80%.

```
EXPander 10 = EXPD, 15, -16,
        PREssure OUT = 200,
        EFFiciency = 0.80
```

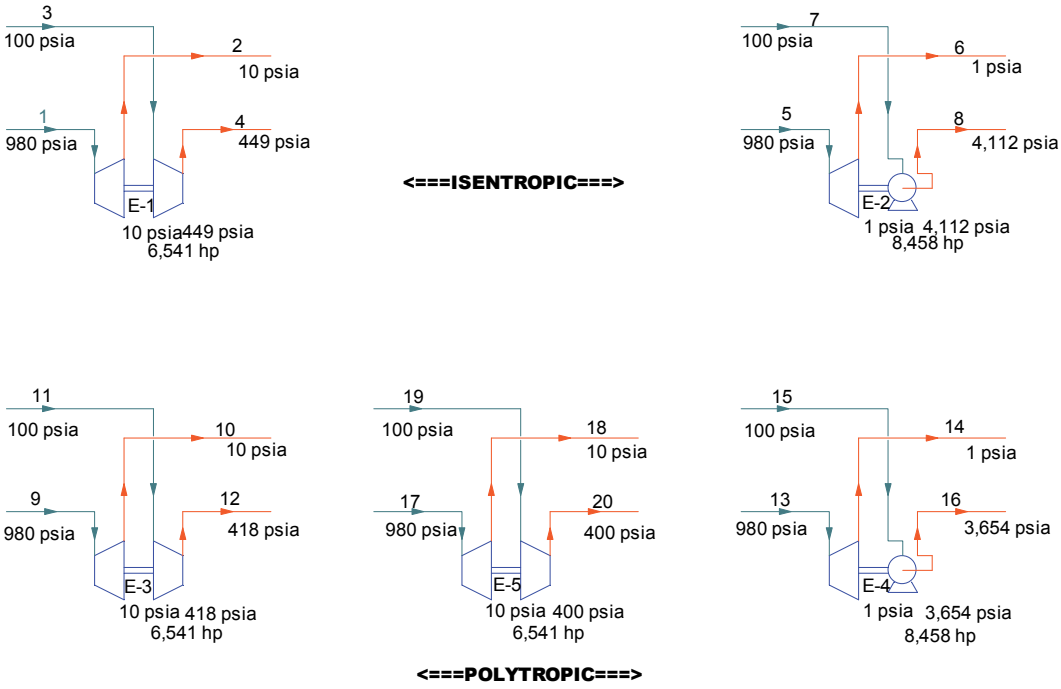
Method of Calculation

All properties of stream 15 are known. The feed is expanded at constant entropy to 200 PSIA. The efficiency is applied to calculate the work produced and then to calculate the temperature and phase distribution.

Chapter 25: Expander Compressor Samples - c:\designii\samples\equipmnt\expcom

Expander Compressor 1.psd - Expansion of a Stream

A sample flowsheet showing all five of the possible options for passing the calculated work by the expander to the compressor/pump's maximum available work. Both isentropic and polytropic options are shown for compressor section of the combined module. Note that the polytropic coefficient cannot be calculated or provided for pump.



Exp Com E-1: Natural gas at 980 psia is expanded at 80% isentropic efficiency to 10 psia which compresses another natural gas stream from 100 psia to 449 psia at an isentropic efficiency of 75%.

Exp Com E-2: Natural gas at 980 psia is expanded at 80% isentropic efficiency to 1 psia which compresses a pure water stream from 100 psia to 4,112 psia at an isentropic efficiency of 90%.

Exp Com E-3: Natural gas at 980 psia is expanded at 80% isentropic efficiency to 10 psia which compresses another natural gas stream from 100 psia to 418 psia at a polytropic efficiency of 72%.

Exp Com E-4: Natural gas at 980 psia is expanded at 80% isentropic efficiency to 1 psia which compresses a pure water stream from 100 psia to 3,654 psia at a polytropic efficiency of 72%.

Exp Com E-5: Natural gas at 980 psia is expanded at 80% isentropic efficiency to 10 psia which compresses another natural gas stream from 100 psia to 449 psia at a polytropic efficiency of 72%.

```
AB123.
*Combined expander compressor module example using all options

C- EQUIPMENT MODULES

EXP COM 1 = E-1, 1,3,-2,-4,
  PRE OUT EXP(PSIA) = 10
  ISE EFF EXP = 0.8
  ISE EFF COM = 0.75
  COU EFF = 0.98
  COU LOS(HP) = 1.5

EXP COM 2 = E-2, 5,7,-6,-8,
  PRE OUT EXP(PSIA) = 1
  ISE EFF EXP = 0.8
  ISE EFF COM = 0.9
  COU EFF = 0.98
```

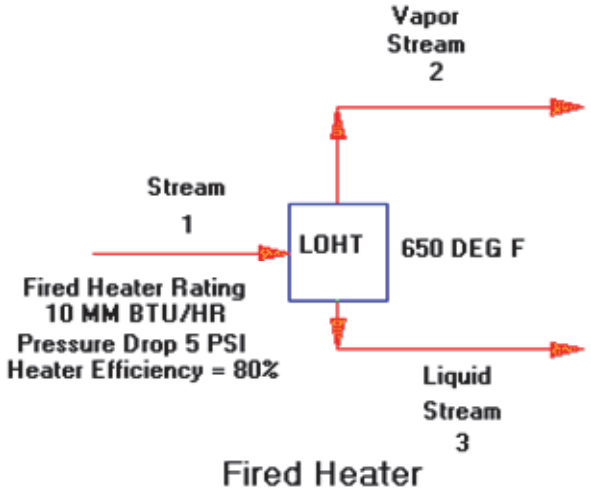
```
  COU LOS(HP) = 1.5
EXP COM 3 = E-3, 9,11,-10,-12,
  PRE OUT EXP(PSIA) = 10
  ISE EFF EXP = 0.8
  POL EFF COM = 0.72
  POL COE COM = 0.22
  COU EFF = 0.98
  COU LOS(HP) = 1.5

EXP COM 4 = E-4, 13,15,-14,-16,
  PRE OUT EXP(PSIA) = 1
  ISE EFF EXP = 0.8
  POL EFF COM = 0.72
  COU EFF = 0.98
  COU LOS(HP) = 1.5

EXP COM 5 = E-5, 17,19,-18,-20,
  PRE OUT EXP(PSIA) = 10
  ISE EFF EXP = 0.8
  POL EFF COM = 0.72
  CAL POL COM
  COU EFF = 0.98
  COU LOS(HP) = 1.5

GENERAL
  COMPONENTS = 62,2,3,4,6,8,10,
  NAM 62 = 'WATER'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'BUTANE'
  NAM 8 = 'N-PENTANE'
  NAM 10 = 'N-HEXANE'
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  BINPAR = PENG1
  AME UNI OUT
  MAX = 10
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 1 = 90,980
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 3 = 60,100
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  FLO(LBMOL/HR)5 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 5 = 90,980
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  FLO(LBMOL/HR)7 = 100000,
  TP (F,PSIA) 7 = 60,100
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  FLO(LBMOL/HR)9 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 9 = 90,980
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  FLO(LBMOL/HR)11 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 11 = 60,100
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  FLO(LBMOL/HR)13 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 13 = 90,980
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  FLO(LBMOL/HR)15 = 100000,
  TP (F,PSIA) 15 = 60,100
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
  FLO(LBMOL/HR)17 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 17 = 90,980
  NAM STR 17 = Strm 17
```

```
  NAM STR 18 = Strm 18
  FLO(LBMOL/HR)19 = 0,6100,500,200,2*100,70,
  TP (F,PSIA) 19 = 60,100
  NAM STR 19 = Strm 19
  NAM STR 20 = Strm 20
END
```

Chapter 26: Fired Heater Samples - c:\designii\samples\equipmnt\firedhea

Fdhtr1.psd - Model

All properties of stream 1 are known. The stream is heated to 650F. If the computed duty at 650F exceeds the rated duty (10 MMBTU/HR), the temperature achievable with rated duty is calculated. The fuel required is figured from the calculated duty, heating value of the fuel (980 BTU/SCF), and the efficiency of the fuel (80%). The outlet stream is separated into its vapor (stream 2) and liquid (stream 3) phases.



Given the information below, code the fired heater module. Rated Duty = 10 MMBTU/HR

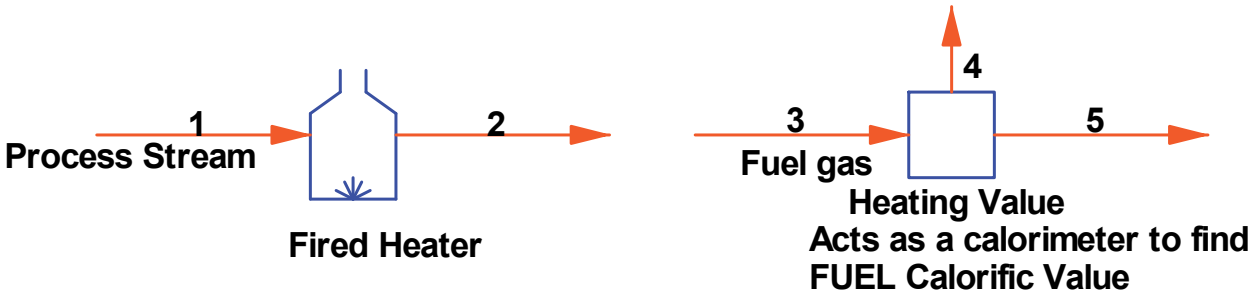
Pressure Drop = 5 PSI
Temperature Out = 650F Fuel Efficiency = 80%
Heating Value = 980 BTU/SCF

Coding

```
FIRed heater 1 = LOHT, 1, -2, -3,
        TEMperature OUT = 650
        RATing of HEAter = 10.E6,
        DELta pressure = 5,
        EFFiciency = .80
        HEAting VALue of FUEl (BTU/SCF) = 980
```

Fdhtr2.psd - Pass Fuel Calorific Value

Use of fired heater with stream manipulator passing heating value of the fuel via Inline Fortran.




```
AB123.
*FIRED HEATER MODEL

C- EQUIPMENT MODULES
```

```
STR 1 = Heating Value, 3,-5,-4,
  HEA VAL

FIR 2 = Fired Heater, 1,-2,
  TEM OUT (F) = 650
  RAT HEA (BTU/HR) = 100000000
  DEL (PSI) = 5
  EFF (FRA) = 0.8
  HEA VAL FUE (BTU/SCF) = 980
  TEM EXC (F) = 250
F-      PRE
F-      CAL=GET(STRMAN(1),CAL MOL)
F-      SET(FIR(2),HEA VAL FUE)=CAL
F-      PRINT CAL


C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,2,3,4,5,
  NAM 49 = 'CARBON DIOXIDE'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 5 = 'I-BUTANE'
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 0,1000,200,2*1,
  TP (F,PSIA) 1 = 200,200
  NAM STR 1 = Process Stream
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 0.02,0.8,0.1,0.05,0.03,
  TOT FLO (LBMOL/HR) 3 = 100
  TP (F,PSIA) 3 = 100,100
  NAM STR 3 = Fuel gas
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
END
```

Chapter 27: Flare Header Samples - c:\designii\samples\flare

PII Flare Header.psd - Flare/Vent System Design & Rating

USE DESIGN II FOR PROCESS FLARE/VENT SYSTEM DESIGN & RATING

TO DESIGN A FLARE/VENT SYSTEM, THE RELIEF LOADS MUST BE KNOWN. THESE ARE NORMALLY DETERMINED BY ANALYSIS OF THE CREDIBLE RELIEF SCENARIOS (BLOCKED OUTLET, FIRE, ETC.) FOR THE SUBJECT EQUIPMENT..THE SCENARIO THAT GENERATES THE HIGHEST RELIEF RATE IS THE "WORST CASE" FOR THE RELIEVING DEVICE (PSV, RUPTURE DISK, ETC) AND PROVIDES THE BASIS FOR DEVICE "SIZING". THIS WORK IS PERFORMED IN ACCORDANCE WITH API 520 (OR OTHER APPLICABLE PROCEDURE). DURING THIS WORK, DESIGN II IS USED TO GENERATE PHYSICAL PROPERTY DATA FOR THE RELIEVING FLUID (UPSTREAM OF THE PSV). THE RELIEF DEVICE DEVICE MAY BE SIZED WHEN THIS WORK IS COMPLETE. SIZING WILL PROVIDE THE PSV INLET/OUTLET PIPING SIZE AND, IF "LOCAL VENTING" IS ADEQUATE, THE DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) WILL ALSO BE KNOWN.

ENVIRONMENTAL CONCERNS, HOWEVER, ARE REQUIRING THAT MORE FACILITIES ROUTE MOST - IF NOT ALL - OF THEIR RELIEF STREAMS TO A FLARE/VENT SYSTEM. IT IS NO LONGER SAFE TO ASSUME THAT, IF A RELIEVING FLUID IS "LIGHTER THAN AIR", THEN A "LOCAL VENT" WILL SUFFICE.

THERE ARE A FEW "SPECIALTY" FLARE/VENT HEADER DESIGN PROGRAMS, AND SOME DO THIS JOB VERY WELL BUT ARE ALSO QUITE EXPENSIVE. OTHERS MAY BE LESS EXPENSIVE, BUT ARE LIMITED IN SCOPE. DESIGN II PROVIDES A VERY SATISFACTORY SOLUTION WITH HEADER DESIGN PERFORMED USING THE PROCESS SIMULATOR THAT PROVIDED THE ORIGINAL PROCESS DESIGN AS WELL AS THE RELIEF LOAD PROPERTIES.

THE STEPS INVOLVED IN FLARE/VENT HEADER DESIGN ARE:

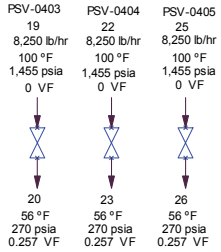
1. FINALIZE THE PHYSICAL LAYOUT OF THE EQUIPMENT THAT WILL RELIEVE TO THE HEADER AS WELL AS THE LOCATION OF THE FLARE/VENT STACK (I.E., PLOT PLAN REQUIRED).

2. PERFORM A PRELIMINARY HEADER PIPING DESIGN TO GENERATE HEADER PIPE SPOOL LENGTHS. ENSURE THAT THE EQUIVALENT PIPING LENGTHS OF FITTINGS ARE INCLUDED. FOR PIPE DIAMETERS, USE PSV OUTLET SIZES - AS DETERMINED ABOVE - FOR THE PSV "TAILPIPES" AND YOUR BEST ESTIMATE FOR THE MAIN HEADER.

3. GENERATE CREDIBLE FLARE/VENT HEADER LOADING SCENARIOS. THIS WORK IS NORMALLY PERFORMED BY A "TEAM". THE TEAM WILL DETERMINE THE AREA(S) AND SITUATIONS FOR ALL FLARE/VENT EVENTS THAT CAN OCCUR. NOTE THAT THE DESIGN II OUTPUT PROVIDES AN EXCELLENT TOOL FOR DOCUMENTING THIS WORK.

4. USING THE PHYSICAL PROPERTY AND PIPING DATA GENERATED FROM THE ABOVE PROCEDURES, CREATE A DESIGN II "CASE" FOR EACH FLARE/VENT HEADER LOADING SCENARIO. THIS WORK WILL REQUIRE SEVERAL ITERATIONS BEFORE AN OPTIMUM DESIGN IS ACHIEVED. AFTER EACH ITERATION, CHECK THE BACK-PRESSURE ON EACH DEVICE THAT "FEEDS" THE HEADER SYSTEM THEN ADJUST THE DESIGN II PIPE SIZES AS REQUIRED TO ACHIEVE A MINIMUM "SYSTEM COST". THE SYSTEM COST INCLUDES THE COSTS FOR ALL COMPONENTS IN THE FLARE/VENT SYSTEM - INCLUDING RELIEF DEVICES AND THE FLARE STACK.

WHEN COMPLETE, THE DESIGN II OUTPUT IS PASSED TO PIPING DESIGN FOR FINAL FLARE/VENT HEADER WORK. ALSO NOTE THAT THE RELIEF DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) HAS BEEN DETERMINED AND THAT RELIEF DEVICE SPECIFICATIONS CAN BE PASSED TO QUALIFIED VENDORS FOR PRICING. FINALLY, THE FLARE/VENT STACK INLET STREAM DATA FOR EACH SCENARIO SHOULD BE TRANSMITTED TO THE STACK FABRICATOR FOR DESIGN AND PRICING.



DISCONNECT PSV-0403/4/5 TAILPIPES FROM FLARE/VENT HEADER AND ROUTE TO PRODUCT SURGE DRUM

THIS IS A FLARE/VENT SYSTEM RATING CASE

THE ABOVE-NOTED PSV-0403/0404/0405 WERE RE-ROUTED TO THE NGL PRODUCT SURGE DRUM AS NOTED ABOVE. NO OTHER CHANGES TO THE SYSTEM WERE REQUIRED.

DATE: 28 Nov 12	DRAWN BY: Young	CHECKED BY:	REVISION:
Process Tech, Inc.			
PROJECT:	Panola II Flare/Vent Header Rating		
DRAWING NO:	PII Flare Header.psd		
TITLE:	**Introduction**		

```
MYXXX.MSS.EXAMPLE
*Calculate Upstream Pressure

C- EQUIPMENT MODULES

VAL 13 = PSV-0403, 19,-20,
  PRE OUT(PSIA) = 270

VAL 15 = PSV-0404, 22,-23,
  PRE OUT(PSIA) = 270

VAL 17 = PSV-0405, 25,-26,
  PRE OUT(PSIA) = 270

VAL 101 = PV-0701, 105,-106,
  PRE OUT(PSIA) = 21

LINE 102 = PV-0701P, 106,-107,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR
```

```
VAL 103 = PV-0502, 101,-102,
  PRE OUT(PSIA) = 15

LINE 104 = PV-0502P, 102,-103,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 105 = S-7, 103,-104,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 106 = Node 1, 104,107,-108,

LINE 107 = S-8, 108,-109,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 108 = S-9, 109,-110,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 38.92
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 109 = KO Drum, 110,-112,-111,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 110 = Node 1 Pressure, 112,-113,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 101
  UNTIL PRE FROM STREAM 107 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 104

CONT 111 = Flare/Vent Pressure, 113,-114,
  VARY PRE OUT (MIN=0,MAX=0,STE=2,PSIA) OF VAL 103
  UNTIL PRE FROM STREAM 112 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

VAL 201 = PV-0701, 215,-216,
  PRE OUT(PSIA) = 28.2

LINE 202 = PV-0701P, 216,-217,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
```

```
  PIP MAT = CAR
  SUR AIR

VAL 203 = PV-0502, 210,-211,
  PRE OUT(PSIA) = 30

LINE 204 = PV-0502P, 211,-212,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 205 = PSV-0401, 205,-206,
  PRE OUT(PSIA) = 34.4

LINE 206 = PSV-0401P, 206,-207,
  DIA (INCH) = 8
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 71.9 FT
  LEN (FT) = 71.9,
  SEG LEN (FT) = 7.19
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 207 = MV-2, 201,-202,
  PRE OUT(PSIA) = 50.2

LINE 208 = MV-2P, 202,-203,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 33.4 FT
  LEN (FT) = 33.4,
  SEG LEN (FT) = 3.34
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 209 = S-5, 203,-204,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 60 FT
  LEN (FT) = 60,
  SEG LEN (FT) = 6
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 210 = Node 3, 207,204,-208,

LINE 211 = S-6, 208,-209,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35 FT
  LEN (FT) = 35,
  SEG LEN (FT) = 3.5
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 212 = Node 2, 212,209,-213,

LINE 213 = S-7, 213,-214,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
```

```
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 214 = Node 1, 214,217,-218,

LINE 215 = S-8, 218,-219,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 216 = S-9, 219,-220,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 19.46
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 217 = KO Drum, 220,-222,-221,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 218 = Node 3 Pressure, 222,-223,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 205
  UNTIL PRE FROM STREAM 207 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 204

CONT 219 = Node 2 Pressure, 223,-224,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 203
  UNTIL PRE FROM STREAM 212 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 209

CONT 220 = Node 1 Pressure, 224,-225,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 201
  UNTIL PRE FROM STREAM 217 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 214

CONT 221 = Flare/Vent Pressure, 225,-226,
  VARY PRE OUT (MIN=0,MAX=0,STE=2,PSIA) OF VAL 207
  UNTIL PRE FROM STREAM 222 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

VAL 301 = PV-0701, 312,-313,
  PRE OUT(PSIA) = 21

LINE 302 = PV-0701P, 313,-314,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 303 = PV-0502, 307,-308,
  PRE OUT(PSIA) = 15

LINE 304 = PV-0502P, 308,-309,
  DIA (INCH) = 1
  PIP WAL = STD
```

```
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 305 = MV-10, 301,-302,
  PRE OUT(PSIA) = 22.6

LINE 306 = MV-10P, 302,-303,
  DIA (INCH) = 2
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.1 FT
  LEN (FT) = 43.1,
  SEG LEN (FT) = 4.31
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 307 = S-4, 303,-304,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10 FT
  LEN (FT) = 10,
  SEG LEN (FT) = 1
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 308 = S-5, 304,-305,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 60 FT
  LEN (FT) = 60,
  SEG LEN (FT) = 6
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 309 = S-6, 305,-306,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35 FT
  LEN (FT) = 35,
  SEG LEN (FT) = 3.5
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 310 = Node 2, 309,306,-310,

LINE 311 = S-7, 310,-311,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 312 = Node 1, 311,314,-315,

LINE 313 = S-8, 315,-316,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
```

```
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 314 = S-9, 316,-317,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 38.92
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 315 = KO Drum, 317,-319,-318,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 316 = Node 2 Pressure, 319,-320,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 303
  UNTIL PRE FROM STREAM 309 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 306

CONT 317 = Node 1 Pressure, 320,-321,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 301
  UNTIL PRE FROM STREAM 314 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 311

CONT 318 = Flare/Vent Pressure, 321,-322,
  VARY PRE OUT (MIN=0,MAX=0,STE=2,PSIA) OF VAL 305
  UNTIL PRE FROM STREAM 319 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

VAL 401 = PV-0701, 413,-414,
  PRE OUT(PSIA) = 21

LINE 402 = PV-0701P, 414,-415,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 403 = PV-0502, 408,-409,
  PRE OUT(PSIA) = 15

LINE 404 = PV-0502P, 409,-410,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 405 = MV-9, 401,-402,
  PRE OUT(PSIA) = 21.2

LINE 406 = MV-9P, 402,-403,
  DIA (INCH) = 2
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 53.1 FT
  LEN (FT) = 53.1,
  SEG LEN (FT) = 5.31
  TEM SUR(F) = 72,
  PIP MAT = CAR
```

```
  SUR AIR

LINE 407 = S-3, 403,-404,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 40 FT
  LEN (FT) = 40,
  SEG LEN (FT) = 4
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 408 = S-4, 404,-405,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10 FT
  LEN (FT) = 10,
  SEG LEN (FT) = 1
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 409 = S-5, 405,-406,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 60 FT
  LEN (FT) = 60,
  SEG LEN (FT) = 6
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 410 = S-6, 406,-407,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35 FT
  LEN (FT) = 35,
  SEG LEN (FT) = 3.5
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 411 = Node 2, 410,407,-411,

LINE 412 = S-7, 411,-412,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 413 = Node 1, 412,415,-416,

LINE 414 = S-8, 416,-417,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 415 = S-9, 417,-418,
  DIA (INCH) = 14
  PIP WAL = STD
```

```
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 38.92
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 416 = KO Drum, 418,-420,-419,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 417 = Node 2 Pressure, 420,-421,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 403
  UNTIL PRE FROM STREAM 410 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 407

CONT 418 = Node 1 Pressure, 421,-422,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 401
  UNTIL PRE FROM STREAM 415 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 412

CONT 419 = Flare/Vent Pressure, 422,-423,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 405
  UNTIL PRE FROM STREAM 420 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

VAL 501 = PV-0701, 515,-516,
  PRE OUT(PSIA) = 21

LINE 502 = PV-0701P, 516,-517,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 503 = PV-0502, 510,-511,
  PRE OUT(PSIA) = 15

LINE 504 = PV-0502P, 511,-512,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 505 = MV-1, 501,-502,
  PRE OUT(PSIA) = 24.5

LINE 506 = MV-1P, 502,-503,
  DIA (INCH) = 3
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 38 FT
  LEN (FT) = 38,
  SEG LEN (FT) = 1.9
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 507 = S-1, 503,-504,
  DIA (INCH) = 3
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 57.8 FT
  LEN (FT) = 57.8,
  SEG LEN (FT) = 5.78
```

```
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 508 = S-2, 504,-505,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 40 FT
  LEN (FT) = 40,
  SEG LEN (FT) = 4
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 509 = S-3, 505,-506,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 40 FT
  LEN (FT) = 40,
  SEG LEN (FT) = 4
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 510 = S-4, 506,-507,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10 FT
  LEN (FT) = 10,
  SEG LEN (FT) = 1
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 511 = S-5, 507,-508,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 60 FT
  LEN (FT) = 60,
  SEG LEN (FT) = 6
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 512 = S-6, 508,-509,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35 FT
  LEN (FT) = 35,
  SEG LEN (FT) = 3.5
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 513 = Node 2, 512,509,-513,

LINE 514 = S-7, 513,-514,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 515 = Node 1, 514,517,-518,

LINE 516 = S-8, 518,-519,
  DIA (INCH) = 14
```

```
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 517 = S-9, 519,-520,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 38.92
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 518 = KO Drum, 520,-522,-521,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 519 = Node 2 Pressure, 522,-523,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 503
  UNTIL PRE FROM STREAM 512 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 509

CONT 520 = Node 1 Pressure, 523,-524,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 501
  UNTIL PRE FROM STREAM 517 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 514

CONT 521 = Flare/Vent Pressure, 524,-525,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 505
  UNTIL PRE FROM STREAM 522 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

VAL 601 = PV-0701, 615,-616,
  PRE OUT(PSIA) = 21

LINE 602 = PV-0701P, 616,-617,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 43.8 FT
  LEN (FT) = 43.8,
  SEG LEN (FT) = 4.38
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 603 = PV--0502, 610,-611,
  PRE OUT(PSIA) = 15

LINE 604 = PV-0502P, 611,-612,
  DIA (INCH) = 1
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 37.8 FT
  LEN (FT) = 37.8,
  SEG LEN (FT) = 3.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

VAL 605 = MV-8, 601,-602,
  PRE OUT(PSIA) = 25.1

LINE 606 = MV-8P, 602,-603,
  DIA (INCH) = 3
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 45.5 FT
```

```
  LEN (FT) = 45.5,
  SEG LEN (FT) = 2.275
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 607 = S-1, 603,-604,
  DIA (INCH) = 3
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 57.8 FT
  LEN (FT) = 57.8,
  SEG LEN (FT) = 5.78
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 608 = S-2, 604,-605,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 40 FT
  LEN (FT) = 40,
  SEG LEN (FT) = 4
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 609 = S-3, 605,-606,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 40 FT
  LEN (FT) = 40,
  SEG LEN (FT) = 4
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 610 = S-4, 606,-607,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10 FT
  LEN (FT) = 10,
  SEG LEN (FT) = 1
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 611 = S-5, 607,-608,
  DIA (INCH) = 10
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 60 FT
  LEN (FT) = 60,
  SEG LEN (FT) = 6
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 612 = S-6, 608,-609,
  DIA (INCH) = 12
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35 FT
  LEN (FT) = 35,
  SEG LEN (FT) = 3.5
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 613 = Node 2, 612,609,-613,

LINE 614 = S-7, 613,-614,
  DIA (INCH) = 12
  PIP WAL = STD
```

```
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 265 FT
  LEN (FT) = 265,
  SEG LEN (FT) = 26.5
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

MIX 615 = Node 1, 614,617,-618,


LINE 616 = S-8, 618,-619,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 110 FT
  LEN (FT) = 110,
  SEG LEN (FT) = 11
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

LINE 617 = S-9, 619,-620,
  DIA (INCH) = 14
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 389.2 FT
  LEN (FT) = 389.2,
  SEG LEN (FT) = 38.92
  MET HOL = DUK
  TEM SUR(F) = 72,
  PIP MAT = CAR
  SUR AIR

FLA 618 = KO Drum, 620,-622,-621,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

CONT 619 = Node 2 Pressure, 622,-623,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 603
  UNTIL PRE FROM STREAM 612 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 609

CONT 620 = Node 1 Pressure, 623,-624,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 601
  UNTIL PRE FROM STREAM 617 =
  PRE (SCA = 1,BEG=1,LOO=30,TOL=0) OF STREAM 614

CONT 621 = Flare/Vent Pressure, 624,-625,
  VARY PRE OUT (MIN=0,MAX=0,STE=0,PSIA) OF VAL 605
  UNTIL PRE FROM STREAM 622 =
  14.7 (BEG=1,LOO=30,TOL=0,PSIA)

SIZE LINE
  USE streams = 101
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 105
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 106
  MAX SON(FRA) = 0.8
  PIP WAL = STD
```

```
  DIR = HORI
  LEN(FT) = 43.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 108
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 109
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 201
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 202
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 33.4
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 203
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 205
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 206
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 208
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 35
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
```

```
SIZE LINE
  USE streams = 210
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 211
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 213
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 265
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 215
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 218
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 219
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 304
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 305
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 35
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 307
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
```

```
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 308
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 310
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 265
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 312
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 315
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 316
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 404
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 405
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 406
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 35
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
```

```
  USE streams = 408
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 409
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 411
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 265
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 413
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 416
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 417
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 502
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 504
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 505
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
```

```
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 506
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 507
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 508
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 35
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 510
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 511
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 513
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 265
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 515
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 518
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 519
```

```
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 602
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 604
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 605
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 606
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 607
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 60
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 608
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 35
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 610
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 611
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 37.8
  ROU(FT) = 0.00015
```

```
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 613
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 265
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 615
  MAX DEL(PSIA) = 0.5
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 100
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 618
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 110
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1
SIZE LINE
  USE streams = 619
  MAX SON(FRA) = 0.9
  PIP WAL = STD
  DIR = HORI
  LEN(FT) = 389.2
  ROU(FT) = 0.00015
  DRA(FRA) = 1
  EFF(FRA) = 1

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,11
  NAM POS 1 = 'NITROGEN'
  NAM POS 2 = 'CO2'
  NAM POS 3 = 'METHANE'
  NAM POS 4 = 'ETHANE'
  NAM POS 5 = 'PROPANE'
  NAM POS 6 = 'I-BUTANE'
  NAM POS 7 = 'N-BUTANE'
  NAM POS 8 = 'I-PENTANE'
  NAM POS 9 = 'N-PENTANE'
  NAM POS 10 = 'N-HEXANE'
  NAM POS 11 = 'N-HEPTANE'
C-  Thermo Comments
  PENK
  PENH
  STDD
  DEN LIQ = STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
  THE CON VAP = NBS81
  THE CON LIQ = NBS81
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  BINPAR = PENG1
  PET UNI OUT
  AME UNI OUT
  MAX = 5
  CON TOL = 0.001
  FLO (FRA MOL) 19 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,0.0143,
            0.03,0.00713,
  TOT FLO (LB/HR) 19 = 8250
  TP (F,PSIA) 19 = 100,1454.696
  NAM STR 19 = Strm 119
```

```
NAM STR 20 = Strm 127
FLO (FRA MOL) 22 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,0.0143,
          0.03,0.00713,
TOT FLO (LB/HR) 22 = 8250
TP (F,PSIA) 22 = 100,1454.696
NAM STR 22 = Strm 127
NAM STR 23 = Strm 127
FLO (FRA MOL) 25 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,0.0143,
          0.03,0.00713,
TOT FLO (LB/HR) 25 = 8250
TP (F,PSIA) 25 = 100,1454.696
NAM STR 25 = Strm 128
NAM STR 26 = Strm 127
FLO (FRA MOL) 101 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 101 = 84
TP (F,PSIG) 101 = 100,50
NAM STR 101 = V-502 Vent
NAM STR 102 = PV-0502 Outlet
NAM STR 103 = PV-0502 @ Header
NAM STR 104 = S7-Node 1
FLO (FRA MOL) 105 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 105 = 19.7267
TP (F,PSIA) 105 = 100,64.696
NAM STR 105 = V-701 Vent
FLO (FRA MOL) 106 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 106 = 19.7267
TP (F,PSIA) 106 = 97.36119,23.43262
NAM STR 106 = PV-0701 Outlet
NAM STR 107 = PV-0701 @ Header
NAM STR 108 = Node 1-S8
NAM STR 109 = S8-S9
NAM STR 110 = S9-KO
NAM STR 111 = KO Drum Liquid
FLO (FRA MOL) 112 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
          0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
TOT FLO (LBMOL/HR) 112 = 1781.44
TP (F,PSIA) 112 = -56.99598,14.70034
NAM STR 112 = Flare/Vent Inlet
NAM STR 113 = Flare/Vent
NAM STR 114 = Flare Vent
FLO (FRA MOL) 201 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LB/HR) 201 = 35000
TP (F,PSIA) 201 = 120,254.696
NAM STR 201 = EC-201 Blowdown
NAM STR 202 = MV-2 Outlet
FLO (FRA MOL) 203 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 203 = 1236.6
TP (F,PSIA) 203 = 107.6384,24.90367
NAM STR 203 = MV-2 @ Header
NAM STR 204 = S5-Node 3
FLO (FRA MOL) 205 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,
          0.0143,0.03,0.00713,
TOT FLO (LB/HR) 205 = 86959
TP (F,PSIG) 205 = 423,350
VAP FRA 205 = 1, 1
NAM STR 205 = V-401 @ Relief
FLO (FRA MOL) 206 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,
          0.0143,0.03,0.00713,
TOT FLO (LBMOL/HR) 206 = 2230.22
TP (F,PSIA) 206 = -46.38606,21.40333
NAM STR 206 = PSV-0401 Outlet
FLO (FRA MOL) 207 = 0,0.0006,0.0084,0.6415,0.1762,0.04817,0.0482,0.0255,
          0.0143,0.03,0.00713,
TOT FLO (LBMOL/HR) 207 = 2230.22
TP (F,PSIA) 207 = 134.96,24.72973
NAM STR 207 = PSV-0401 @ Header
NAM STR 208 = Node 3-S6
NAM STR 209 = S6-Node 2
FLO (FRA MOL) 210 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 210 = 84
TP (F,PSIG) 210 = 100,50
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NAM STR 210 = V-502 Vent
FLO (FRA MOL) 211 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 211 = 4.72092
TP (F,PSIA) 211 = 97.18234,20.65752
NAM STR 211 = PV-0502 Outlet
FLO (FRA MOL) 212 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 212 = 24.4476
TP (F,PSIA) 212 = 79.09395,14.69933
NAM STR 212 = PV-0502 @ Header
NAM STR 213 = Node 2-S7
NAM STR 214 = S7-Node 1
FLO (FRA MOL) 215 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 215 = 19.7267
TP (F,PSIA) 215 = 100,64.696
NAM STR 215 = V-701 Vent
NAM STR 216 = PV-0701 Outlet
FLO (FRA MOL) 217 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 217 = 19.7267
TP (F,PSIA) 217 = 95.87918,14.69966
NAM STR 217 = PV-0701 @ Header
NAM STR 218 = Node 1-S8
NAM STR 219 = S8-S9
NAM STR 220 = S9-KO
NAM STR 221 = KO Drum Liquid
FLO (FRA MOL) 222 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
          0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
TOT FLO (LBMOL/HR) 222 = 1781.44
TP (F,PSIA) 222 = -56.99598,14.70034
NAM STR 222 = Flare/Vent Inlet
FLO (FRA MOL) 223 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
          0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
TOT FLO (LBMOL/HR) 223 = 3491.28
TP (F,PSIA) 223 = 126.9753,14.6999
NAM STR 223 = Flare/Vent
FLO (FRA MOL) 224 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
          0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
TOT FLO (LBMOL/HR) 224 = 3491.28
TP (F,PSIA) 224 = 126.9753,14.6999
NAM STR 224 = Flare/Vent
NAM STR 225 = Flare/Vent
NAM STR 226 = Flare/Vent
FLO (FRA MOL) 301 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 301 = 2500
TP (F,PSIA) 301 = 125,914.696
NAM STR 301 = F-202A/B Blowdown
NAM STR 302 = MV-10 Outlet
NAM STR 303 = MV-10 @ Header
FLO (FRA MOL) 304 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 304 = 1236.6
TP (F,PSIA) 304 = 107.6384,24.90367
NAM STR 304 = S4-S5
NAM STR 305 = S5-S6
NAM STR 306 = S6-Node 2
FLO (FRA MOL) 307 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 307 = 84
TP (F,PSIG) 307 = 100,50
NAM STR 307 = V-502 Vent
FLO (FRA MOL) 308 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 308 = 4.72092
TP (F,PSIA) 308 = 97.18234,20.65752
NAM STR 308 = PV-0502 Outlet
FLO (FRA MOL) 309 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 309 = 24.4476
TP (F,PSIA) 309 = 79.09395,14.69933
NAM STR 309 = PV-0502 @ Header
NAM STR 310 = Node 2-S7
NAM STR 311 = S7-Node 1
FLO (FRA MOL) 312 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
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          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 312 = 19.7267
TP (F,PSIA) 312 = 100,64.696
NAM STR 312 = V-701 Vent
NAM STR 313 = PV-0701 Outlet
FLO (FRA MOL) 314 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 314 = 19.7267
TP (F,PSIA) 314 = 95.87918,14.69966
NAM STR 314 = PV-0701 @ Header
NAM STR 315 = Node 1-S8
NAM STR 316 = S8-S9
NAM STR 317 = S9-KO
NAM STR 318 = KO Drum Liquid
FLO (FRA MOL) 319 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
          0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
TOT FLO (LBMOL/HR) 319 = 1781.44
TP (F,PSIA) 319 = -56.99598,14.70034
NAM STR 319 = Flare/Vent Inlet
FLO (FRA MOL) 320 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
          0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
TOT FLO (LBMOL/HR) 320 = 3491.28
TP (F,PSIA) 320 = 126.9753,14.6999
NAM STR 320 = Flare/Vent
NAM STR 321 = Flare/Vent
NAM STR 322 = Flare/Vent
FLO (FRA MOL) 401 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 401 = 2000
TP (F,PSIA) 401 = 125,914.696
NAM STR 401 = T-601 Blowdown
NAM STR 402 = MV-9 Outlet
NAM STR 403 = MV-9 @ Header
FLO (FRA MOL) 404 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 404 = 1236.6
TP (F,PSIA) 404 = 107.6384,24.90367
NAM STR 404 = S3-S4
FLO (FRA MOL) 405 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 405 = 1236.6
TP (F,PSIA) 405 = 107.6384,24.90367
NAM STR 405 = S4-S5
NAM STR 406 = S5-S6
NAM STR 407 = S6-Node 2
FLO (FRA MOL) 408 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 408 = 84
TP (F,PSIG) 408 = 100,50
NAM STR 408 = V-502 Vent
FLO (FRA MOL) 409 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 409 = 4.72092
TP (F,PSIA) 409 = 97.18234,20.65752
NAM STR 409 = PV-0502 Outlet
FLO (FRA MOL) 410 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 410 = 24.4476
TP (F,PSIA) 410 = 79.09395,14.69933
NAM STR 410 = PV-0502 @ Header
NAM STR 411 = Node 2-S7
NAM STR 412 = S7-Node 1
FLO (FRA MOL) 413 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 413 = 19.7267
TP (F,PSIA) 413 = 100,64.696
NAM STR 413 = V-701 Vent
NAM STR 414 = PV-0701 Outlet
FLO (FRA MOL) 415 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 415 = 19.7267
TP (F,PSIA) 415 = 95.87918,14.69966
NAM STR 415 = PV-0701 @ Header
NAM STR 416 = Node 1-S8
NAM STR 417 = S8-S9
NAM STR 418 = S9-KO
NAM STR 419 = KO Drum Liquid
```
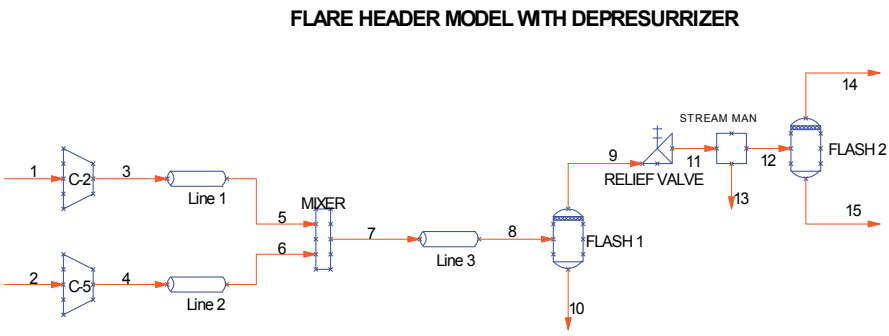

```
FLO (FRA MOL) 420 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
          0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
TOT FLO (LBMOL/HR) 420 = 1781.44
TP (F,PSIA) 420 = -56.99598,14.70034
NAM STR 420 = Flare/Vent Inlet
FLO (FRA MOL) 421 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
          0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
TOT FLO (LBMOL/HR) 421 = 3491.28
TP (F,PSIA) 421 = 126.9753,14.6999
NAM STR 421 = Flare/Vent
NAM STR 422 = Flare/Vent
NAM STR 423 = Flare/Vent
FLO (FRA MOL) 501 = 0.01704,0.0073,0.9052,0.0464,0.0121,2*0.0033,0.00175,
          0.001,2e-005,0.00259,
TOT FLO (LB/HR) 501 = 5550
TP (F,PSIA) 501 = 100,900
NAM STR 501 = F-101 Blowdown
FLO (FRA MOL) 502 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 502 = 4.72092
TP (F,PSIA) 502 = 97.18234,20.65752
NAM STR 502 = MV-1 Outlet
NAM STR 503 = MV-1 @ Header
FLO (FRA MOL) 504 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 504 = 1236.6
TP (F,PSIA) 504 = 107.6384,24.90367
NAM STR 504 = S1-S2
FLO (FRA MOL) 505 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 505 = 1236.6
TP (F,PSIA) 505 = 107.6384,24.90367
NAM STR 505 = S2-S3
FLO (FRA MOL) 506 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 506 = 1236.6
TP (F,PSIA) 506 = 107.6384,24.90367
NAM STR 506 = S3-S4
FLO (FRA MOL) 507 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 507 = 1236.6
TP (F,PSIA) 507 = 107.6384,24.90367
NAM STR 507 = S4-S5
NAM STR 508 = S5-S6
NAM STR 509 = S6-Node 2
FLO (FRA MOL) 510 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 510 = 84
TP (F,PSIG) 510 = 100,50
NAM STR 510 = V-502 Vent
FLO (FRA MOL) 511 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 511 = 4.72092
TP (F,PSIA) 511 = 97.18234,20.65752
NAM STR 511 = PV-0502 Outlet
FLO (FRA MOL) 512 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 512 = 24.4476
TP (F,PSIA) 512 = 79.09395,14.69933
NAM STR 512 = PV-0502 @ Header
NAM STR 513 = Node 2-S7
NAM STR 514 = S7-Node 1
FLO (FRA MOL) 515 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 515 = 19.7267
TP (F,PSIA) 515 = 100,64.696
NAM STR 515 = V-701 Vent
NAM STR 516 = PV-0701 Outlet
FLO (FRA MOL) 517 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 517 = 19.7267
TP (F,PSIA) 517 = 95.87918,14.69966
NAM STR 517 = PV-0701 @ Header
NAM STR 518 = Node 1-S8
NAM STR 519 = S8-S9
NAM STR 520 = S9-KO
NAM STR 521 = KO Drum Liquid
```

```
FLO (FRA MOL) 522 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
          0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
TOT FLO (LBMOL/HR) 522 = 1781.44
TP (F,PSIA) 522 = -56.99598,14.70034
NAM STR 522 = Flare/Vent Inlet
FLO (FRA MOL) 523 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
          0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
TOT FLO (LBMOL/HR) 523 = 3491.28
TP (F,PSIA) 523 = 126.9753,14.6999
NAM STR 523 = Flare/Vent
NAM STR 524 = Flare/Vent
NAM STR 525 = Flare/Vent
FLO (FRA MOL) 601 = 0.01704,0.0073,0.9052,0.0464,0.0121,2*0.0033,0.00175,
          0.001,2e-005,0.00259,
TOT FLO (LB/HR) 601 = 5550
TP (F,PSIA) 601 = 100,900
NAM STR 601 = V-101 Blowdown
FLO (FRA MOL) 602 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 602 = 4.72092
TP (F,PSIA) 602 = 97.18234,20.65752
NAM STR 602 = MV-8 Outlet
NAM STR 603 = MV-8 @ Header
FLO (FRA MOL) 604 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 604 = 1236.6
TP (F,PSIA) 604 = 107.6384,24.90367
NAM STR 604 = S1-S2
FLO (FRA MOL) 605 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 605 = 1236.6
TP (F,PSIA) 605 = 107.6384,24.90367
NAM STR 605 = S2-S3
FLO (FRA MOL) 606 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 606 = 1236.6
TP (F,PSIA) 606 = 107.6384,24.90367
NAM STR 606 = S3-S4
FLO (FRA MOL) 607 = 0.00800063,0,0.989578,0.00238019,4e-005,1e-006,5.5e-007,
          3e-008,7.501e-009,2*0,
TOT FLO (LBMOL/HR) 607 = 1236.6
TP (F,PSIA) 607 = 107.6384,24.90367
NAM STR 607 = S4-S5
NAM STR 608 = S5-S6
NAM STR 609 = S6-Node 2
FLO (FRA MOL) 610 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LB/HR) 610 = 84
TP (F,PSIG) 610 = 100,50
NAM STR 610 = V-502 Vent
FLO (FRA MOL) 611 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 611 = 4.72092
TP (F,PSIA) 611 = 97.18234,20.65752
NAM STR 611 = PV-0502 Outlet
FLO (FRA MOL) 612 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 612 = 24.4476
TP (F,PSIA) 612 = 79.09395,14.69933
NAM STR 612 = PV-0502 @ Header
NAM STR 613 = Node 2-S7
NAM STR 614 = S7-Node 1
FLO (FRA MOL) 615 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 615 = 19.7267
TP (F,PSIA) 615 = 100,64.696
NAM STR 615 = V-701 Vent
NAM STR 616 = PV-0701 Outlet
FLO (FRA MOL) 617 = 0.00747,3e-005,0.9209,0.0472,0.01233,2*0.00336,0.00178,
          0.001,0.00021,0.00236,
TOT FLO (LBMOL/HR) 617 = 19.7267
TP (F,PSIA) 617 = 95.87918,14.69966
NAM STR 617 = PV-0701 @ Header
NAM STR 618 = Node 1-S8
NAM STR 619 = S8-S9
NAM STR 620 = S9-KO
NAM STR 621 = KO Drum Liquid
```

```
  FLO (FRA MOL) 622 = 0.00010246,0.00074031,0.0230816,0.764852,0.16453,
            0.0246768,0.0177591,0.00275373,0.00114087,0.00034854,1.433e-005,
  TOT FLO (LBMOL/HR) 622 = 1781.44
  TP (F,PSIA) 622 = -56.99598,14.70034
  NAM STR 622 = Flare/Vent Inlet
  FLO (FRA MOL) 623 = 0.00288612,0.00038349,0.362321,0.410963,0.112657,
            0.0307948,0.0308138,0.0163019,0.00914183,0.0191654,0.00457116,
  TOT FLO (LBMOL/HR) 623 = 3491.28
  TP (F,PSIA) 623 = 126.9753,14.6999
  NAM STR 623 = Flare/Vent
  NAM STR 624 = Flare/Vent
  NAM STR 625 = Flare/Vent
END
```

flrehdr.psd - Flare Header Model with Depressurizer



A description of this simulation is under HELP...DESIGN II Reference Guide... Example Problems...General: Flrehdr.psd Flare Header Simulation

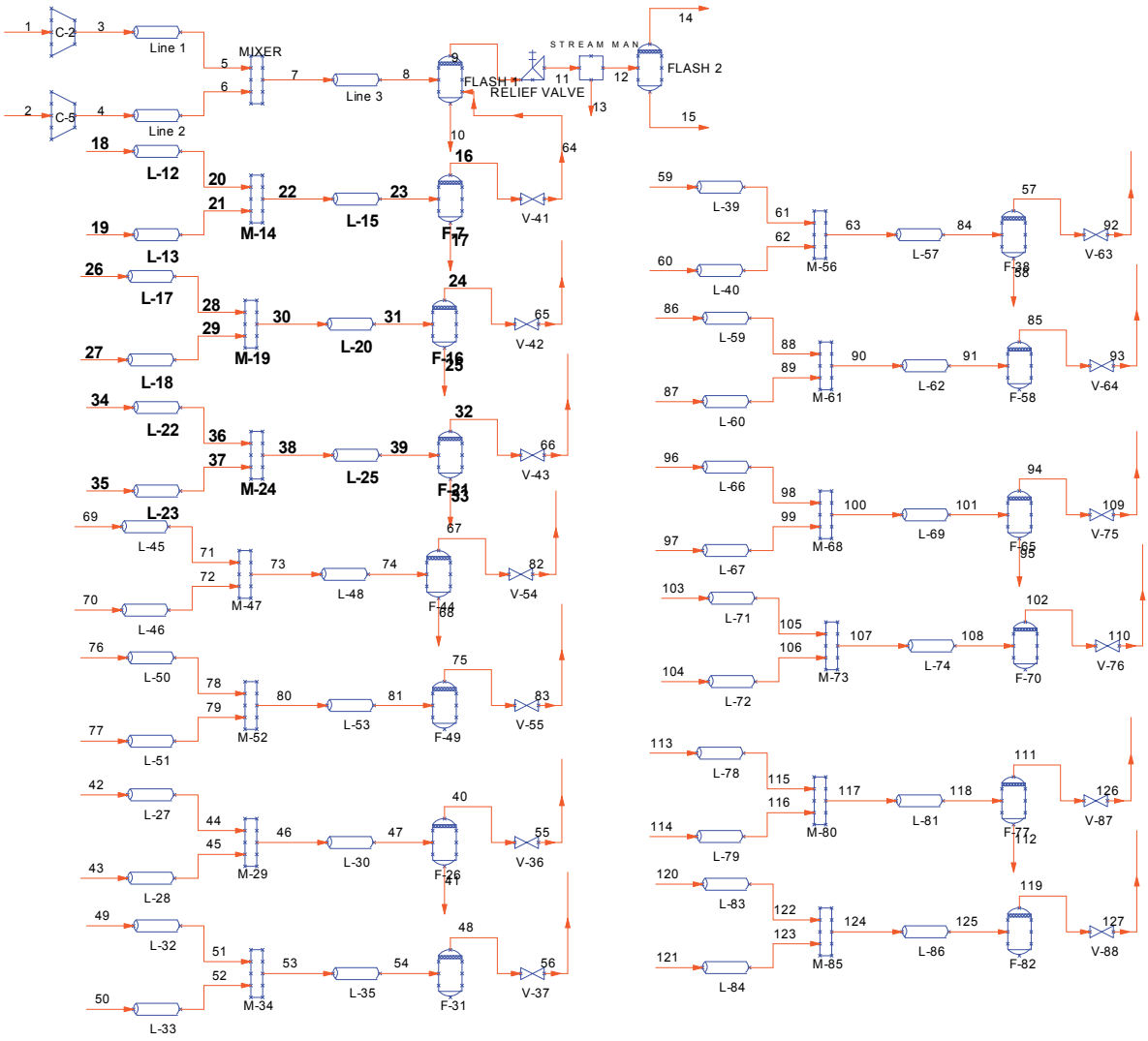
```
MYXXX.CGR.EXAMPLE
*Gas Flare Header Sample Problem

C- EQUIPMENT MODULES

STR 1 = STREAM MAN, 11,-13,-12,
  MUL
  SAM FAC, FAC = 1

COMPRE 2 = C-2, 1,-3,
  PRE OUT(PSIA) = 650
  EFF = 0.76
  CAL POL
  ELE DRI
  STA = 1

DEP 3 = RELIEF VALVE, 9,-11,
  CRO(IN2) = 3.2648
  COE = 0.975
  DIA(FT) = 15
  VER
  LEN(FT) = 120
  LIQ HEI(MM) = 0.01
  RED(PSIA) = 300
  BAC(PSIA) = 25
  SET(PSIA) = 650
  RES(PSIA) = 0
  TIME STEP(MIN) = 2
  MIN TIME(MIN) = 15
  MAX TIME(MIN) = 60
  ENV = 1
  PRI = 2
  WET = 34500
  RUP = 0
  BAL = 0
  BAC FAC = 0
  VAP(FRA) = 0.1

FLA 4 = FLASH 1, 8,-9,-10,
  TEM FEE
  PRE FEE

COMPRE 5 = C-5, 2,-4,
  PRE OUT(PSIA) = 650
```

```
  EFF = 0.76
  CAL POL
  ELE DRI
  STA = 1

LINE 6 = Line 1, 3,-5,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 100 FT
  LEN (FT) = 100,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0
  ROU (IN) = 2
  MET ACC
  ADI
      PREssure OUT (PSIA) =600

FLA 8 = FLASH 2, 12,-14,-15,
  TEM FEE
  PRE FEE

LINE 9 = Line 2, 4,-6,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 100 FT
  LEN (FT) = 100,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0
  SEG LEN (FT) = 10
  ROU (IN) = 2
  ADI
      PREssure OUT (PSIA) =600

MIX 10 = MIXER, 5,6,-7,

LINE 11 = Line 3, 7,-8,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 1000 FT
  LEN (FT) = 2*500,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0,0
  SEG LEN (FT) = 100
  ROU (IN) = 2
  ADI
      PREssure OUT (PSIA) =580

C- FLOWSHEET LINE SIZING
SIZ LIN
  USE streams = 5, 6, 8

  MAXimum SONic fraction =0.75

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,6,62
  NAM POS 1 = 'NITROGEN'
  NAM POS 2 = 'CO2'
  NAM POS 3 = 'METHANE'
  NAM POS 4 = 'ETHANE'
  NAM POS 5 = 'PROPANE'
  NAM POS 6 = 'N-BUTANE'
  NAM POS 7 = 'WATER'
C-  Thermo Comments
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  AME UNI OUT
```

```
  MAX = 25
  CON TOL = 0.001
  FLO (FRA) 1 = 0.009,0.006,0.85,0.12,0.01,0.005,
  TOT FLO (LBMOL/HR) 1 = 100
  TP (F,PSIA) 1 = -35,350
  NAM STR 1 = Strm 1
  FLO (FRA) 2 = 0.009,0.006,0.79,0.18,0.01,0.005,5e-005,
  TOT FLO (LBMOL/HR) 2 = 50
  TP (F,PSIA) 2 = -35,350
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
END
```

flrehdr1.psd - Flare Header Model with Depressurizer

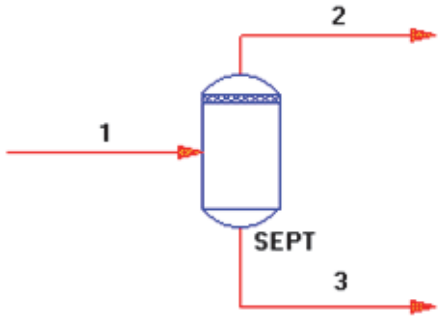
FLARE HEADER MODEL WITH DEPRESURRIZER



Chapter 28: Flash Samples - c:\designii\samples\equipmnt\flash

Flash1.psd - Phase Separation

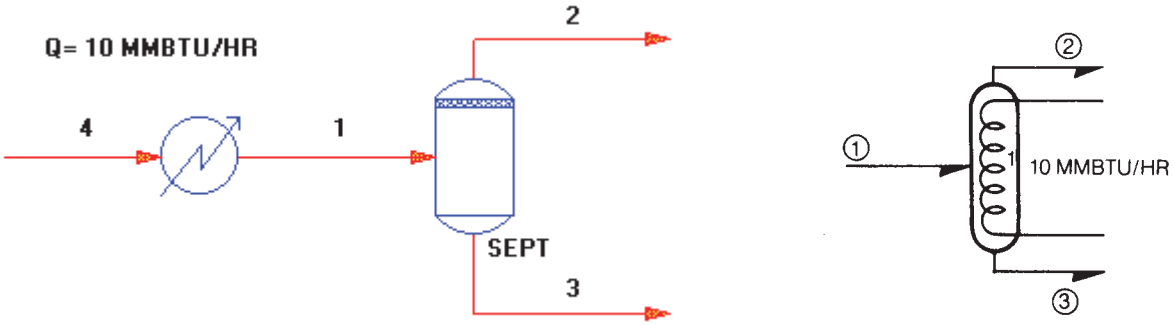
All properties of stream 1 are known. The vapor and liquid phase are separated.




```
FLAsh 1 = SEPT, 1, -2, -3
```

Flash2.psd - Heat Change And Phase Separation

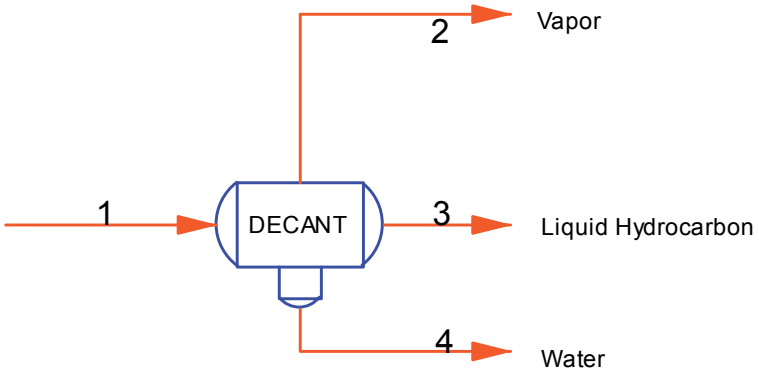
The enthalpy change is added to stream 1 and then a flash is performed at the new conditions. The vapor and liquid phases are separated into streams 2 and 3.




```
FLAsh 1 = HTFL, 1, -2, -3, HEAt change = 10.E6
```

Flash3.psd - Hydrocarbon and Water Decant

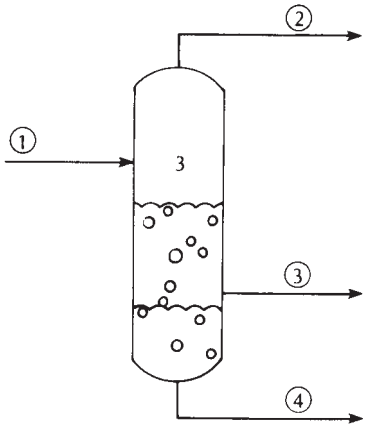
All properties of stream 8 are known. If there is any free water present, the vapor phase (stream 2) will be saturated. Stream 3 will contain liquid hydrocarbons and soluble water. Stream 4 will contain free water and soluble hydrocarbons. If CO2, H2S, and/or NH3 are present the APISOUR correlation will be used to estimate their solubility in the water phase. Enthalpies and properties of the liquid streams will be recalculated.




```
FLAsh 1 = DECANT, 1, -2, -3, -4,
GENeral,
     IMMiscible = 62
```

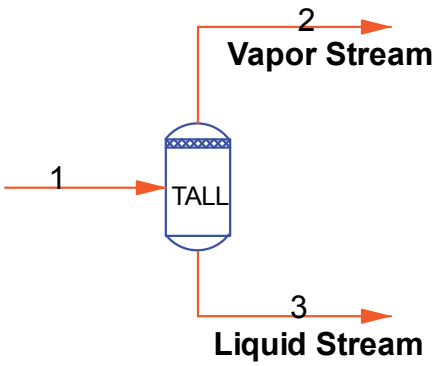
Flash4.psd - Oil and Water Entrainment

All properties of stream 1 are known. If there is any free water present, the vapor phase (stream 2) will be saturated. The composition of the liquid streams will be calculated on an iterative basis. Enthalpy and properties of the liquid streams will be recalculated.




```
FLAsh 3 = ENT, 1, -2, -3, -4,
     OIL ENTrained in WATer (VOL) = .0015,
     WATer ENTrained in OIL (VOL) = .02,
GENeral,
        IMMiscible = 62
```

Flash5.psd - Vertical Separator With Specified Height



Sizing of a vertical flash drum of specified height.

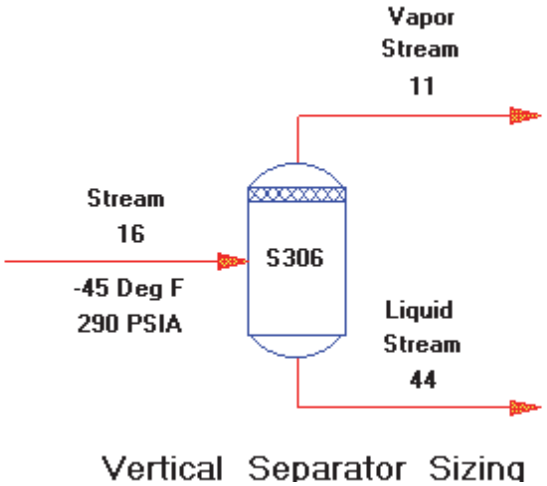
```
FLAsh 1 = TALL, 1, -2, -3,
       VERTical,
       LENgth = 7,
GENeral,
       COMponents = 2, 3, 4,
       FLOw 1 = 800, 100, 100,
       TP 1 = -45, 290,
END
```

Flash6.psd - Vertical Separator With Default Values

Sizing of a vertical flash drum with default values.

```
AB123.
*EXAMPLE OF VERTICAL SEPARATOR SIZING
FLAsh 19 = S306, 16, -11, -44,
       VERTical,
GENeral,
       COMponents = 2, 3, 4,
       FLOw 16 = 800, 100,100,
       TP 16 = -45, 290,
END
```

A sample output for separator sizing, Example 7, is shown. Additional coding examples can be found in VALVE, as well as a sample output for a horizontal separator.



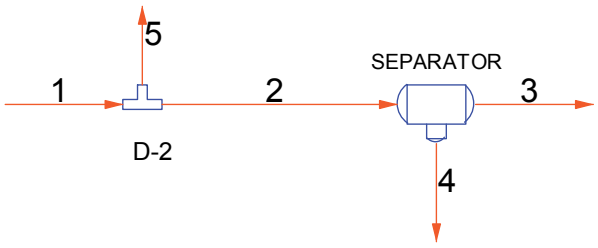

```
****** SEPARATOR SIZING RESULTS FOR VERTICAL
       SEPARATOR   19 *******
      VAPOR STREAM =  11    LIQUID STREAM =  44

CONDITIONS
----------
LIQUID      FLOW RATE          T-P         1.9953
       FT3/MIN
VAPOR       FLOW RATE          T-P         200.69
       FT3/MIN
TOTAL       FLOW RATE          T-P         202.69      FT3/MIN
TEMPERATURE                               -45.000      DEG F
PRESSURE                                   290.00      PSIA

PROPERTIES
----------
DENSITY                      LIQUID        31.118      LB/FT3
DENSITY                      VAPOR        1.3723      LB/FT3
Z-FACTOR                     VAPOR       0.88021

DESIGN PARAMETERS
-----------------
K  CONSTANT                  DEFAULT     0.35000      FT/SEC
SURGE  TIME                  DEFAULT      2.0000      MIN
MIN DISENGAGING HEIGHT       DEFAULT      3.0000      FT
MIN LIQ TO INLET HEIGHT      DEFAULT      2.0000      FT
MIN LEVEL NOZZLE HEIGHT      DEFAULT      1.5000      FT
MIST ELIMINATOR              DEFAULT     0.50000      FT
DESIGN PRESSURE              DEFAULT      319.00      PSIA
ALLOWABLE STRESS             DEFAULT      15015.      PSIA
JOINT EFFICIENCY             DEFAULT      1.0000
CORROSION ALLOWANCE (CA)     DEFAULT     0.10417E-01  FT
WEIGHT PERCENT ALLOWANCE(WA) DEFAULT      20.000

RESULTS
-------
MAX. DROPLET SETTLING VELOCITY                1.6295      FT/SEC
ACTUAL  VAPOR VELOCITY                        1.1345      FT/SEC
DIAMETER BASED ON SEPARATION                  1.6167      FT
SELECTED INTERNAL DIAMETER - PIPE             23.250      IN.
PIPE NOMINAL DIAMETER                         24.000      IN.
LENGTH                                        8.0000      FT
SURGE TIME TO HLL                             2.2164      MIN
HIGH LIQUID LEVEL (HLL) HEIGHT                1.5000      FT
HLL TO INLET HEIGHT                           3.0000      FT
DISENGAGING HEIGHT                            3.0000      FT
WALL THICKNESS (STD )                        0.31250E-01  FT
ACTUAL HEAD THICKNESS                        0.31250E-01  FT
MIN SHELL THICKNESS INCLUDING CA             0.30292E-01  FT
```

```
MIN HEAD  THICKNESS INCLUDING CA         0.30090E-01  FT
APPROX WEIGHT OF STEEL VESSEL(EMPTY)      802.65      LB
APPROX WEIGHT OF VESSEL (FULL)            1536.6      LB
APPROX TOTAL WEIGHT INCLUDING WA          1697.1      LB

NOZZLES (STD WALL)  INLET      VAP OUT     LIQ OUT
                      A           B           C
MINIMUM VELOCITY   46.497      51.218      10.756     FT/SEC
ACTUAL  VELOCITY   65.802      65.155      5.5409     FT/SEC
MAXIMUM VELOCITY   77.495      85.363      17.926     FT/SEC
NOMINAL DIAMETER   3.0000      3.0000      1.0000     IN.
     ESTIMATED SONIC VELOCITY IN VAPOR IS 1117.6 FT/SEC
 NOTE : ONE OR MORE OF THESE NOZZLES DOES NOT MEET MIN AND MAX
        VELOCITY SPECS BECAUSE SMALLEST NOZZLE SIZE IS 1 IN.
 NOTE: INSULATION MAY BE NEEDED AS TEMPERATURE =   -45.0 DEG F
THESE RESULTS ARE INTENDED FOR PRELIMINARY DESIGN WORK ONLY AND
SHOULD NOT BE USED FOR DETAILED VESSEL DESIGN TO A.S.M.E. CODE.

**** SKETCH OF VERTICAL   FLASH DRUM  19 (TO APPROX.SCALE ONLY) ****

                     === B
                      I
                     ***
                   ****   ****
                 **           **
               I+++++++++++++++I      -
               I   MIST PAD    I      I
         -     I+++++++++++++++I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
      3.00 FT  I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I      I
         I     I               I    6.50 FT
         I   I I               I      I
         -   I==I    VAPOR      I      I
             I  I               I      I
                I               I      I
             A  I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I      I
                I               I  I   I
                I-----HLL-------I==I   -
                I               I  I   I
                I               I      I
                I    LIQUID     I      I
                I               I    1.50 FT
                I               I      I
                I               I  I   I
                I               I==I   -
                 **           **   I
                   ****   ****
                       *I*
                       === C

                 --   2.00 FT---
```

Flash7.psd - Separator sizing with Case Study

Detailed sizing of a vertical flash drum along with a case study for multiple flowrates though the drum.

```
AB123.
*Separator Sizing...CASE STUDY

FLA 1 = SEPARATOR, 2,-3,-4,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0
  VERT
  MIS ELI
  KCO (FT/SEC) = 0.32
  RAT MAJ  = 2
  LEN SHE (FT) = 6.25
  SUR (MIN) = 2
  MIN DIS (FT) = 1.5
  MIN LIQ (FT) = 0.5

DIV 2 = D-2, 1,-2,-5,
  FLO(FRA) = 0.65

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,46,2,3,22,23,
  NAM 1 = 'HYDROGEN'
  NAM 46 = 'NITROGEN'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 22 = 'ETHYLENE'
  NAM 23 = 'PROPYLENE'
  SOAVEK
  SOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  FLO(LB/HR)1 = 4474,188,19928,2745,10265,11,
  TP (F,PSIG) 1 = -200,434
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
CASE
CASE BLOCK , NAME = SEP. SIZE
STEP, DIVIDER 2, FLO RAT = .95, INCR = -.1, LIMIT = .05
TABLE,STREAM 5, FLO, LABEL = TOTAL FLOW
C-PLOTX,FLA 1, PRESSURE OUT, LABEL=K101 PRESSURE OUT
C-PLOTY,FLA 1, TEM , LABEL=K101 TEMPERATURE
END
```



CASE STUDY TABLE

	BASE CASE	CASE 1 SEP. SIZE	CASE 2 SEP. SIZE	CASE 3 SEP. SIZE
DIV 2,D-2				
FLO RAT	0.65000	0.95000	0.85000	0.75000
TOTAL FLOW				
FLO, LBMOL/HR	1374.0	196.28	588.85	981.42

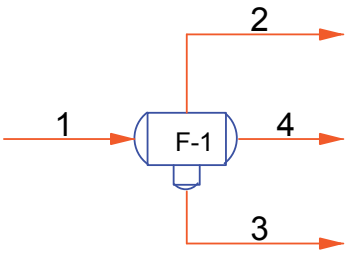
	CASE 4 SEP. SIZE	CASE 5 SEP. SIZE	CASE 6 SEP. SIZE	CASE 7 SEP. SIZE
DIV 2,D-2				
FLO RAT	0.65000	0.55000	0.45000	0.35000
TOTAL FLOW				
FLO, LBMOL/HR	1374.0	1766.6	2159.1	2551.7

	CASE 8 SEP. SIZE	CASE 9 SEP. SIZE	CASE 10 SEP. SIZE
DIV 2,D-2			
FLO RAT	0.25000	0.15000	0.50000E-01
TOTAL FLOW			
FLO, LBMOL/HR	2944.3	3336.8	3729.4

Flash8.psd - Horizontal 3 Phase Separator With Boot

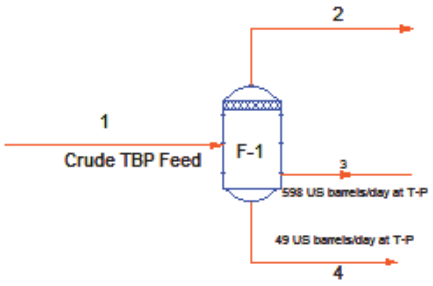
Sizing of a horizontal vessel for a specified residence time and boot diameter.

```
FLA 1 = F-1, 1,-2,-4,-3,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0
  HORI
  MIS ELI
  RES (MIN) = 6
  BOO DIA (M) = 0.75

****** SEPARATOR SIZING RESULTS FOR HORIZONTAL SEPARATOR    1 *******

+++ WARNING: NO IMMISCIBLE PHASE EXISTS.    BOOT IGNORED

  VAPOR STREAM =   2   LIQUID1 STREAM =   4   LIQUID2 STREAM =   3

+++ WARNING : REQUIRED WALL THICKNESS  1.1239 IN. NOT AVAILABLE
         IN  32.0 IN. NOMINAL PIPE.    PLATE STEEL USED INSTEAD.

CONDITIONS
----------
LIQUID   1  FLOW RATE         T-P         7.0133      M3 /HR
LIQUID   2  FLOW RATE         T-P       0.00000E+00   M3 /HR
LIQUID  TOT FLOW RATE         T-P         7.0133      M3 /HR
VAPOR       FLOW RATE         T-P         1399.5      M3 /HR
TOTAL       FLOW RATE         T-P         1406.5      M3 /HR
TEMPERATURE                               3.3333      DEG C
PRESSURE                                  60.001      BAR

PROPERTIES
----------
DENSITY                     LIQUID 1      564.77      KG/M3
DENSITY                     LIQUID 2    0.00000E+00   KG/M3
DENSITY                     VAPOR         58.076      KG/M3
Z-FACTOR                    VAPOR        0.81741

DESIGN PARAMETERS
-----------------
LENGTH TO DIAMETER RATIO    DEFAULT       4.0000
K  CONSTANT                 DEFAULT      0.10668      M/SEC
MIN LIQUID RESIDENCE TIME   SPEC          6.0000      MIN
MIN VAPOR SPACE HEIGHT      DEFAULT      0.38100      M
MIST ELIMINATOR             SPEC         0.15240      M
DESIGN PRESSURE             DEFAULT       66.001      BAR
ALLOWABLE STRESS            DEFAULT       1035.2      BAR
JOINT EFFICIENCY            DEFAULT       1.0000
CORROSION ALLOWANCE (CA)    DEFAULT     0.31750E-02   M
WEIGHT PERCENT ALLOWANCE(WA) DEFAULT      20.000

RESULTS
-------
MAX. DROPLET SETTLING VELOCITY            0.31511      M/SEC
ACTUAL  VAPOR VELOCITY                     1.5789      M/SEC
DIAMETER BASED ON SEPARATION              0.62823      M
SELECTED INTERNAL DIAMETER - PLATE STEEL  0.81280      M
LENGTH                                     3.6576      M
CORRESPONDING RESIDENCE TIME               8.5314      MIN
NORMAL LIQUID LEVEL                       0.42266      M
VAPOR SPACE HEIGHT                        0.39014      M
ACTUAL PLATE THICKNESS                  0.30163E-01    M
ACTUAL HEAD THICKNESS                   0.30163E-01    M
MIN SHELL THICKNESS INCLUDING CA        0.29686E-01    M
MIN HEAD  THICKNESS INCLUDING CA        0.28848E-01    M
APPROX WEIGHT OF STEEL VESSEL(EMPTY)       2433.6      KG
APPROX WEIGHT OF VESSEL (FULL)             3505.5      KG
APPROX TOTAL WEIGHT INCLUDING WA           3992.2      KG

NOZZLES (STD WALL)  INLET     VAP OUT    LIQ1 OUT   LIQ2 OUT
MINIMUM VELOCITY    9.402      9.604      3.080     0.0000E+00 M/SEC
ACTUAL  VELOCITY    12.11      12.04      3.494     0.0000E+00 M/SEC
MAXIMUM VELOCITY    15.67      16.01      5.133     0.0000E+00 M/SEC
NOMINAL DIAMETER    8.000      8.000      1.000      1.000     IN.

     ESTIMATED SONIC VELOCITY IN VAPOR IS  370.8 M/SEC
```



Flash9.psd – Crude Feed With Water Flash

This example shows a crude feed with water flash.



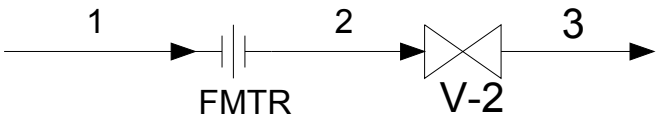

```
AB123.
*Crude Maker -

C- EQUIPMENT MODULES

FLA 1 = F-1, 1,-2,-3,-4,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0
  VERT
  MIS ELI
  SUR (MIN) = 5

C- CRUDE FEED CHARACTERIZATION
CRUDE
  FEE BLEND  1 = 1
  FEE VOL  1 = 0,5,10,20,30,40,50,60,70,80,100,
  FEE TBP (F) 1 = 50,100,175,200,250,300,350,400,450,500,550,
  FEE (BBL/DAY) 1 = 600
  FEE WAT (BBL/DAY) 1 = 50
  FEE GRA (API) 1 = 22.5
  FEE MOL VOL  1 = 0,5,10,20,30,40,50,60,70,80,90,100,
  FEE NAME  1 = Crude TBP Feed
  FEE REA () 1 = 0.12,2,1.9,
  FEE VIS TEM(F) = 100,210,
  FEE BUL VIS SET 1 (CST) STR 1 = 86
  FEE VIS VOL SET 1  STR 1 = 0,32,38,50,60,70,75,78,100,
  FEE VIS SET 1 (CST) STR 1 = 0,3.8,6,16,52,216,1000,2600,39000,
  FEE BUL VIS SET 2 (CST) STR 1 = 7.8
  FEE VIS VOL SET 2  STR 1 = 0,32,38,50,60,70,75,78,100,
  FEE VIS SET 2 (CST) STR 1 = 0,1.4,1.8,3.2,5.9,14,30,50,245,
  FEE BULK PRO SET 1  STR 1 = 0.94
  FEE PRO VOL SET 1  STR 1 = 0,5,6.5,10,21,30,40,50,61,71,81.7,100,
  FEE PRO SET 1 (WGT) STR 1 = Sulfur,0,2*0.002,0.004,0.032,0.16,0.46,0.81,0.94,
           1.11,1.57,2.38,
  FEE BULK PRO SET 2  STR 1 = 80.6
  FEE PRO VOL SET 2  STR 1 = 0,10,30,55,85,100,
  FEE PRO SET 2 (PPT) STR 1 = Pour Point,-68.8,-56.2,-40,89.6,104,111.2,


C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,2,3,4
  NAM POS 1 = 'WATER'
  NAM POS 2 = 'METHANE'
  NAM POS 3 = 'ETHANE'
  NAM POS 4 = 'PROPANE'
C-  Thermo Comments
  MOD ESSO
  APIMOD
  STDD
  DEN LIQ = STDD
  VIS VAP = MODAPI
  VIS LIQ = LNAV
  SUR TEN = STD
```

```
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  VIS UNI OUT = CST
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  REI PRO = 4,
  TP (F,PSIG) 1 = 80,1000
  NAM STR 1 = 'Crude TBP Feed'
  NAM STR 2 = 'Strm 2'
  NAM STR 3 = 'Strm 3'
  NAM STR 4 = 'Strm 4'
END
```

Chapter 29: Flow Meter Samples - c:\designii\samples\equipmnt\flowmetr

Fmtr1.psd - Vapor Phase Flow: Calculate Flowrate

Sizing a 303 stainless steel, flange tap orifice in a 3.97 inch carbon steel pipe to measure 12.5 MMSCF/DAY of methane vapor at 100 F and 500 psig. The pressure drop reading is 200 inches of water. Also, generating a table and a parametric plot with diameter ratio at 0.4, 0.45, 0.5 and range of differential pressure is 150 and 200 in H2O.




```
AB123.
*FLOW METER: CALCULATE FLOWRATE FOR A SINGLE PHASE

C- EQUIPMENT MODULES

FMTR 1 = FMTR, 1,-2,
     ORIfice
         TYPE= SQUared edged
         TAP = FLAnge
     PIPe WALl thickness code = STD
     INSide DIAmeter (IN) =3.97
     METhod of computation = API
     CALculate FLOw rate
     DELta pressure (INH2O) = 200
     RATio of the flow meter and pipe diameter =0.4
     PLOt RATio = 0.4,0.45,0.5
     RANge of DELta pressure (INH2O) = 150,200

VAL 2 = V-2, 2,-3,
  PRE OUT(PSIA) = 400.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  TIM UNI OUT = HR
  MAX = 1
  CON TOL = 1.E-003
  FLO(MMSCF/DAY)1 = 12.5,
  TP (F,PSIG) 1 = 100.,500.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
END
```

Simulation Results

```
****FLOWMETER           ****

           FLOW METER    1
         -------------------
EQUIPMENT NAME              =  FMTR
TYPE OF FLOW METER          =   ORI
TYPE OF ORIFICE             =   SQU
TYPE OF TAP                 =   FLA
METHOD FOR DIS. COEFFICIENT =   API
DIAMETER OF FLOW METER      =    0.1323     FT
DIAMETER OF PIPE            =    0.3308     FT

                       SPECIFIED        ACTUAL      UNITS
```

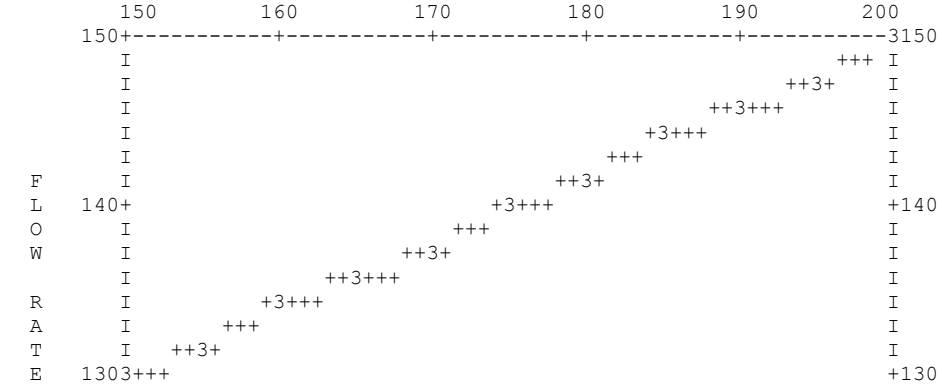
	CALCULATION	FLOWRATE	
FLOW RATE	9339.30	22021.9	LB/HR
	582.142	1372.68	LBMOL/HR
	220881.	520833.	SCF/HR
PRESSURE DROP	200.000	950.052	INH2O
VAPOR COEF DISCHARGE	0.600454	0.663601	
LIQUID COEF DISCHARGE	0.000000E+00	0.000000E+00	
VAPOR COEF EXPANSION	0.995760	0.974786	
LIQUID COEF EXPANSION	0.000000E+00	0.000000E+00	
STATIC/TOTAL P FACTOR	0.000000E+00	1.00596	
VAPOR REYNOLDS NUMBER	0.120791E+07	0.284822E+07	
LIQUID REYNOLDS NUMBER	0.000000E+00	0.000000E+00	
CRITICAL FLOWRATE	0.000000E+00	53055.9	LB/HR

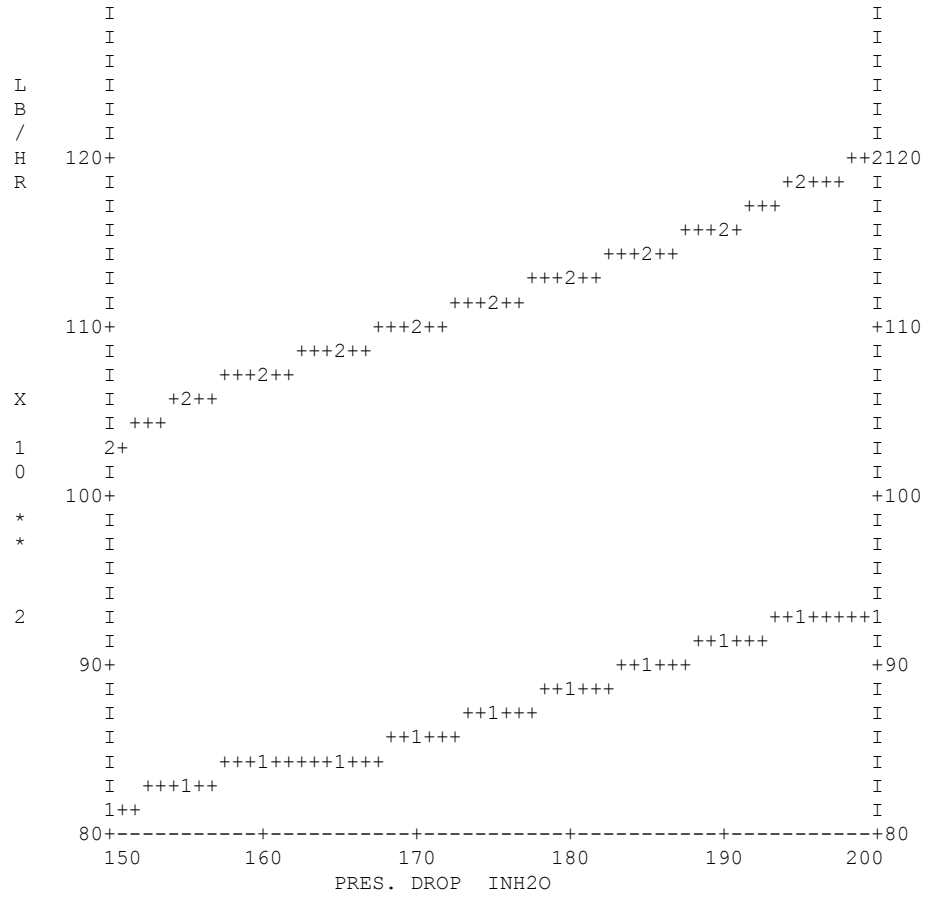
*** TABLE 1 ***

DIAMETER RATIO	PRES DROP INH2O	FLOW RATE LB/HR
0.4000	150.0	8098.
0.4000	155.0	8231.
0.4000	160.0	8361.
0.4000	165.0	8490.
0.4000	170.0	8616.
0.4000	175.0	8741.
0.4000	180.0	8864.
0.4000	185.0	8985.
0.4000	190.0	9105.
0.4000	195.0	9223.
0.4000	200.0	9339.
0.4500	150.0	0.1035E+05
0.4500	155.0	0.1052E+05
0.4500	160.0	0.1069E+05
0.4500	165.0	0.1085E+05
0.4500	170.0	0.1101E+05
0.4500	175.0	0.1117E+05
0.4500	180.0	0.1133E+05
0.4500	185.0	0.1149E+05
0.4500	190.0	0.1164E+05
0.4500	195.0	0.1179E+05
0.4500	200.0	0.1194E+05
0.5000	150.0	0.1295E+05
0.5000	155.0	0.1317E+05
0.5000	160.0	0.1337E+05
0.5000	165.0	0.1358E+05
0.5000	170.0	0.1378E+05
0.5000	175.0	0.1398E+05
0.5000	180.0	0.1418E+05
0.5000	185.0	0.1437E+05
0.5000	190.0	0.1456E+05
0.5000	195.0	0.1475E+05
0.5000	200.0	0.1494E+05

* The SPECIFIED data are the parameters defined in the FMTR input. Pressure drop is across the orifice and does not include pressure recovery.
** The actual data is the inlet stream flow rate and the pressure drop including pressure recovery.

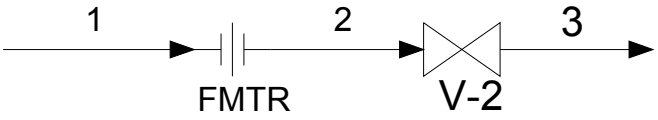
*** PLOT 1 ***





Fmtr2.psd - Two Phase Flow: Calculate Size

Sizing a 303 stainless steel, corner tap orifice in a 4 inch carbon steel pipe to calculate the size of the flowmeter given the flowrate of liquid hydrocarbons at 150 F and 869 psig and the pressure drop of 2.5 inches of water. Also, generating a table and a parametric plot with diameter ratio at 0.4, 0.45, 0.5 and range of differential pressure is 150 and 200 in H2O.

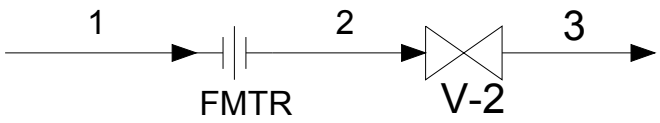



```
AB123.
*TWO-PHASE FLOW METER: CALCULATE SIZE

C- EQUIPMENT MODULES

FMTR 1 = FMTR, 1,-2,
     ORIfice
         TYPE= SQUared edged
         TAP =CORner
     PIPe WALl thickness code = STD
     INSide DIAmeter (IN) =4
     CALculate SIZe
     FLOw RATe (LBMOL/HR) =327
     DELta pressure (INH2O) =2.5
     PLOt RATio = 0.4,0.45,0.5
     RANge of DELta pressure (INH2O) = 150,200

VAL 2 = V-2, 2,-3,
  PRE OUT(PSIA) = 500.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,2,3,4,5,6,7,8,10,
```

```
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  QUA UNI OUT = GAL
  TIM UNI OUT = HR
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 2.9E-002,0.19236,0.1763,0.2237,7.66E-002,0.1203,5.26E-002,
  4.77E-002,0.1101,
  TOT FLO (LBMOL/HR) 1 = 100.
  TP (F,PSIG) 1 = 150.,869.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
END
```

Fmtr3.psd - Liquid Phase Flow: Calculate Delta Pressure

Sizing a 303 stainless steel, square-edged radius tap orifice in a 2.5 inch carbon steel pipe to calculate the delta pressure of water at 483 F and 580 psig. The flowrate of water is 10,320 lb/hr and the diameter ratio of flowmeter and pipe diameter at 0.50. Also, generating a table and a parametric plot with diameter ratio at 0.4, 0.45, 0.5 and range of differential pressure is 150 and 200 in H2O.




```
AB123.
*FLOW METER: CALCULATE DELTA PRESSURE

C- EQUIPMENT MODULES

FMTR 1 = FMTR, 1,-2,
     ORIfice
         TYPE= SQUared edged
         TAP =RADius
     PIPe WALl thickness code = STD
     INSide DIAmeter (IN) =2.5
     CALculate delta PREssure
     FLOw RATe (LB/HR) =10500
     RATio of the flow meter and pipe diameter =0.5
     PLOt RATio = 0.4,0.45,0.5
     RANge of DELta pressure (INH2O) = 150,200

VAL 2 = V-2, 2,-3,
  PRE OUT(PSIA) = 500.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  STE TAB
  QUA UNI OUT = GAL
  TIM UNI OUT = HR
  MAX = 1
  CON TOL = 1.E-003
  FLO(LB/HR)1 = 10320.,
  TP (F,PSIA) 1 = 483.,580.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
STEAM TABLES
END
```

Chapter 30: Heat Exchanger Samples - c:\designii\samples\equipmnt\heaexc

Exchgr1.psd - Single Stream Specified Outlet Temperature

An air cooler is used to cool a process stream to 120F. The pressure drop is 5 PSIA. All properties of stream 17 are known. The pressure and temperature are adjusted and a new enthalpy calculated for stream 18. The change in enthalpy from stream 17 to stream 18 is reported as the duty.

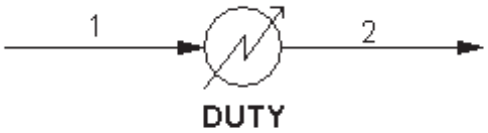
```
AB123.
*Example of Single Stream Heat Exchanger 1

C- EQUIPMENT MODULES

HEA EXC 23 = DELP=5 PSI, 17,-18,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 120.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  QUA UNI OUT = LB
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)17 = 800.,2*100.,
  TP (F,PSIA) 17 = 200.,290.
  NAM STR 17 = Strm 17
  NAM STR 18 = 120 F
END
```

Exchgr2.psd - Single Stream Specified Duty

Hot oil is used as the reboiler heating material for a column, and a heating curve on the column bottoms is required. The cooling curve on the hot oil should be run separately since it is probably used for other applications with various temperature ranges, and only one curve through the entire range would be run. The reboiler duty is 50 MMBTU/HR.

All properties of stream 1 are known. The duty is added to the enthalpy of stream 1 and an outlet temperature is calculated. The temperature change is broken into 15 increments and a report similar to the one shown in Output Control is generated.

```
AB123.
*Example of Heat Exchanger 2: Specified Duty

C- EQUIPMENT MODULES

HEA EXC 23 = DUTY, 1,-2,
  DUT(BTU/HR) = 50000000.
  U(BTU/HR/FT2/F) = 50.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5.

C- GENERAL SECTION
GENERAL
```

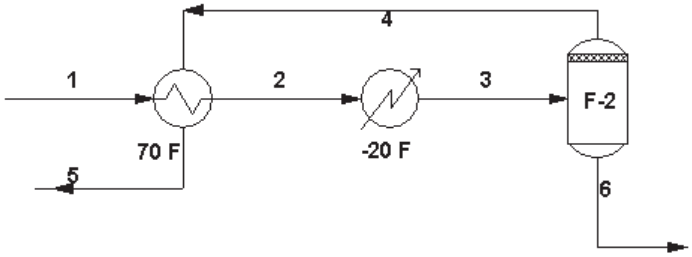


```
  COMPONENTS = 100,
  NAM 100 = HOT_OIL (Pet-100)
  AMW 100 = 245.
  AMB (F) 100 = 300.
  API (API) 100 = 45.
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  QUA UNI OUT = LB
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  FLO(GAL/MIN)1 = 300.,
  TP (F,PSIA) 1 = 100.,50.
  NAM STR 1 = Strm 1
  HEA CUR 1
  NAM STR 2 = STRM 2
END
```

Exchgr3.psd - Counter-Current Specified Outlet Temperature

To calculate equipment 1 you need to know stream 5, which will not be known until equipment 3 is calculated. This is a recycle problem, and you must make a guess for one of the streams. Guess stream 3 and code equipment 1. Stream 3, the feed stream to equipment 3, has the same flowrate and composition as stream 1. Therefore enter FLO 3 = FLO 1 in the GENeral section. Assuming the other properties for streams 1 and 3 are known, enter corresponding TP and FLO commands to define the streams. If equipment 3 is calculated first, stream 5 will be known when equipment 1 is calculated and each equipment will be calculated only once. This leads to a quicker solution than guessing stream 5 and looping until convergence.

```
AB123.
*Example of Heat Exchanger 3: Specified Outlet Temperature

C- EQUIPMENT MODULES

HEA EXC 1 = 70 F, 4,1,-5,-2,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 70.
  Q-T
  MIN TEM APP(F) = 10.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,

FLA 2 = F-2, 3,-4,-6,
  ADI, HEA(BTU/HR) = 0.
  DEL(PSI) = 0.

HEA EXC 23 = -20 F, 2,-3,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = -20.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  REC STR = 3,
  FLO(LBMOL/HR)1 = 200.,100.,50.,
```



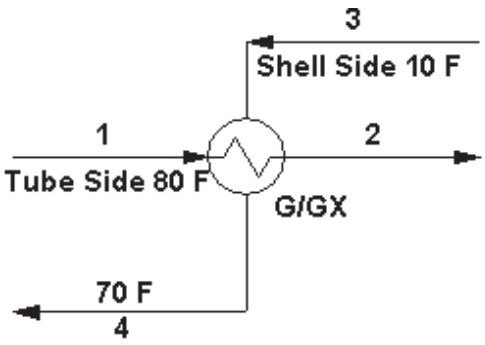
```
  TP (F,PSIA) 1 = 100.,250.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 200.,100.,50.,
  TP (F,PSIA) 3 = -20.,250.
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
END
```

Exchgr4.psd - Counter-Current Outlet Temperature Specified

All properties of the inlet streams are known. The temperature and pressure of stream 4 are changed and the enthalpy calculated for stream 4. The change in enthalpy from stream 3 to stream 4 is applied to stream 1, and using the enthalpy, the temperature of stream 2 is calculated. The corrected LMTD is calculated from the known temperatures in and out. The area is then calculated from the equation: A = Q/(U*LMTD) where Q is the duty or enthalpy change.

NOTE: The TEMperature OUT specification for stream 4 will be met whether or not duty is available from stream 1.

```
AB123.
*Example of Heat Exchanger 4: Outlet Temperature Specified

C- EQUIPMENT MODULES

HEA EXC 1 = G/GX, 3,1,-4,-2,
  U(BTU/HR/FT2/F) = 80.
  TEM OUT(F) = 70.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 8., 5.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 200.,100.,25.,0.,
  TP (F,PSIA) 1 = 80.,290.
  NAM STR 1 = Tube Side 80 F
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 3*0.,100.,
  TP (F,PSIA) 3 = 10.,50.
  NAM STR 3 = Shell Side 10 F
  NAM STR 4 = 70 F
END
```



Exchgr5.psd - Counter-Current Specified Approach Temperature

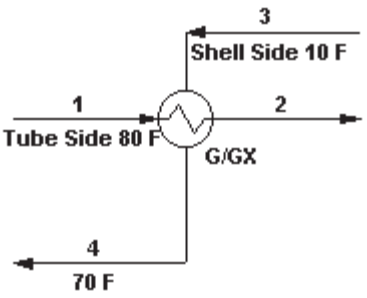
The TEM APP command description tells you to enter the absolute value of the temperature difference between the stream coded as the *first outlet stream* and the stream coded as the second input stream. You must code the stream order as 1, 3, -2, -4 since you want to set the approach between stream 8 and stream 1. If the temperature approach cannot be met (controlling approach is on the other side of the exchanger, i.e., between stream and 20) the approach will be applied to the other side. The temperature approach must be greater than 2F (1.1K).

Method of Calculation

The temperature approach is subtracted from the temperature of stream 8 to obtain the temperature of stream 21. The calculation is the same as described in the Temperature Out Specification section of the COMMAND DETAILS.

```
AB123.
*Example of Heat Exchanger 5: Specified Approach Temperature
```

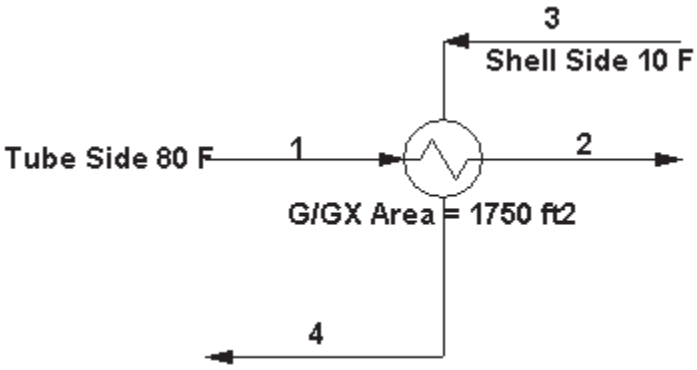
```
C- EQUIPMENT MODULES

HEA EXC 1 = G/GX, 3,1,-4,-2,
  U(BTU/HR/FT2/F) = 80.
  TEM APP(F) = 10.
  SHE PAS = 2
  TUB PAS = 4
  DEL(PSI) = 8., 5.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,62,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 200.,100.,25.,0.,
  TP (F,PSIA) 1 = 80.,290.
  NAM STR 1 = Tube Side 80 F
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 3*0.,100.,
  TP (F,PSIA) 3 = 10.,50.
  NAM STR 3 = Shell Side 10 F
  NAM STR 4 = 70 F
END
```



Exchgr6.psd - Counter-Current Simple Exchanger With Specified Area

A guess is made for the first outlet stream temperature and the calculation, as described under the Temperature Out Specification, is performed. The enthalpy change must balance with the U*Area*LMTD product. If the balance is not met, additional temperatures are calculated and tried until convergence is obtained.

```
AB123.
*Example of Heat Exch 6: Specified Area
C- EQUIPMENT MODULES

HEA EXC 1 = G/GX Area = 1750 ft2, 3,1,-4,-2,
  SIM
  U(BTU/HR/FT2/F) = 80.
  ARE(FT2) = 1750.
  SHE PAS = 2
  TUB PAS = 4
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 8., 5.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 200.,100.,25.,0.,
  TP (F,PSIA) 1 = 80.,290.
  NAM STR 1 = Tube Side 80 F
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 3*0.,100.,
  TP (F,PSIA) 3 = 10.,50.
  NAM STR 3 = Shell Side 10 F
  NAM STR 4 = Strm 4
END
```

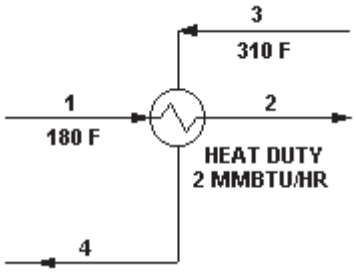


Exchgr7.psd - Counter-Current Specified Duty

The duty specification is normally used in simulating reboilers or condensers associated with column calculations and is done in a separate run from the column.

The duty is added to the enthalpy of the first stream coded and the outlet temperature is calculated. The calculations proceed as described in the Temperature Out Specification section of the COMMAND DETAILS. Heating curves with 10 temperature increments are also provided.

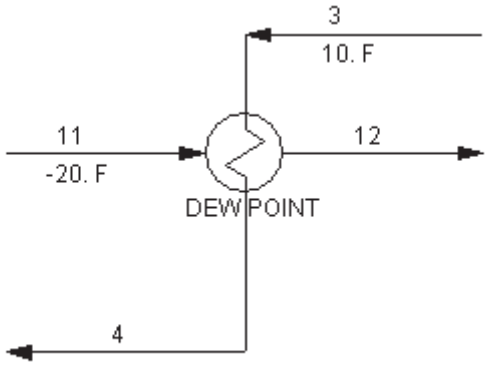
```
AB123.
*Example of Heat Exchanger 7: Specified Duty

C- EQUIPMENT MODULES

HEA EXC 1 = HEAT DUTY, 3,1,-4,-2,
  ARE(FT2) = 1750.
  DUT(BTU/HR) = 2000000.
  U(BTU/HR/FT2/F) = 80.
  SHE PAS = 2
  TUB PAS = 4
  DEL(PSI) = 8., 5.,
     CURve INCrements =10

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,7,8,9,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 0.,200.,100.,50.,
  TP (F,PSIA) 1 = 180.,50.
  NAM STR 1 = 180 F
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 100.,3*0.,
  TP (F,PSIA) 3 = 310.,50.
  NAM STR 3 = 310 F
  NAM STR 4 = Strm 4
END
```



Exchgr8.psd - Counter-Current Specified Temperature Out Is Dew Point

All properties of the inlet streams are known. The dew point temperature for stream 12 is calculated. The enthalpy of stream 12 is calculated for the dew point vapor and the change in enthalpy from stream 11 calculated. The enthalpy change is applied to stream 3 to calculate the enthalpy of stream 4. The pressure drop is used to adjust the pressure of stream 4 and the temperature is calculated. The corrected LMTD can now be calculated and then the area.

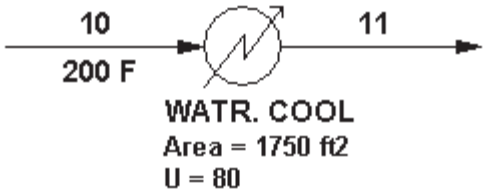
```
AB123.
*Example of Heat Exchanger 8: Specified Temperature Out is Dew

C- EQUIPMENT MODULES

HEA EXC 1 = DEW PT., 3,11,-4,-12,
  U(BTU/HR/FT2/F) = 80.
  TEM OUT DEW
  SHE PAS = 2
  TUB PAS = 4
  DEL(PSI) = 8., 5.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,62,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)3 = 3*0.,100.,
  TP (F,PSIA) 3 = 10.,50.
  NAM STR 3 = 10 F
  NAM STR 4 = 70 F
  FLO(LBMOL/HR)11 = 200.,100.,25.,0.,
  TP (F,PSIA) 11 = -20.,290.
  NAM STR 11 = - 20 F
  NAM STR 12 = Strm 2
END
```



Exchgr9.psd - Water Cooled Outlet Temperature Calculated

The temperature out for the process stream is guessed, and the duty is calculated. The corrected LMTD is calculated and the product of U*AREa*LMTD is checked against the duty (Q = U*A*LMTD). Subsequent guesses for the process stream outlet temperature are calculated by the program, and the above procedure followed until convergence is obtained. Then, knowing the duty, the amount of water can be calculated.

```
AB123.
*Example of Heat Exchanger 9: Water Cooled.

C- EQUIPMENT MODULES

HEA EXC 2 = WATR. COOL, 10,-11,
  U(BTU/HR/FT2/F) = 80.
  ARE(FT2) = 1750.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.
  WATCOO
  TEM WAT IN(F) = 90.
  TEM WAT OUT(F) = 105.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)10 = 200.,100.,25.,
  TP (F,PSIA) 10 = 200.,290.
  NAM STR 10 = 200 F
  NAM STR 11 = Strm 2
END
```

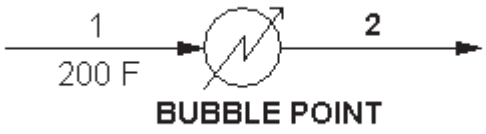


NOTE: Water streams need not be numbered.

Exchgr10.psd - Water Cooled Condenser Bubble Point Calculated

The bubble point temperature of the process stream is calculated, allowing the enthalpy change or duty to be calculated. The amount of water can then be calculated from the duty required. AREa will be calculated from A = (Q/U*LMTD)

```
AB123.
*Example of Heat Exchanger 10: Water Cooled.

C- EQUIPMENT MODULES

HEA EXC 2 = BUBBLE POINT, 1,-2,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT BUB
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5.
  TEM WAT IN(F) = 80.
  TEM WAT OUT(F) = 100.
  WAT COO CON

C- GENERAL SECTION
GENERAL
  COMPONENTS = 4,5,6,7,8,62,
  APISOAVEK
  APISOAVEH
  STDD
```

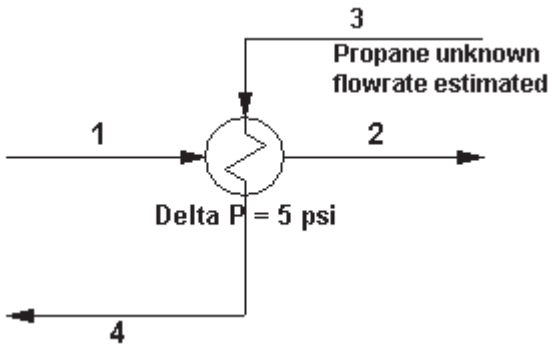


```
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 920.,184.,271.,3.5,0.9,0.,
  TP (F,PSIA) 1 = 200.,200.
  NAM STR 1 = 200 F
  NAM STR 2 = Strm 2
END
```

Exchgr11.psd - Single Component Refrigerant Enthalpy Calculated

The enthalpy of the process stream out is calculated from the specified temperature. The enthalpy change (duty) is then used to calculate the required refrigerant flow. The bubble point pressure of the propane (component id 4) refrigerant at 10F will also be calculated.

```
AB123.
*Example of Heat Exch 11: Enthalpy Calculated

C- EQUIPMENT MODULES

HEA EXC 2 = Delta P = 5 psi, 1,3,-2,-4,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 20.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,
  TEM REF(F)4 = 10.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 200.,100.,25.,0.,
  TP (F,PSIA) 1 = 90.,290.
  NAM STR 1 = STRM 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
END
```

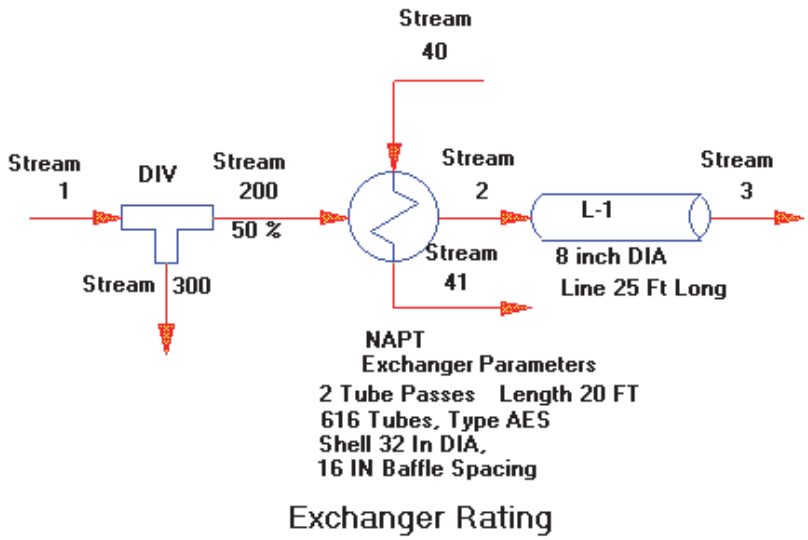


NOTE: A guess for the pressure of the refrigerant should be coded in the GENeral command section in a TP command. (e.g. TP3 = 10,50)

Exchgr12.psd - Single Phase Rating

A counter-current exchanger is used to transfer heat between a crude feed and product stream.

```
AB123.
*EXCHANGER 12:  RATING CALCULATION

C- EQUIPMENT MODULES

HEA EXC 1 = NAPH, 40,200,-41,-2,
  SIM
  U(BTU/HR/FT2/F) = 50.
  ARE(FT2) = 2419.
  SHE PAS = 1
  TUB PAS = 2
  SHE = 1
```



```
  TUB FEE = 1
  INL = OPP
  HOR
  RAT
  NUM INC = 5
  MET = BELL
  EXC = AES
  TOL = 1.E-006
  NUM TUB = 616
  TUB LAY = SQU
  TUB MAT = ADM
  TUB LEN(FT) = 20.
  TUB PIT(IN) = 1.
  TUB FOU(1/BTU/FT2/HR/F) = 3.E-003
  TUB INS DIA(IN) = 0.584
  TUB BWG = 16
  BAF CUT = 0.25
  NUM SEA = 2
  BAF SPA(IN) = 16.
  SHE FOU(1/BTU/FT2/HR/F) = 0.
  SHE DIA(IN) = 32.
  DEL(PSI) = 0.6, 3.,

LINE 2 = L-1, 2,-3,
  DIA (INCH) = 8
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 25. FT
  LEN (FT) = 25.,
  SEG LEN (FT) = 2.5
  SEG ELE(FT) = 20.
  ISO

DIV 100 = Divider, 1,-200,-300,
  FLO(FRA) = 0.5

C- CRUDE FEED CHARACTERIZATION
CRUDE
  TEM BRE(F) = 500., 900.,
  TEM INC(F) = 30., 60., 120.,
  FEE VOL  1 = 2.,5.,10.,20.,30.,40.,50.,60.,70.,80.,90.,95.,98.,100.,
  FEE TBP (F) 1 = -50.,100.,200.,300.,400.,470.,550.,650.,750.,850.,1100.,1300.,
  1475.,1670.,
  FEE (BBL/DAY) 1 = 50000.
  FEE GRA VOL  1 = 2.,5.,10.,20.,30.,40.,50.,60.,70.,80.,90.,95.,98.,
  FEE GRA (API) 1 = 150.,95.,65.,45.,40.,38.,33.,30.,25.,20.,15.,10.,5.,
  FEE NAME  1 = STRM 1
  FEE REA () 1 = 0.1,0.2,2*0.5,1.,1.5,2.5,
  FEE VOL  40 = 0.,5.,10.,30.,50.,70.,90.,95.,100.,
  FEE TBP (F) 40 = 65.,94.,120.,192.,232.,268.,317.,339.,371.,
  FEE (BBL/DAY) 40 = 8020.
  FEE GRA (API) 40 = 60.5
  FEE NAME  40 = STRM 4

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,2,3,4,5,6,7,8,
  ESSO
  GRA
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  TP (F,PSIA) 1 = 90.,150.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  TP (F,PSIA) 40 = 217.,100.
  NAM STR 40 = Strm 4
  NAM STR 41 = Strm 41
  NAM STR 200 = 50%
  NAM STR 300 = Strm 4
PET STR=NONE
END
```

A partial output for Example 12 is shown on the following page.

```
*** PROCESS RATING - HEAT EXCHANGER    1 = NAPH
================================================
 SIZE   21-240      TYPE  AES
 ------------------------------------------------------------------------------
                                I     SHELL SIDE      I     TUBE SIDE
 ------------------------------------------------------------------------------
 FEED STREAMS                   I   40                I  200
 PRODUCT STREAMS                I   41                I    2
 TEMPERATURE (IN/OUT)  DEG F    I   217.000/   130.330I    90.000/   118.400
 INLET PRESSURE        PSIA     I   100.00            I   150.00
 VELOCITY(IN/OUT)      FT/SEC   I   0.75073/   0.70384I     2.924/     2.963
 HEAT EXCHANGED        BTU/HR   I   -4.226943E+06     I    4.226944E+06
 FOULING FACTOR    HR-FT2-F/BTUI    2.000000E-03     I    3.000000E-03
 HEAT TRANS. COEF. BTU/HR/FT2/FI     163.989         I     58.6589
 PRES DROP SPEC/CALC   PSI      I   0.60000/   0.55739I     3.000/     2.501
 TOTAL FLUID           LB/HR    I  86551.7            I  310433.
       LIQUID      (IN/OUT)     I 86551.711/ 86551.711I 3.104E+05/ 3.104E+05
 PROPERTIES                     I                     I
 ----------                     I                     I
 SP.GRAVITY AT 60 F  DEG API    I      59.799         I      34.760
 MOLECULAR WEIGHT  (IN/OUT)     I                     I
       LIQUID                   I   104.733/   104.733I   202.186/   202.186
 VISCOSITY          (IN/OUT)    I                     I
       LIQUID      CP           I   0.22496/   0.30010I     4.169/     3.010
 DENSITY            (IN/OUT)    I                     I
       LIQUID      LB/FT3       I    41.352/    44.107I    51.476/    50.803
 SPECIFIC HEAT      (IN/OUT)    I                     I
       LIQUID      BTU/LB/F     I   0.58619/   0.54028I   0.47100/   0.48785
 THE. CONDUCTIVITY (IN/OUT)     I                     I
       LIQUID      BTU/FT/HR/F I   0.07109/   0.07719I   0.07768/   0.07531
 ------------------------------------------------------------------------------
 HEAT EXCHANGED       BTU/HR   =  4.226943E+06
 OVERALL U COEFF. BTU/HR/FT2/F =    30.320
 MTD  (= Q/UA)        DEG F    =    57.633
 CORRECTION FACTOR             =     0.88420
 NO. OF SHELL PASSES/SHELL     =     1
 NO. OF SHELLS IN PARALLEL     =     1
 SURFACE AREA SPEC/CALC FT2    =   2419.0    /  2302.9
 EST. REQUIRED AREA/SHELL FT2  =   2419.0
 METHOD OF CALCULATION         =  BELL
 TUBE MATERIAL                 =  CARBON STEEL
 ------------------------------------------------------------------------------
       CONSTRUCTION OF ONE SHELL
 ------------------------------------------------------------------------------
 SHELL SIDE                          TUBE SIDE
 ==========                          =========
 INSIDE DIAMETER  IN   =    21.250   OUTSIDE DIAMETER IN  =     0.714
 BAFFLE SPACING   IN   =    16.000   PITCH SQUARE     IN  =     1.000
 TUBE BUNDLE DIA. IN   =    20.850   LENGTH           FT  =    20.000
 BAFFLE-CLEARANCE IN   =     0.150   INSIDE DIAMETER  IN  =     0.584
 BAFFLE CUT            =     0.250   NO. OF PASSES        =     2
 SEALING STRIPS        =     2       NO. OF TUBES         =   616
 ------------------------------------------------------------------------------
```

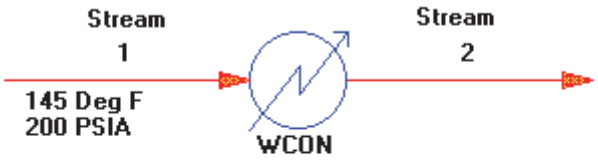
Exchgr13.psd - Two Phase Rating

A water cooled heat exchanger is used to condense the vapor from a debutanizer. In this example, the WATer COOled CONdenser option is used to calculate the required cooling water before performing the rating calculation.

All properties of stream 1 are known. Also, the geometric configuration of the exchanger is known. The WATer COOled CONdenser tells DESIGN II to condense the inlet gas to its bubble point (i.e., a TEM OUT specification). It also tells DESIGN II to use cooling water with default temperatures of 90 F in and 105 F out. The printed REQUIRED AREA is the amount needed to meet the TEM OUT or DUTy specification. It is estimated with the U value calculated for the actual surface area (based on user specified tube geometry).

```
AB123.
*Example of Heat Exchanger 13: Water Cooled Rating.

C- EQUIPMENT MODULES
```

```
HEA EXC 2 = WCON, 10,-11,
  ARE(FT2) = 0.
  U(BTU/HR/FT2/F) = 80.
  TEM OUT BUB
  SHE PAS = 1
  TUB PAS = 2
  DEL(PSI) = 0.
  TEM WAT IN(F) = 90.
  TEM WAT OUT(F) = 105.
   RATe,
   EXChanger type = AES,
   INLet = OPPosite,
   TUBe PITch(IN) = 1,
   NUMber of TUBes = 811,
   SHEll DIAmeter(IN) = 35,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 4,5,6,7,8,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)10 = 920.,184.,271.,3.5,0.9,
  TP (F,PSIA) 10 = 145.,200.
  NAM STR 10 = 145 F
  NAM STR 11 = Strm 2
END
```



2 Tube Passes, Type AES
Tube Pitch 1, 811 Tubes
Shell Diameter 35 Inches

Water Cooled Condensor

```
*** PROCESS RATING - HEAT EXCHANGER    2 = WCON
================================================
 SIZE   35-192      TYPE  AES
 ------------------------------------------------------------------------------
                               I     SHELL SIDE      I     TUBE SIDE
 ------------------------------------------------------------------------------
 FEED STREAMS                  I   10                I
 PRODUCT STREAMS               I   11                I
 TEMPERATURE (IN/OUT)  DEG F   I   145.000/   124.425I    90.000/   105.000
 INLET PRESSURE        PSIA    I   200.00            I   14.696
 VELOCITY(IN/OUT)      FT/SEC  I    23.260/     1.485I     3.288/     3.298
 HEAT EXCHANGED        BTU/HR  I   -9.328335E+06     I    9.328336E+06
 FOULING FACTOR    HR-FT2-F/BTUI    0.000000E+00     I    0.000000E+00
 HEAT TRANS. COEF. BTU/HR/FT2/FI     372.416         I     971.021
 PRES DROP SPEC/CALC   PSI     I   0.00000/     5.269I   0.00000/   0.00000
 TOTAL FLUID           LB/HR   I  67328.5           I  624301.
       VAPOR       (IN/OUT)    I 67328.516/   0.00000I   0.00000/   0.00000
       LIQUID      (IN/OUT)    I   0.00000/ 67328.516I 6.243E+05/ 6.243E+05
 PROPERTIES                    I                     I
 ----------                    I                     I
 SP.GRAVITY AT 60 F  DEG API   I      134.09         I      10.000
 MOLECULAR WEIGHT  (IN/OUT)    I                     I
       VAPOR                   I    48.810/   0.00000I   0.00000/   0.00000
       LIQUID                  I   0.00000/    48.810I    18.015/    18.015
 VISCOSITY         (IN/OUT)    I                     I
       VAPOR       CP          I   0.00971/   0.00000I   0.00000/   0.00000
       LIQUID      CP          I   0.00000/   0.08466I   0.76057/   0.64459
 DENSITY           (IN/OUT)    I                     I
       VAPOR       LB/FT3      I     1.890/   0.00000I   0.00000/   0.00000
       LIQUID      LB/FT3      I   0.00000/    29.617I    62.035/    61.846
 SPECIFIC HEAT     (IN/OUT)    I                     I
       VAPOR       BTU/LB/F    I   0.50793/   0.00000I   0.00000/   0.00000
       LIQUID      BTU/LB/F    I   0.00000/   0.73385I     1.002/     1.002
 THE. CONDUCTIVITY (IN/OUT)    I                     I
       VAPOR       BTU/FT/HR/F I   0.01418/   0.00000I   0.00000/   0.00000
       LIQUID      BTU/FT/HR/F I   0.00000/   0.05090I   0.35818/   0.36405
 ------------------------------------------------------------------------------
 HEAT EXCHANGED       BTU/HR   =  9.328335E+06
 OVERALL U COEFF. BTU/HR/FT2/F =   239.647
 MTD  (= Q/UA)        DEG F    =    35.143
 CORRECTION FACTOR             =     0.94616
 NO. OF SHELL PASSES/SHELL     =     1
 NO. OF SHELLS IN PARALLEL     =     1
 SURFACE AREA SPEC/CALC FT2    =  0.00000E+00/  2547.8
 EST. REQUIRED AREA/SHELL FT2  =   1107.6
 TUBE MATERIAL                 =  CARBON STEEL
 ------------------------------------------------------------------------------
```

```
      CONSTRUCTION OF ONE SHELL
--------------------------------------------------------------------------
SHELL SIDE                          TUBE SIDE
==========                          =========
INSIDE DIAMETER  IN   =    35.000   OUTSIDE DIAMETER IN  =     0.750
BAFFLE SPACING   IN   =     7.000   PITCH SQUARE     IN  =     1.000
TUBE BUNDLE DIA. IN   =    34.600   LENGTH           FT  =    16.000
BAFFLE-CLEARANCE IN   =     0.175   INSIDE DIAMETER  IN  =     0.620
BAFFLE CUT            =     0.200   NO. OF PASSES        =     2
SEALING STRIPS        =     2       NO. OF TUBES         =   811
--------------------------------------------------------------------------

           TEMPERATURE PROFILES IN HEAT EXCHANGER    2

       0.0       1.0        2.0        3.0        4.0       5.0
  150.0+---------+---------+---------+---------+---------+150.0
       I                                                   I
       I                                                   I
       I                                                   I
       S                                S SHELLSIDE        I
       I**                              * INTERPOLATED     I
       I  **                            T TUBESIDE         I
       I    **                          + INTERPOLATED     I
  140.0+      **                                           +140.0
       I        **                                         I
       I          S*                                       I
       I            ***                                    I
       I               **                                  I
       I                 ***                               I
T      I                    *S*                            I
E      I                       ****                        I
M 130.0+                           ****                    +130.0
P      I                               *S**                I
E      I                                   ******          I
R      I                                         **S*****  I
A      I                                                *****S
T      I                                                   I
U      I                                                   I
R      I                                                   I
E 120.0+                                                   +120.0
       I                                                   I
D      I                                                   I
E      I                                                   I
G      I                                                   I
       I                                                   I
F      I                                                   I
       I                                                   I
  110.0+                                                   +110.0
       I                                                   I
       I                                                   I
       I                                                   I
       I                                                   I
       I                                  +++++T++++++++++T
       I                             +++++T+++++           I
       I                  +++++T+++++                      I
  100.0+        ++T+++++                                   +100.0
       I  ++++++                                           I
       T++                                                 I
       I  ++++++                                           I
       I        ++T+++++                                   I
       I                 +++++T+++++                       I
       I                             +++++T+++++           I
       I                                        +++++T+++++I
   90.0+---------+---------+---------+---------+---------+T90.0
       0.0       1.0        2.0        3.0        4.0       5.0
                             INCREMENT
```

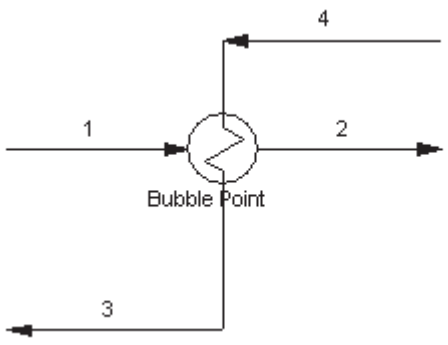
Exchgr14.psd - Counter-Current Rating

This example is similar to the previous example except the exchanger is coded as a counter-current exchanger with the water rate specified in the GENeral section.

All properties of stream 1 are known. Also, the geometric configuration of the exchanger is known. The TEMperature OUT is BUBble tells DESIGN II to condense the inlet gas to its bubble point (i.e. a TEM OUT specification). Cooling water is used on the tube side of the exchanger with the temperature of the inlet water set by the TP command in the general section. The

printed REQUIRED AREA is the amount needed to meet the TEM OUT BUB specification. It is estimated with the U value calculated for the actual surface area (based on user specified tube geometry).

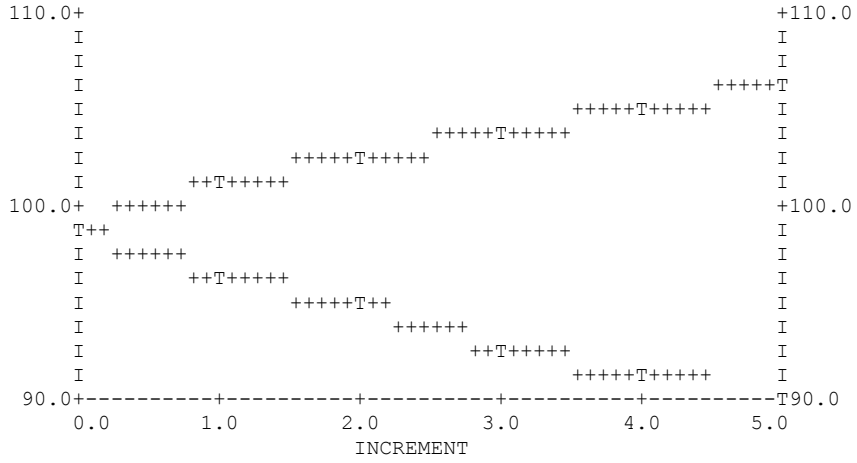
```
AB123.
*Example of Heat Exchanger 14: Exchanger Rating

C- EQUIPMENT MODULES

HEA EXC 1 = Bubble Point, 1,4,-2,-3,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT BUB
  SHE PAS = 1
  TUB PAS = 2
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  RAT
  NUM INC = 5
  MET = KERN
  EXC = AES
  TOL = 1.E-005
  NUM TUB = 811
  TUB LAY = SQU
  TUB MAT = STA
  TUB LEN(FT) = 16.
  TUB PIT(IN) = 1.
  TUB FOU(1/BTU/FT2/HR/F) = 1.E-003
  TUB OUT DIA(FT) = 6.25E-002
  TUB BWG = 16
  BAF CUT = 0.2
  NUM SEA = 2
  BAF SPA(IN) = 7.
  SHE FOU(1/BTU/FT2/HR/F) = 1.E-003
  SHE DIA(IN) = 35.
  DEL(PSI) = 0., 0.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 4,5,6,7,8,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  STE TAB
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 920.,184.,271.,3.5,0.9,0.,
  TP (F,PSIA) 1 = 145.,200.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  FLO(GAL/MIN)4 = 5*0.,1200.,
  TP (F,PSIA) 4 = 90.,55.
  NAM STR 4 = Strm 4
END
```



****PROCESS RATING RESULTS FOR HEAT EXCHANGER*****

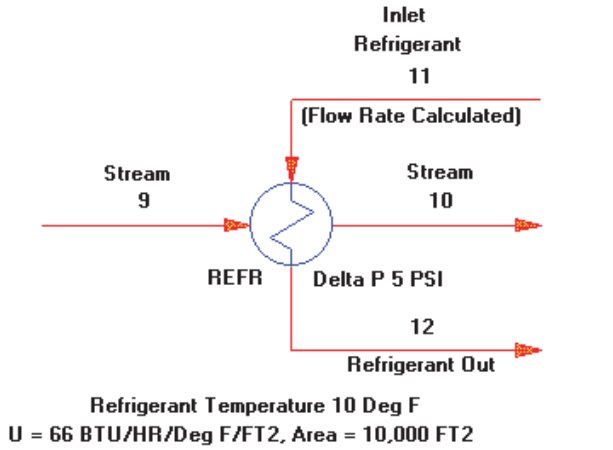
```
Example of Heat Exchanger 14: Exchanger Rating                        PAGE   10
                          SOLUTION REACHED
*** PROCESS RATING - HEAT EXCHANGER    2 = BUB POINT
=========================================================
 SIZE   35-192      TYPE  AES
 ------------------------------------------------------------------------
                               I     SHELL SIDE      I     TUBE SIDE
 ------------------------------------------------------------------------
 FEED STREAMS                  I    1                I    3
 PRODUCT STREAMS               I    2                I    4
 TEMPERATURE (IN/OUT)  DEG F   I   145.000/   124.429I    90.000/   105.704
 INLET PRESSURE        PSIA    I   200.00           I   55.000
 VELOCITY(IN/OUT)      FT/SEC  I    23.260/     1.485I     3.103/     3.127
 HEAT EXCHANGED        BTU/HR  I   -9.328196E+06     I    9.328192E+06
 FOULING FACTOR    HR-FT2-F/BTUI    0.000000E+00     I    0.000000E+00
 HEAT TRANS. COEF. BTU/HR/FT2/FI     372.328         I     1261.24
 PRES DROP SPEC/CALC   PSI     I   0.00000/     5.262I   0.00000/     1.866
 TOTAL FLUID           LB/HR   I  67328.5           I  600334.
       VAPOR       (IN/OUT)    I 67328.516/   0.00000I   0.00000/   0.00000
```

```
      LIQUID      (IN/OUT)    I   0.00000/ 67328.516I 6.003E+05/ 6.003E+05
PROPERTIES                    I                     I
----------                    I                     I
SP.GRAVITY AT 60 F  DEG API   I      134.09         I      9.9867
MOLECULAR WEIGHT  (IN/OUT)    I                     I
      VAPOR                   I    48.810/   0.00000I   0.00000/   0.00000
      LIQUID                  I   0.00000/    48.810I    18.015/    18.015
VISCOSITY         (IN/OUT)    I                     I
      VAPOR       CP          I   0.00971/   0.00000I   0.00000/   0.00000
      LIQUID      CP          I   0.00000/   0.08466I   0.67973/   0.66275
DENSITY           (IN/OUT)    I                     I
      VAPOR       LB/FT3      I     1.890/   0.00000I   0.00000/   0.00000
      LIQUID      LB/FT3      I   0.00000/    29.617I    63.203/    62.723
SPECIFIC HEAT     (IN/OUT)    I                     I
      VAPOR       BTU/LB/F    I   0.50793/   0.00000I   0.00000/   0.00000
      LIQUID      BTU/LB/F    I   0.00000/   0.73386I   0.98265/   0.99345
THE. CONDUCTIVITY (IN/OUT)    I                     I
      VAPOR       BTU/FT/HR/F I   0.01418/   0.00000I   0.00000/   0.00000
      LIQUID      BTU/FT/HR/F I   0.00000/   0.05089I   0.67135/   0.67500
-------------------------------------------------------------------------------
HEAT EXCHANGED       BTU/HR   =  9.328196E+06
OVERALL U COEFF. BTU/HR/FT2/F =   257.310
MTD  (= Q/UA)        DEG F    =    34.089
CORRECTION FACTOR             =     0.92613
NO. OF SHELL PASSES/SHELL     =     1
NO. OF SHELLS IN PARALLEL     =     1
SURFACE AREA SPEC/CALC FT2    =  0.00000E+00/  2547.8
EST. REQUIRED AREA/SHELL FT2  =   1063.5
TUBE MATERIAL                 =  CARBON STEEL
-------------------------------------------------------------------------------
      CONSTRUCTION OF ONE SHELL
-------------------------------------------------------------------------------
SHELL SIDE                          TUBE SIDE
==========                          =========
INSIDE DIAMETER  IN   =    35.000   OUTSIDE DIAMETER IN  =     0.750
BAFFLE SPACING   IN   =     7.000   PITCH SQUARE     IN  =     1.000
TUBE BUNDLE DIA. IN   =    34.600   LENGTH           FT  =    16.000
BAFFLE-CLEARANCE IN   =     0.175   INSIDE DIAMETER  IN  =     0.620
BAFFLE CUT            =     0.200   NO. OF PASSES        =     2
SEALING STRIPS        =     2       NO. OF TUBES         =   811
-------------------------------------------------------------------------------

           TEMPERATURE PROFILES IN HEAT EXCHANGER    2

       0.0       1.0        2.0        3.0        4.0       5.0
  150.0+----------+----------+----------+----------+----------+150.0
       I                                                      I
       I                                                      I
       I                                                      I
       S                                 S SHELLSIDE          I
       I**                               * INTERPOLATED       I
       I  **                             T TUBESIDE           I
       I    **                           + INTERPOLATED       I
  140.0+      **                                              +140.0
       I        **                                            I
       I          S*                                          I
       I            ***                                       I
       I               **                                     I
       I                 ***                                  I
T      I                    *S*                               I
E      I                       ****                           I
M 130.0+                           ****                       +130.0
P      I                               *S**                   I
E      I                                   ******             I
R      I                                         **S*****     I
A      I                                                *****S
T      I                                                      I
U      I                                                      I
R      I                                                      I
E 120.0+                                                      +120.0
       I                                                      I
D      I                                                      I
E      I                                                      I
G      I                                                      I
       I                                                      I
F      I                                                      I
       I                                                      I
```



Exchgr15.psd - Refrigerant Exchanger, UA Specification

Example showing heat exchanger simulation with specified UA and refrigerant temperature. The temperature or pressure of the refrigerant is calculated along with the flowrate required and all properties. The properties of the process stream out are calculated. The duty, area, and corrected LMTD are calculated and reported in the equipment summary.



Refrigerant Exchanger - Calculated Flow

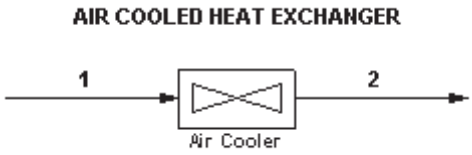
```
AB123.
*Example of Heat Ex 15: Refrigerant UA Spec.

C- EQUIPMENT MODULES

HEA EXC 1 = REFR, 9,11,-10,-12,
  SIM
  U(BTU/HR/FT2/F) = 66.
  ARE(FT2) = 10000.
  SHE PAS = 2
  TUB PAS = 4
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5., 0.,
  TEM REF(F)4 = 10.
REFR = TUBE
C-Refrigerant exchanger with U and AREa specification to calculate the
C-flow rate of C3 refrigerant such that refrigerant will vaporize from
C-a bubble point liquid to a dew point vapor.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 3,4,5,6,7,8,10,11,62,
  RKK
  STDH
```

```
  STDD
  SUR TEN = STD
  IMM = 62
  AME UNI OUT
  MAX = 5
  CON TOL = 1.E-003
  FLO(GMOL/SEC)9 = 500.,2*200.,4*100.,0.,150.,
  TP (F,PSIA) 9 = 200.,150.
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  FLO(LBMOL/HR)11 = 8*0.,150.,
  TP (F,PSIA) 11 = 10.,50.
  NAM STR 11 = Inlet Refrigerant
  NAM STR 12 = Refrigerant out
END
```

Exchgr16.psd - Air Cooled Example

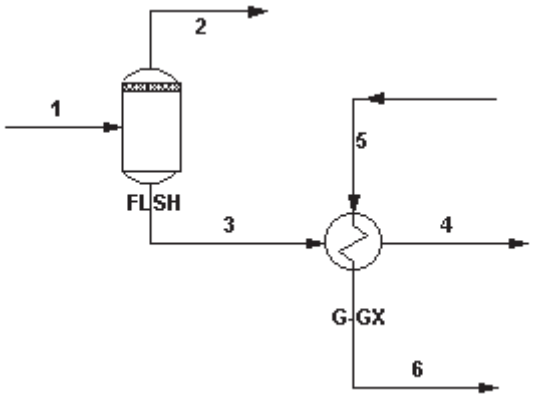
The FIN FAN exchanger example allows to specify the AREa and U of the exchanger. Calculates the required flow, duty, corrected LMTD and outlet stream properties for the process stream.

```
AB123.
*AIR COOLED HEAT EXCHANGER EXAMPLE

C- EQUIPMENT MODULES

HEA EXC 2 = Air Cooler, 1,-2,
  U(BTU/HR/FT2/F) = 4.2
  ARE(FT2) = 30000.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 5.
  FINFAN
  TEM AIR IN(F) = 90.
  TEM AIR OUT(F) = 105.
C-
C-  DESIGN II WILL CALCULATE A REQUIRED AIR
C-  RATE FOR THE HEAT EXCHANGER MODULE WHEN A
C-  FIN FAN COMMAND IS ENTERED.  THE DEFAULT
C-  INLET AIR TEMPERATURE IS 90 F, WITH A 15
C-  DEGREE F TEMPERATURE RISE THROUGH THE EXCHANGER.
C-  THE AIR RATE IS REPORTED IN THE EQUIPMENT SUMMARY.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 100,
  NAM 100 = PETFRAC
  AMW 100 = 175.
  AMB (F) 100 = 350.
  API (API) 100 = 37.
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  FLO(LB/HR)1 = 273000.,
  TP (F,PSIA) 1 = 250.,50.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
END
```



Exchgr17.psd - Rating E Shell in Series

Example rating of an E shell using KERN's method. With the addition of a RATing command and appropriate commands for shell-and-tube geometry, DESIGN II will perform a rating calculation on any single phase counter-current exchanger and almost all two-phase exchangers in your flowsheet.

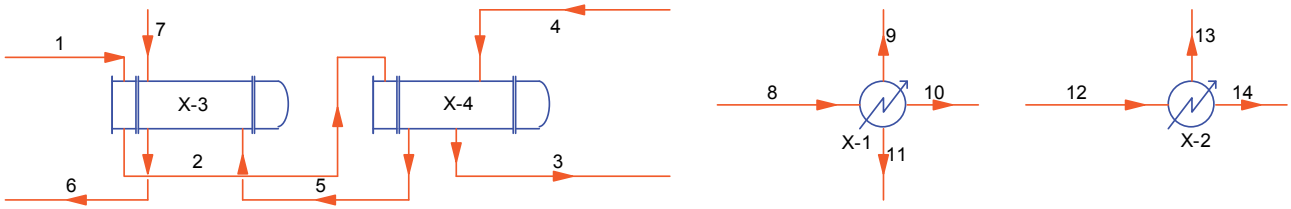



```
AB123.
*EXCHANGER RATING E-SHELL IN SERIES

C- EQUIPMENT MODULES

FLA 1 = FLSH, 1,-2,-3,

HEA EXC 3 = G-GX, 3,5,-4,-6,
  SIM
  U(BTU/HR/FT2/F) = 50.
  ARE(FT2) = 450.
  SHE PAS = 2
  TUB PAS = 2
  SHE = 1
  TUB FEE = 2
  INL = SAM
  HOR
  RAT
  NUM INC = 5
  MET = KERN
  EXC = AES
  TOL = 1.E-006
  NUM TUB = 160
  TUB LAY = TRI
  TUB MAT = ADM
  TUB LEN(FT) = 15.
  TUB PIT(FT) = 9.E-002
  TUB FOU(1/BTU/FT2/HR/F) = 1.E-003
  TUB OUT DIA(FT) = 7.E-002
  TUB BWG = 18
  BAF CUT = 0.25
  NUM SEA = 2
  BAF SPA(FT) = 0.4
  SHE FOU(1/BTU/FT2/HR/F) = 1.E-003
  SHE DIA(FT) = 1.9
  DEL(PSI) = 0., 0.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,2,3,4,5,6,7,8,10,11,12,100,
  NAM 100 = C8+
  AMB (F) 100 = 300.
  API (API) 100 = 40.
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  NO IMM
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 100.,2*500.,300.,60.,40.,2*30.,2*20.,10.,5.,
  TP (F,PSIA) 1 = 0.,900.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  FLO(LB/HR)5 = 0.,50.,100.,2000.,100.,53.,10.,5*0.,
  TP (K,ATM) 5 = 200.,13.6
  NAM STR 5 = Strm 5
```

```
  NAM STR 6 = Strm 6
END
```

Exchgr18.psd – Multiple inlet and outlet streams

An example of cascading heat eachangers using multiple inlet streams and an example of multiple outlet streams for a single stream heat exchanger.



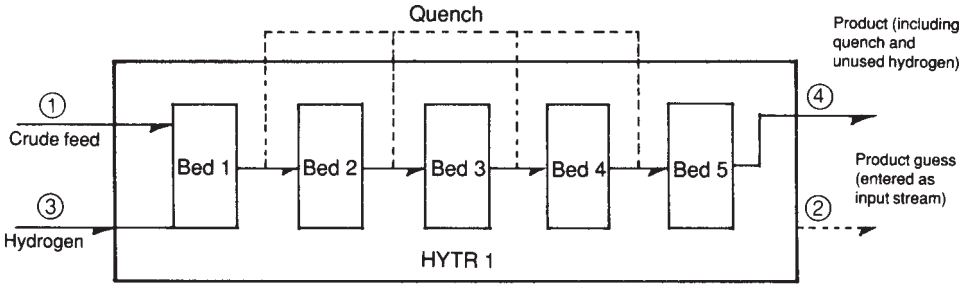

```
AB123.
*Example of Heat Exchanger 3: Specified Outlet Temperature

C- EQUIPMENT MODULES

HEA EXC 1 = X-1, 8,-9,-10,-11,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 35
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 4
  PRO 1 = THR

HEA EXC 2 = X-2, 12,-13,-14,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 35
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0
  PRO 1 = TWO

HEA EXC 3 = X-3, 5,7,1,-6,-2,
  U(BTU/HR/FT2/F) = 50
  TEM OUT TUB(F) = 70
  NUM FEE 1 = 2
  NUM FEE 2 = 1
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

HEA EXC 4 = X-4, 4,2,-5,-3,
  U(BTU/HR/FT2/F) = 50
  TEM OUT TUB(F) = -20
  NUM FEE 1 = 1
  NUM FEE 2 = 1
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,
```

```
C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,5,6,7,8,62,
  NAM POS 1 = 'METHANE'
  NAM POS 2 = 'ETHANE'
  NAM POS 3 = 'PROPANE'
  NAM POS 4 = 'I-BUTANE'
  NAM POS 5 = 'BUTANE'
  NAM POS 6 = 'I-PENTANE'
  NAM POS 7 = 'N-PENTANE'
  NAM POS 8 = 'WATER'
C-  Thermo Comments
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 5
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 200,100,50,
  TP (F,PSIA) 1 = 100,250
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  FLO (FRA) 4 = 2*0.1,0.8,
  TOT FLO (LBMOL/HR) 4 = 1000
  TP (F,PSIA) 4 = -100,15
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  FLO (FRA) 7 = 2*0.1,0.8,
  TOT FLO (LBMOL/HR) 7 = 1000
  TP (F,PSIA) 7 = -100,15
  NAM STR 7 = Strm 7
  FLO(LBMOL/HR)8 = 50,2*100,2*50,2*40,10,
  TP (F,PSIA) 8 = 100,200
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  FLO(LBMOL/HR)12 = 200,100,50,4*0,10,
  TP (F,PSIA) 12 = 100,50
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
END
```

Chapter 31: Hydrotreater Samples - c:\designii\samples\equipmnt\hytr

Hydrtr.psd - Crude Treating

A crude stream (1) is fed to a hydrotreater to be hydrogenated, denitrogenated, and desulfurized. The hydrogen stream (3) composition and flowrate are specified by the user. Hydrogen will be used for the quench stream. Composition and flowrates are also entered for the product stream (2).



Specifications

delta P = 5 PSI per bed
Temperature of quench = 140 ºF
Pressure of quench = 170 PSIA

Temperature rise:	Bed 1 = 50 ºF	Bed 2 = 51 ºF	Bed 3 = 49 ºF	Bed 4 = 50 ºF	Bed 5 = 51 ºF
Temperature drop per quench:	Bed 1 =20 ºF	Bed 2 =21 ºF	Bed 3 =19 ºF	Bed 4 =20 ºF	

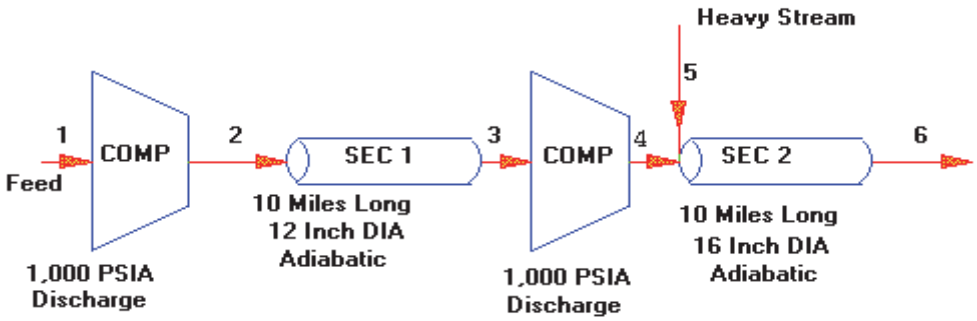
```
HYdroTReater 1 = REAC, 1,2,3,-4,
       SULfur REActed = 0.9,
       NITrogen REActed  = 0.9,
       OLEfin REActed = 0.9,
       AROmatic REActed  = .05,
       SULfur CONtent = .00025,
       NITrogen CONtent  = .0001,
       OLEfin CONtent = .344,
       AROmatic CONtent  = .274,
       TEMperature of QUEnch = 140,
       PREssure of QUEnch  = 170,PRInt = 2,
       OPTion for temperature RISe per bed   = 1,
       OPTion for temperature DROp per quench = 1,
       PREssure DROp per bed = 5,
       OPTion QUEnch = 1,
       MAXimum number of beds = 5,
       TEMperature RISe per bed = 50,51,49,50,51,
       TEMperature DROp per quench = 20,21,19,20
GENeral
       COMponents =1,2,3,4,5,6,7,8,100,101,102,103,104,105,106,107,108,63,50
       TP 1=550,200
       TP 2=600,200
       TP 3=300,200
       API 100 = 82.1, AMB 100 = 136, AMW 100 = 87.89, NAMe 100 = 136 ABP
       API 101 = 68.3, AMB 101 = 195, AMW 101 = 100.44, NAMe 101 = 195 ABP
       API 102 = 59.1, AMB 102 = 235, AMW 102 =108.81, NAMe 102 = 235 ABP
       API 103 = 53.2, AMB 103 = 275, AMW 103 =119.25, NAMe 103 = 275 ABP
       API 104 = 49.9, AMB 104 = 315, AMW 104 = 131.86, NAMe 104 = 315 ABP
       API 105 = 46.8, AMB 105 = 355, AMW 105 =145.47, NAMe 105 = 355 ABP
       API 106 = 43.6, AMB 106 = 395, AMW 106 =159.88, NAMe 106 = 395 ABP
       API 107 = 40.2, AMB 107 = 435, AMW 107 =174.69, NAMe 107 = 435 ABP
       API 108 = 36.7, AMB 108 = 475, AMW 108 =190.24, NAMe 108 = 475 ABP
       FLOw 1 =  0,0,15.4,63.89,14.88,40.45,11.96,10.86,
       355.87,218.22,263.3,261.11,249.96,232.4,232.35,232.13,220.74,2*0,
       FLOw  2 =  0,0,15.4,63.89,14.88,40.45,11.96,10.86,
       358.67,219.26,270.11,274.04,241.2,228.5,227.3,228.4,218.5,2*0,
       FLOw  3 = 85,7.7,3.2,2.0,.5,.6,1,12*0,
END
```

FLOw 1 is the crude feed. FLOw 2 is the estimated hydrocarbon product flowrates excluding quench and is coded as the next to last stream. FLOw 3 is the hydrogen stream flowrates and is coded as the last stream.

Chapter 32: Line Module Samples - c:\designii\samples\equipmnt\line

Line1.psd - Intermediate Compression and Junction

Gas is compressed to 1000 PSIA (COMPREssor 1) for transmission through a 12-inch diameter adiabatic line (LINE 10). At the next compressor station (COMPREssor 2) there is a junction with another stream of heavier composition (stream 5) entering the next line section (LINE 20) along with the compressor outlet stream (stream 4). LINE 20 is a 16-inch, two-phase adiabatic line. The plot of pressure and elevation profiles includes both LINE 10 and LINE 20.



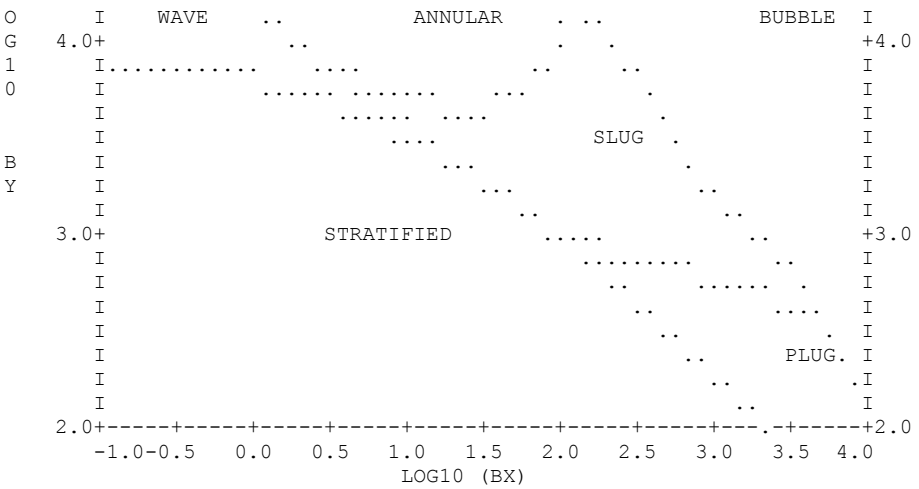
20 Mile Pipeline with Intermediate Compression

```
COMPREssor 1 = COMP,1,-2, ENTropy, PREssure OUT = 1000,
LINE 10 = SEC 1,2,-3,
        DIAmeter (IN) = 12.0,
        ADIabatic,
        LENgth (MILES) = 10*1.0,
        ELEvation (FT) = 5,-2,3,4*1,3*2
COMPREssor 2 = COMP, 3,-4,
        PREssure OUT = 1000,
        ENTropy
LINE 20 = SEC 2,4,5,-6,
        DIAmeter (IN) = 16,
        ADIabatic
        LENgth (MILES) = 10*1.0,
        ELEvation (FT) = 10,5,2,5,-5,10,3,4,2,-8,
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,11,12,13,14,
        FLOw 1 = 5,3,5000,3000,1000,700,500,7*0
        TP 1 = 120,800
        SOAVEK
        TRANSport properties = NBS81
        FLOw 5 = 2*0, 50,100,3*50,7*200
        TP 5 = 120,1000
        PLOT PREssure AND ELEvation against LINEs = 10,20
END
```

A partial output file from Example 1 appears below.

```
LIN        10 = SEC
             RESULTS OF LINE    10     -     GAS FLOW
              INLET STREAM(S)   2
             OUTLET STREAM(S)   3
             LINE CHARACTERISTICS: HORIZONTAL
                                   ADIABATIC
             ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       3.8732E+06  LIQUID FLOWRATE GAL/HR       0.0000E+00
PIPE ROUGHNESS    FT        1.5000E-04  PIPE EFFICIENCY FACTOR       1.000
```

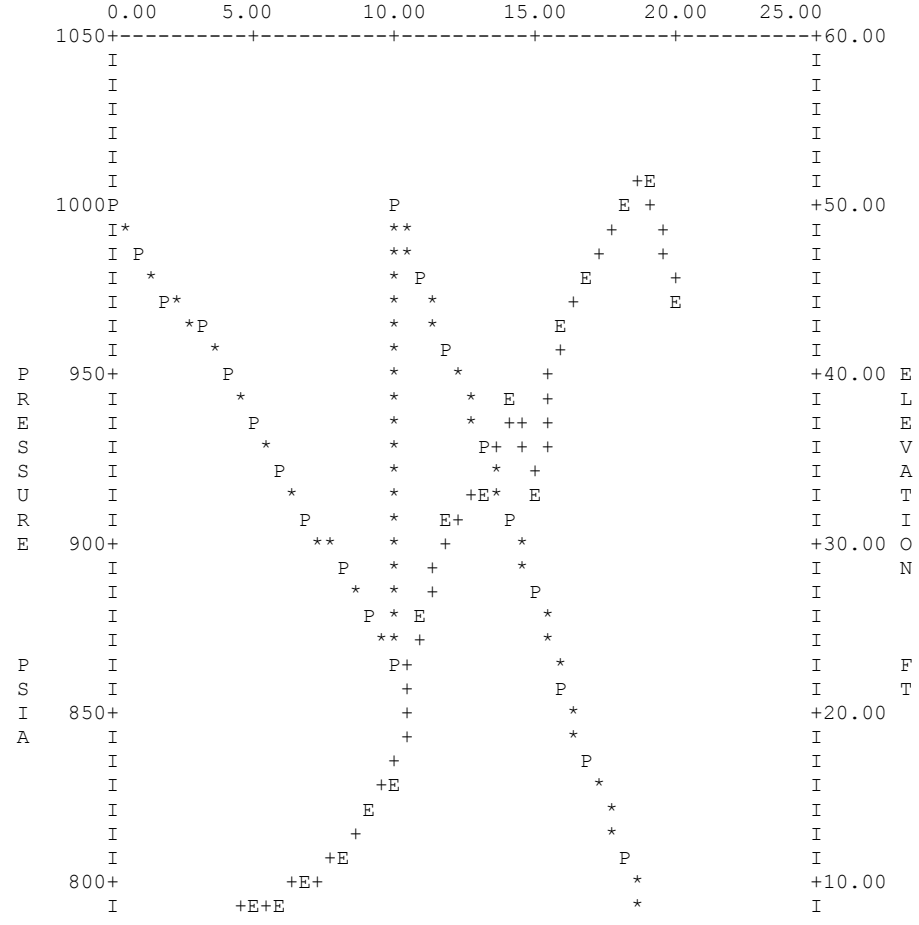
```
INTERNAL DIAMETER INCHES   12.000       OUTSIDE DIAMETER INCHES     12.750
NOMINAL DIAMETER  INCHES   12.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              DARCY               MOODY               PHASE DENSITY

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.

LINE    LENGTH   ELEV.  TEMP. DELP-FRIC DELP-ELEV PRESS. VELOC. LIQUID  FLOW
NODE       FT      FT   DEG F   PSI       PSI     PSIA   FT/SEC HOLDUP REGIME

  0        0.0     0.0 150.5                    1000.00   16.8 0.0000  TURB
  1     5280.0     5.0 149.5     12.60     0.21   987.19   17.0 0.0000  TURB
  2    10560.0     3.0 148.5     12.77    -0.08   974.51   17.2 0.0000  TURB
  3    15840.0     6.0 147.4     12.94     0.12   961.44   17.5 0.0000  TURB
  4    21120.0     7.0 146.4     13.12     0.04   948.28   17.7 0.0000  TURB
  5    26400.0     8.0 145.3     13.31     0.04   934.93   18.0 0.0000  TURB
  6    31680.0     9.0 144.1     13.51     0.04   921.38   18.3 0.0000  TURB
  7    36960.0    10.0 143.0     13.72     0.04   907.63   18.6 0.0000  TURB
  8    42240.0    12.0 141.8     13.94     0.08   893.62   18.9 0.0000  TURB
  9    47520.0    14.0 140.5     14.16     0.07   879.38   19.2 0.0000  TURB
 10    52800.0    16.0 139.3     14.41     0.07   864.90   19.5 0.0000  TURB

      ESTIMATED SONIC VELOCITY IN GAS PHASE
        =    940.3 FT/SEC   AT  INLET CONDITIONS.
        =    931.4 FT/SEC   AT OUTLET CONDITIONS.

LIN        20 = SEC
             RESULTS OF LINE    20     -     TWO PHASE FLOW
              INLET STREAM(S)   6
             OUTLET STREAM(S)   7
             LINE CHARACTERISTICS: HORIZONTAL
                                   ADIABATIC
             ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       2.8999E+06  LIQUID FLOWRATE GAL/HR       5.1535E+04
PIPE ROUGHNESS    FT        1.5000E-04  PIPE EFFICIENCY FACTOR       1.000
INTERNAL DIAMETER INCHES   15.250       OUTSIDE DIAMETER INCHES     16.000
NOMINAL DIAMETER  INCHES   16.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              LOCKHART-MARTINELLI MOODY               PHASE DENSITY
METHODS FOR:  LIQUID HOLDUP       FLOW REGIME
              HUGHMARK            BAKER

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.

LINE    LENGTH   ELEV.  TEMP. DELP-FRIC DELP-ELEV PRESS. VELOC. LIQUID  FLOW
NODE       FT      FT   DEG F   PSI       PSI     PSIA   FT/SEC HOLDUP REGIME

  0        0.0     0.0 166.1                    1000.00   10.5 0.3092  DISP
  1     5280.0    10.0 165.1     22.26     0.87   976.88   10.7 0.3026  DISP
  2    10560.0    15.0 164.0     22.68     0.43   953.77   11.0 0.2960  DISP
  3    15840.0    17.0 163.0     23.11     0.17   930.49   11.4 0.2896  DISP
  4    21120.0    22.0 161.9     23.57     0.41   906.51   11.7 0.2830  DISP
  5    26400.0    17.0 160.8     24.05    -0.41   882.87   12.1 0.2766  DISP
  6    31680.0    27.0 159.5     24.54     0.80   857.53   12.5 0.2698  DISP
  7    36960.0    30.0 158.3     25.09     0.23   832.20   12.9 0.2631  DISP
  8    42240.0    34.0 157.0     25.67     0.31   806.22   13.4 0.2563  DISP
  9    47520.0    36.0 155.6     26.28     0.15   779.79   13.9 0.2494  DISP
 10    52800.0    28.0 154.2     26.93    -0.59   753.45   14.4 0.2426  DISP

++NOTE: NO SLUG FLOW REGIME IS FOUND, SLUG ANALYSIS BYPASSED

      ESTIMATED SONIC VELOCITY IN GAS PHASE
        =   1022.0 FT/SEC   AT  INLET CONDITIONS.
        =   1030.2 FT/SEC   AT OUTLET CONDITIONS.
     *******  BAKER TWO-PHASE FLOW REGIME MAP FOR LINE    20 *******
       -1.0-0.5   0.0   0.5   1.0   1.5   2.0   2.5   3.0   3.5  4.0
       -1.0-0.5   0.0   0.5   1.0   1.5   2.0   2.5   3.0   3.5  4.0
    5.0+-----+----.+-----+-----+-----+-----+--.--+-----+-----+-----+5.0
       I..          ..                    .                     I
       I  ..          ...        DISPERSED  .                    I
       I    ..           .....              .                    I
       I      ..              .........XXX...                    I
       I        ..                         .                    I
L      I          ..                       .                    I
```



BAKER FLOW REGIME COORDINATES

SEG NO.	X	Y	LOG10(X)	LOG10(Y)
1	67.214348	29006.177734	1.827462	4.462491
2	64.482086	29520.214844	1.809439	4.470119
3	61.824677	30055.173828	1.791162	4.477919
4	59.181705	30624.833984	1.772187	4.486074
5	56.664265	31206.027344	1.753309	4.494238
6	54.057697	31851.724609	1.732858	4.503133
7	51.542576	32522.601562	1.712166	4.512185
8	49.053478	33238.515625	1.690670	4.521642
9	46.609001	33998.683594	1.668470	4.531462
10	44.255524	34791.738281	1.645967	4.541476

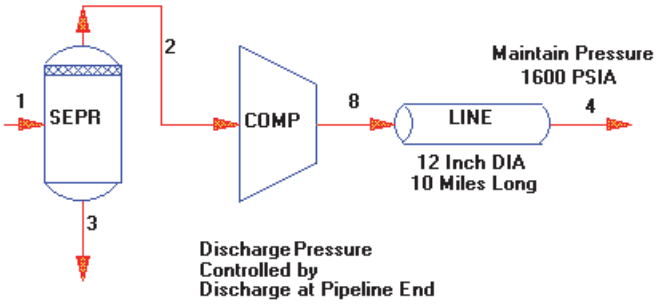
LINE PLOT 1 OF PRESSURE AND ELEVATION AGAINST LENGTH
FOR LINE(S) 10 20



```
        I       +E                                  *            I
        I      E                                    P            I
        I E+ ++                                      *           I
        I+  E                                        *           I
        I+                                             *         I
     750E                                              P         +0.00
        I                                                        I
        I                                                        I
        I                                                        I
        I                                                        I
        I                                                        I
        I                                                        I
     700+----------+----------+----------+----------+----------+-10.00
        0.00      5.00       10.00      15.00      20.00     25.00
                             LENGTH      MILES
```

Line2.psd - Back Pressure Calculation

An offshore transmission line has to meet an onshore pressure of 1600 PSIA. The coding below shows how this problem can be solved using a CONTroller module to back calculate the compression discharge pressure. Pressure drops due to friction and changes in elevation are calculated by the AGA method.



Pipeline with 1600 PSIA Discharge Pressure

```
FLAsh drum 1 = SEPR, -1,-2,-3,
COMPREssor 4 = COMP, 2,-8,
   ENTropy,PREssure OUT = 1700
LINE 7 = LINE, 8, 11,
   nominal DIAmeter (IN) = 12,
   DOWN,
   LENgth = 150,
   METhod for FRIction = AGA,
   DRAg factor = 0.96,
   METhod for ELEvation = AGA
LINE 8 = LINE, 11,12,
        DIAmeter (IN) = 12,
        LENgth (MILES) = 10*1,
        ELEvation (FT) = 3*10, 5,2,20,30,3*10,
        METhod for FRIction = AGA,
        DRAg factor = 0.96,
        METhod for ELEvation = AGA
CONTroller 10 = FIXP, 12,13,
        VARY PRE OUT (MIN=1500,MAX=2500) OF COMPRE 4
        UNTIL PREssure FROm STReam 12 = 1600 (TOL=25)
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,11,12,13,14,62
        IMMiscible = 62
        FLOw 1 = 5, 3, 5000, 3000, 1000, 700, 500, 233,
         339, 226, 427, 207, 266, 234, 100
        TP 1 = 120,1400
        SOAVEK, SOAVEH, STDD
END
```

Line3.psd - Diameter Case Studies

If you are studying alternative pipe sizes for a line with known flow and topography, you can calculate all of your cases in a single LINE module. In this case the downstream pressure out of the LINE module will be the outlet pressure from the last diameter calculated (20 inch).

14, 16, 18, 20 Inch
Line Diameter Study

```
LINE 2 = LINE,1,2,
        nominal DIAmeter (IN) = 14.0,16.0,18.0,20.0
        LENgth (MILES) = 10*1.0
        ELEvation (FT) = 5,2,3,4*1,3*2
GENeral
        COMponents = 2,3,4
        FLOw 1 = 500,1000,500
        TP 1 = 150,100
        TRANSport properties = NBS81
END
```

Line4.psd - Two Phase Line Calculations

An offshore transmission line is transporting 300 MMSCFD of gas through a 24 inch line over a distance of 35 miles. The example below uses a heat transfer coefficient and a temperature of the surrounding to determine the pressure drop and slugging characteristics of the line.

```
AB123.
*Two Phase Line Calculations

C- EQUIPMENT MODULES

LINE 1 = OFFSHORE, 1,-2,
  DIA (INCH) = 24
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35. MILE
  LEN (MILE) = 25.,10.,
  ELE(FT) = 1.,5.,
  SEG LEN (MILE) = 3.5
  MET FRI = BEG
  TEM SUR(F) = 2*29.,
  U(BTU/HR/FT2/F) = 2*2.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,11,12,13,
  PENK
  PENH
  STDD
  VIS = NBS81
  THE CON = NBS81
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO(FRA)1 = 3.E-003,1.1E-002,0.802,8.9E-002,6.E-002,7.E-003,1.7E-002,3.E-003,
  4.E-003,2.E-003,1.6E-003,3.E-004,1.E-004,
  TOT FLO(MMSCF/DAY)1 = 300.
  TP (F,PSIA) 1 = 50.,1350.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
END
```

Offshore Pipeline

The tabular results generated by the line module for the above input file is shown on the following page.

```
LIN         1 = OFFS
              RESULTS OF LINE     1      -      WET GAS FLOW
               INLET STREAM(S)    1
              OUTLET STREAM(S)    2
              LINE CHARACTERISTICS: HORIZONTAL
```

```
                                  HEAT TRANSFR
            ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       1.2256E+07  LIQUID FLOWRATE GAL/HR        6009.
PIPE ROUGHNESS    FT        1.5000E-04  PIPE DRAG FACTOR             1.000
INTERNAL DIAMETER INCHES   23.250       OUTSIDE DIAMETER INCHES     24.000
NOMINAL DIAMETER  INCHES   24.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              BEGGS AND BRILL     BEGGS AND BRILL     PHASE DENSITY
METHODS FOR:  LIQUID HOLDUP       FLOW REGIME
              BEGGS AND BRILL     BEGGS AND BRILL

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.

LINE    LENGTH   ELEV.  TEMP. DELP-FRIC DELP-ELEV PRESS. VELOC. LIQUID  FLOW
NODE       FT      FT   DEG F   PSI       PSI     PSIA   FT/SEC HOLDUP REGIME

  0        0.0     0.0  50.0                     1350.00    8.3 0.1038  SEGR
  1    18480.0     0.1  44.3      8.41     0.01  1341.58    8.1 0.1262  SEGR
  2    36960.0     0.3  40.1      8.20     0.01  1333.37    8.0 0.1428  SEGR
  3    55440.0     0.4  37.0      8.06     0.01  1325.29    7.9 0.1552  SEGR
  4    73920.0     0.6  34.7      7.98     0.01  1317.31    7.9 0.1645  SEGR
  5    92400.0     0.7  33.0      7.93     0.01  1309.37    7.9 0.1714  SEGR
  6   110880.0     0.8  31.7      7.90     0.01  1301.45    7.9 0.1764  SEGR
  7   129360.0     1.0  30.8      7.90     0.01  1293.54    7.9 0.1801  SEGR
  8   132000.0     1.0  30.6      1.13     0.00  1292.41    7.9 0.1808  SEGR
  9   150480.0     2.8  29.9      7.93     0.13  1284.35    7.9 0.1833  SEGR
 10   168960.0     4.5  29.4      7.95     0.13  1276.27    8.0 0.1850  SEGR
 11   184800.0     6.0  29.1      6.84     0.11  1269.31    8.0 0.1857  SEGR
```

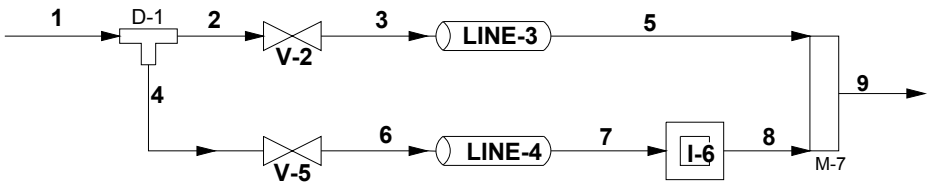
Line5.psd - Example Calculating Upstream Pressure

This flowsheet shows a line module calculating backwards by way of using a command for "calculating upstream pressure" (see Calculation Methodology below) when a target pressure out is needed for each line. After using this command one can set this calculated upstream pressure for the stream entering each of the line module. While doing this the divider flow fraction to each line can be varied to accommodate similar pressure fed to EACH line module. This is accomplished in the following way.

1. The first requires accessing CALculate UPStream PREssure command for a specified PRE OUT of each line module (see Inline Fortran and Keyword Input of each line module).
2. The valve pressure out is reset from the line module's calculated upstream pressure. Note that we have used user specified CALculation SEQuence command from the Specify...Keyword Input dialog.
3. The controller is set-up to vary the divider flow fraction until the stream pressures of #3 and #6 are close to each other. In this example the values were calculated as 365.48 and 365.28 psia.

Note that this example is specific for the line module specifications and feed pressure, temperature and flow specifications. If any change is done then you would have re-specify, change criteria in appropriate modules and re-run the case.

CALCULATE UPSTREAM PRESSURE (#3 & #6)



Calculation Methodology

The LINe module now allows a downstream pressure specification. When the down stream pressure is specified, the required upstream pressure is calculated. The calculated upstream pressure is available to other DESIGN II modules through Inline FORTRAN. For example, this feature may be used to specify an upstream compressor discharge pressure. The calculated upstream pressure is only reported when specified with Inline FORTRAN (see example below).

Major benefits of this improvement include direct calculation of downstream pressure. Previously, an iterative loop with a controller module was used to adjust the upstream pressure in order to achieve a specified downstream pressure. The calculated upstream pressure can be passed to an upstream module using Inline FORTRAN. You will need two additional commands in the GENeral section so that the pressure changing module and the Line module will be recalculated with the

new pressure value.

CALSEQ = n1, n2, ... or REC SEQ = n1, n2, ...

where n1, n2, ... are equipment module numbers, and MAX = 10

Example:

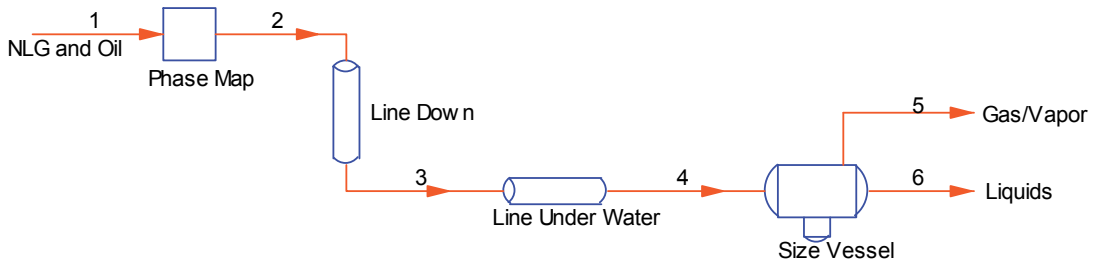
```
AB123.
*Calculate Upstream Pressure

C- EQUIPMENT MODULES

DIV 1 = D-1, 1,-4,-2,
  FLO(FRA) = 5.E-002

VAL 2 = V-2, 2,-3,
  PRE OUT(PSIA) = 10.

LINE 3 = LINE-3, 3,-5,
  DIA (INCH) = 18
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10. MILE
  LEN (MILE) = 10.,
  SEG LEN (MILE) = 1.
  MET FRI = BEG
  ADI
PRE OUT (PSIA) = 360
F-      POST
F-      UPRE =GET(LINE(3),CAL UPS PRE)
F-      WRITE(2,95)
F-  95  FORMAT (' ')
F-      WRITE(6,100)UPRE
F-      WRITE(1,100)UPRE
F- 100  FORMAT(2X,'UPSTREAM PRESSURE LINE 3 (PSIA)=',G12.5)
F-      SET(VALVE(2),PRE OUT)=UPRE

LINE 4 = LINE-4, 6,-7,
  DIA (INCH) = 18
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 14. MILE
  LEN (MILE) = 14.,
  SEG LEN (MILE) = 1.4
  MET FRI = BEG
  ADI
PRE OUT (PSIA) = 320
F-      POST
F-      UPRE =GET(LINE(4),CAL UPS PRE)
F-      WRITE(2,95)
F-  95  FORMAT (' ')
F-      WRITE(6,100)UPRE
F-      WRITE(1,100)UPRE
F- 100  FORMAT(2X,'UPSTREAM PRESSURE LINE 4 (PSIA)=',G12.5)
F-      SET(VALVE(5),PRE OUT)=UPRE

VAL 5 = V-5, 4,-6,
  PRE OUT(PSIA) = 10.

CONT 6 = I-6, 7,-8,
VARY FLO (MIN=0.05, MAX=1.0, STEP=0.05) OF DIV 1
UNTIL PRE FROM STR 3/PRE FROM STR 6=1.0 (TOL=0.005,LOO=100)

MIX 7 = M-7, 5,8,-9,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,2,3,4,5,6,7,8,10,11,
  PENK
  PENH
  STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
  THE CON VAP = NBS81
  THE CON LIQ = NBS81
  BINPAR = PENG1
  MAX = 5
```

```
  CON TOL = 1.E-003
  FLO(FRA)1 = 1.1E-002,0.802,8.9E-002,6.E-002,7.E-003,1.7E-002,3.E-003,4.E-003,
  2.E-003,1.6E-003,
  TOT FLO(MMSCF/DAY)1 = 100.
  TP (F,PSIA) 1 = 50.,900.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
CAL SEQ=1,2,3,2,5,4,5,6,7,
END
```

For Information on retrieving and printing calculated values such as upstream pressure, see Inline Fortran: Table 2. Variables Available to GET and SET.

Line6.psd - Line Surroundings with Heat Transfer

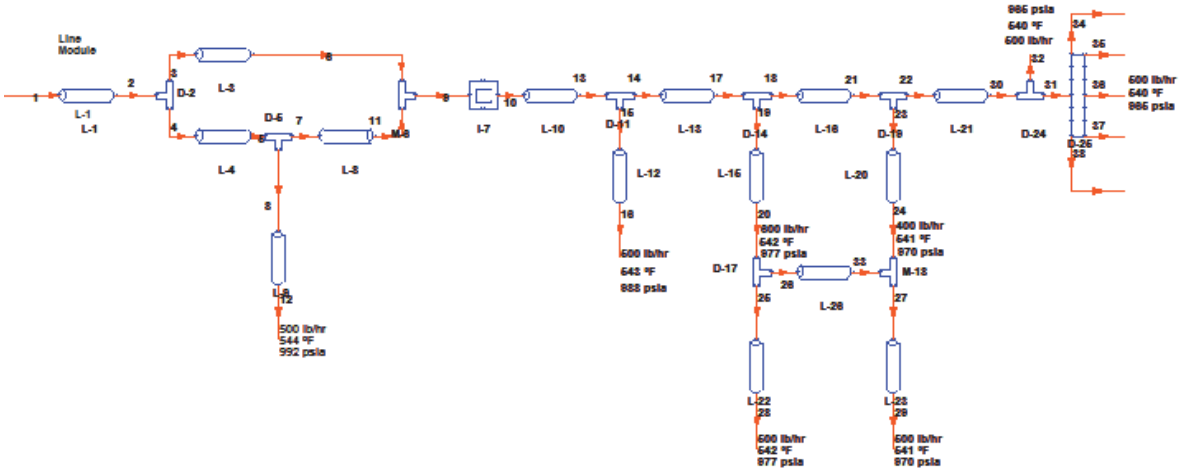
The example is to illustrate the heat transfer calculations equipped in the line module. Line down is exposed to Air at 70 F and 4 Ft/sec and the line under water is exposed to water flowing at .05 m/min. The external resistance due to heat transfer (conduction & convection) is included for the line modules.




```
LINE 2 = Line Down, 2,-3,
  DIA (INCH) = 16
  PIP WAL = STD
  DOWN
C- ENTRIES IN LEN COMMAND TOTAL TO 45.72 M
  LEN (M) = 45.72,
  SEG LEN (M) = 4.572
  MET HOL = DUN
  TEM SUR(F) = 70,
  PIP MAT = STA
  SUR AIR
  VEL SUR(FT/SEC) = 4,

LINE 3 = Line Under Water, 3,-4,
  DIA (INCH) = 16
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 16000 M
  LEN (M) = 10*1600,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (M) = 3.048,3.048,3.048,1.524,0.609,6.097,9.144,3.049,3.047,3.047
  SEG LEN (M) = 1600
  MET HOL = MUK
  TEM SUR(F) = 10*60,
  PIP MAT = CAR
  SUR WAT CON(KCAL/HR/M/C) = 10*0.1709,
  VEL SUR(M/MIN) = 10*0.051,
```

Steam Line for wells.psd – Simulation of Steam Line for Various Wells

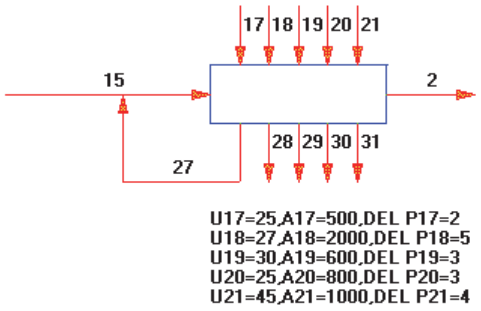
Simulation of steam line for various wells.



Chapter 33: LNG Exchanger Samples - c:\designii\samples\equipmnt\lngexc

Lngex1.psd - Simple Exchanger

The outlet temperature of the tubeside streams is estimated and the changes in enthalpies ("D"H) are calculated. These "D"H's are summed and the total "D"H used to calculate the shell temperature out. LMTD's are calculated for each tubeside stream and the heat transferred (Q) calculated from the Q = U*"A*"LMTD equation. The sum of the Q's is checked against the sum of the "D"H's and a new tubeside temperature out is tried until convergence is obtained.

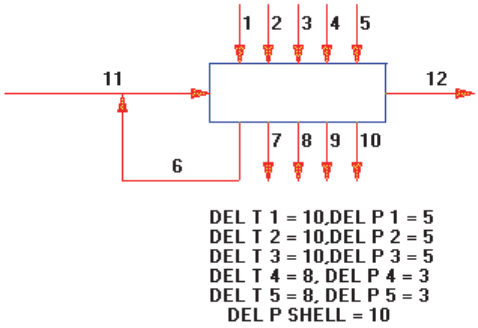


Number of Shells = 2, Shell Passes = 1, Tube Pass = 1, Pressure Drop across Shell =10

```
LNG 11 = E201, 17, 18, 19, 20, 21, 15, -27, -28, -29, -30,-31, -2
   SIMple,
   AREa = 500, 2000, 600, 800, 1000,
   U = 25, 27, 30, 25, 45,
   DELta pressure of TUBes = 2, 5, 3, 3, 4,
   SHElls =2, SHEll PASes = 1,
   TUBe PASes = 1,
   DELta pressure of SHElls = 10,
   RECycle STReams = 1
```

Lngex2.psd - Temperature Approach

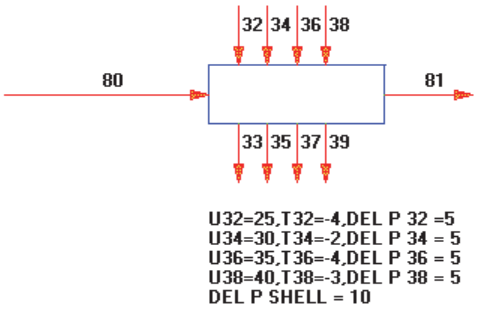
An approach can be set between the shellside stream in and the tubeside streams out. The approach is the absolute difference in temperature between the Shell inlet stream and the tubeside outset stream and is used calculating the shellside temperature out.




```
LNG 12 = E202, 1, 2, 3, 4, 5, 11, -6, -7, -8, -9, -10, -12
   TEMperature APProach = 3*10, 2*8,
   DELta pressure of TUBes = 3*5, 2*3
   DELta pressure of SHElls = 10,
   RECycle STReams = 1
```

Lngex3.psd - Temperature Specification

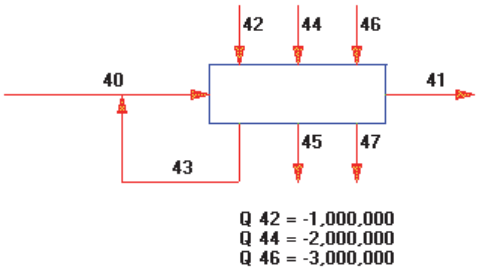
The outlet temperature of the tubeside streams can be specified. The areas and duties will be calculated.




```
LNG 13 = E203, 32, 34, 36, 38, 80, -33, -35, -37, -39, -81,
   TEMperature OUT = -4, -2, -4, -3,
   U = 25, 30, 35, 40,
   DELta pressure of TUBes = 4*5,
   DELta pressure of SHElls = 10,
   SHElls = 2
```

Lngex4.psd - Enthalpy Change Specification

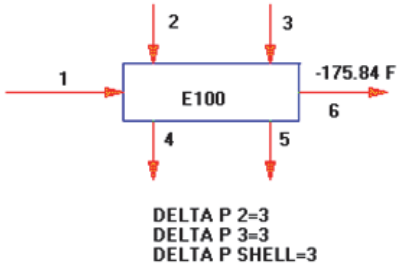
The tubeside streams can be heated or cooled by a specified change in enthalpy (Q). The tubeside temperatures out will be calculated at the enthalpy obtained by adding the specified cooling (-) or heating (+) Q to the feed stream enthalpy. The enthalpy change is specified in BTU/HR. Stream 43 is mixed with stream 40, and this becomes the shellside feed stream. The recycle mix temperature is reported in the LNG equipment summary.




```
LNG 14 = E204, 42, 44, 46, 40, -43, -45, -47, -41,
DUTy = -1E6, -2E6, -3E6, RECycle STReams = 1
```

Lngex5.psd - Shellside Temperature Specification Coding

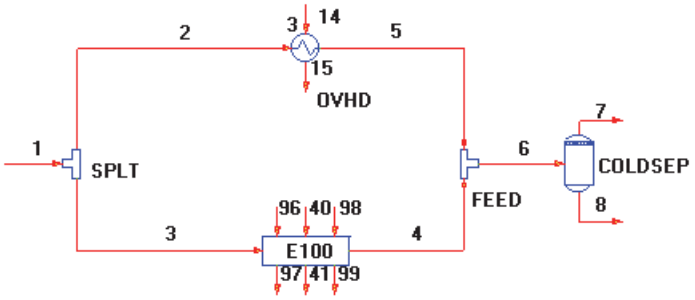
The shellside stream is being heated to -175.84F by the tubeside streams. The inlet enthalpy for the shellside stream is known. The enthalpy change is calculated, as is the outlet temperature for the tubeside streams. Both tubeside streams will exit at the same temperature.




```
LNG 1 = E100, 2, 3, 1, -4, -5, -6
   TEMperature OUT of SHEll = -175.84,
   DELta pressure of TUBes = 3,3,
   SHEll STR = 1
   DELta pressure of SHElls = 3
```

Lngex6.psd - Q-T Diagram

Use of Q-T command to show duty versus temperature curves for composite hot and cold streams. The results show all critical information such as narrow temperature approach, temperature crossovers, and discontinuity in composite curves.




```
DIVider 1 = SPLT,1, -3, -2,
   FLOw RATe = 0.41
LNG 2 = REBOILERS, 96, 40, 98, 3, -97, -41, -99, -4,
   DELta pressure TUBes =5,0,5,
   DELta pressure SHEll = 15,
   DUTy 1= 2.0696E6,
   DUTy 2= 2.6406E6,
   TEMperature OUT 3 = -38.1,
   Q-T
HEAt EXChanger 3 = OVHD, 14, 2, -15, -5,
   TEMperature APProach = 10,
   DELta PRESsure = 10, 10,
   Q-T
MIXer 4 = FEED, 4, 5, -6
FLAsh 5 = COLDSEP, 6, -7, -8
GENeral
       COMponents = 46, 49, 2, 3, 4, 5, 6, 7, 8, 10
       TP 1 = 90, 1000
       TP 14 = -136.1, 300
       TP 40 = -40, 21.761
       TP 96 = 1.5, 305.3
       TP 98 = -63.05, 303.3
       FLOw 1 = 1393.3, 500.47, 6131.4, 419.69, 133.4, 40.36,
       46.95, 19.77,  15.65, 31.29
       FLOw 14 = 1393.3, 268.4, 6124.3, 134.42, 3.6666,
       0.1553, 0.078664,  0.0027506, 0.86009E-3, 0.3633E-5
       FLOw(LB/HR)40 = 3*0.0, 6.6050E+02, 1.2453E+04,5.4720E+02
       FLOw(FRAction) 96 = 0, 0.3695, 0.0215, 0.3526, 0.123,
       0.0357, 0.041, 0.0169, 0.0133, 0.0265
       FLOw(FRAction) 98 = 0.2446E-4, 0.3215, 0.1853, 0.2848,
       .09821, 0.02915,  0.0337, 0.01405, 0.0111, 0.02214
       TOTal FLOw 96 = 1186.4
       TOTal FLOw 98 = 1414.6
       APISOAVEK
       APISOAVEH
       STDD
       VAPor = 14
END
```

Sample output for LNG Exchanger 2 follows. Notice that the cold stream in the Q-T results is a composite of streams 96, 40, and 98.

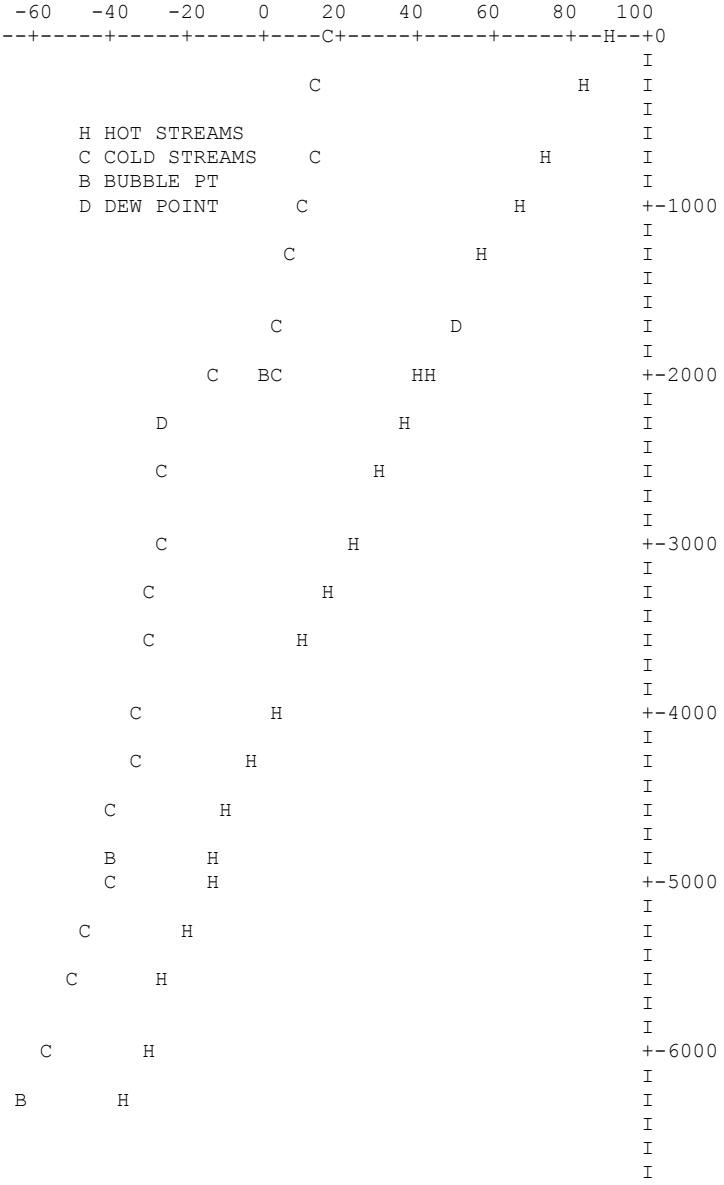
Table of Heat Content vs. Temperature for Exchanger 2

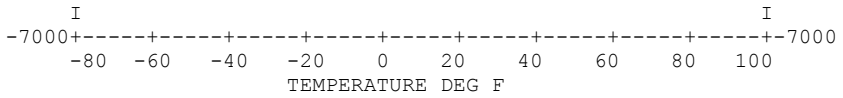
TABLE OF HEAT CONTENT-TEMPERATURE DIAGRAM

NO	TEMPERATURE DEG F HOT	COLD	HEAT FLOW(Q) BTU/HR	DELTA T DEG F	UA=DELTA Q/LMTD BTU/HR /F	
1	90.00	17.47	0.00000E+00	72.53	0.00000E+00	
2	81.88	14.63	-3.30128E+05	67.25	4725.7	
3	73.80	12.02	-6.60255E+05	61.78	5120.3	
4	65.84	9.62	-9.90383E+05	56.22	5599.8	
5	57.88	7.35	-1.32051E+06	50.53	6191.3	
6	50.03	4.96	-1.65064E+06	45.07	6914.2	
7	49.05	4.63	-1.69195E+06	44.42	923.31	DH
8	43.05	2.33	-1.98077E+06	40.71	6789.5	
9	42.96	1.53	-1.98516E+06	41.43	106.98	BC

```
10    41.22      1.50  -2.06960E+06     39.72    2081.5
11    41.22    -13.94  -2.06960E+06     55.16   0.13027      *
12    36.37    -25.69  -2.30483E+06     62.07    4017.9     DC
13    36.25    -26.39  -2.31089E+06     62.64    97.174
14    29.48    -27.37  -2.64102E+06     56.86    5529.5
15    22.82    -28.29  -2.97115E+06     51.11    6121.0
16    16.24    -29.33  -3.30128E+06     45.57    6836.9
17     9.77    -30.52  -3.63140E+06     40.29    7699.7
18     3.44    -32.09  -3.96153E+06     35.53    8719.8
19    -2.79    -34.61  -4.29166E+06     31.82    9813.3
20    -8.82    -38.40  -4.62179E+06     29.58    10758.
21   -11.98    -40.00  -4.79675E+06     28.02    6076.3     BC
22   -14.78    -41.42  -4.95192E+06     26.64    5678.3
23   -20.46    -46.34  -5.28204E+06     25.88    12573.
24   -26.11    -51.58  -5.61217E+06     25.47    12860.
25   -31.42    -57.16  -5.94230E+06     25.74    12893.
26   -36.58    -62.88  -6.26283E+06     26.30    12318.     BC
27   -36.73    -63.05  -6.27243E+06     26.32    364.68
                                                ----------------
                                    TOTAL UA =  1.60808E+05

 WHERE
   * IS DISCONTINUED Q-T CURVE
  BC IS BUBBLE PT OF COLD STREAMS
  DC IS DEW POINT OF COLD STREAMS
  DH IS DEW POINT OF HOT STREAMS
```

FIGURE- HEAT CONTENT-TEMPERATURE DIAGRAM (Q-T DIAGRAM)

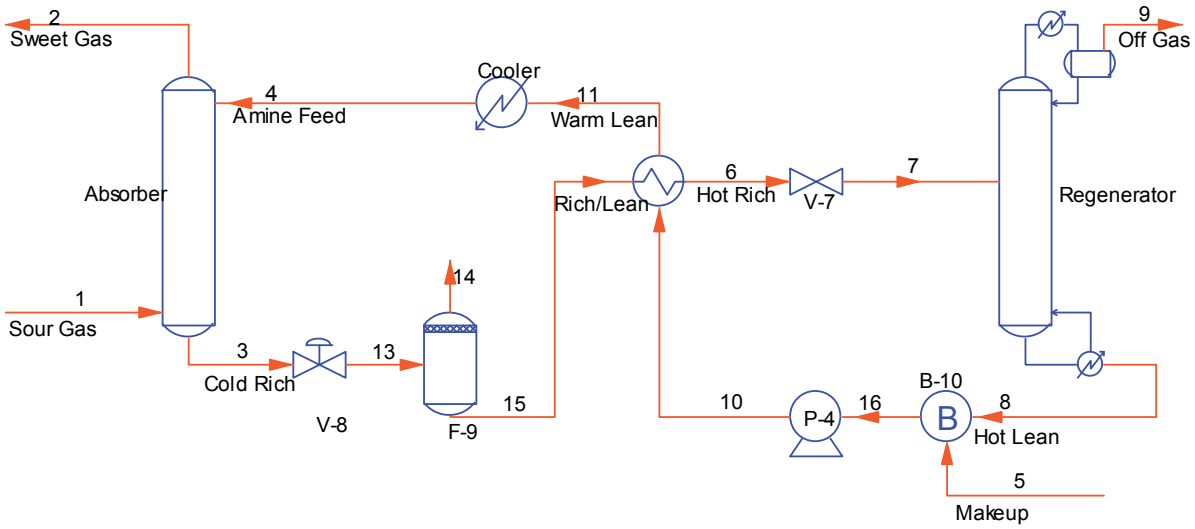



I
I
-7000+-----+-----+-----+-----+-----+-----+-----+-----+-----+-7000
-80 -60 -40 -20 0 20 40 60 80 100
TEMPERATURE DEG F

Chapter 34: Mass Balance Samples - c:\designii\samples\equipmnt\masbal

MxpInt1.psd - Mixed Amine Plant for CO2 / H2S Removal

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



The Mass Balance module (B-10) will both set the recycle flowrate to 105 gpm and vary the flowrate on feed stream 5 to meet the makeup flowrate requirements of the recycle for water, DEA and MDEA.

```
MAS BAL 10 = B-10, 5,8,-16,
  REC(GAL/MIN) = 105
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 5
  INP = -1
  OUT = -1

Sample Output:

NOW CALLING EQUIPMENT     10   +   MASBAL     =   B-10            19:55:57
KVALUES: AMIN   ENTHALPY: AMIN   VAPOR DENSITY: AMIN   LIQUID DENSITY: AMIN

    Setting flowrate of the product stream to   105.00          GAL/MIN

    Feed Streams:    1    5
    Product Streams:    2    9   14

            Component Id:      62
            Recycle Feed:      44.929        LBMOL/HR
            Recycle Product:   44.930        LBMOL/HR
            Old Makeup:        44.929        LBMOL/HR
            Adding Makeup:    0.11036E-02    LBMOL/HR
            New Makeup:        44.930        LBMOL/HR

            Component Id:    4051
            Recycle Feed:     0.15916E-02    LBMOL/HR
            Recycle Product:  0.15905E-02    LBMOL/HR
            Old Makeup:       0.15916E-02    LBMOL/HR
            Adding Makeup:   -0.11313E-05    LBMOL/HR
            New Makeup:       0.15905E-02    LBMOL/HR

            Component Id:    7051
            Recycle Feed:     0.10310E-01    LBMOL/HR
            Recycle Product:  0.10306E-01    LBMOL/HR
            Old Makeup:       0.10310E-01    LBMOL/HR
            Adding Makeup:   -0.33489E-05    LBMOL/HR
```

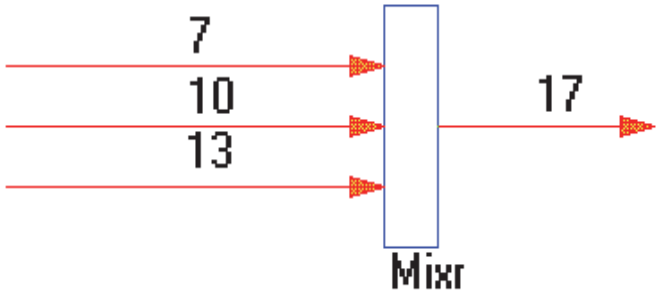
```
New Makeup:       0.10306E-01    LBMOL/HR

STREAM NUMBER    =     5
VAPOR FRACTION   =  0.00000E+00
TEMPERATURE      =   85.003       F
PRESSURE         =   30.000       PSIA
ENTHALPY         = -0.82669E+06   BTU/HR
FLOWRATE         =   44.942       LBMOL/HR
```

Chapter 35: Mixer Samples - c:\designii\samples\equipmnt\mixer

Mix.psd - Stream Mixing

The MIXer is used to mix multiple input streams. The feed streams are adiabatically flashed to the lowest inlet stream pressure.

The component flowrates of the inlet streams (7, 10, 13) are added, and the outlet stream (17) pressure is set to equal the lowest inlet pressure. The enthalpies of the feeds are added and used with the pressure and totaled flowrates to calculate the temperature of the mixture.



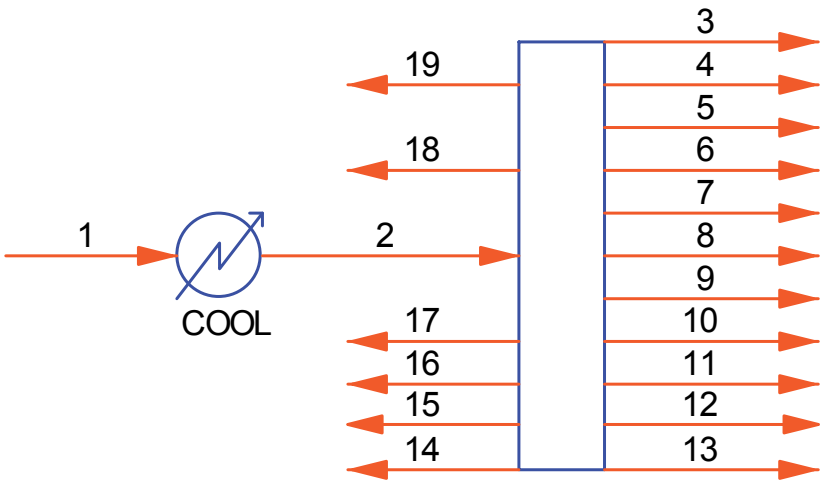

```
MIX 6 = MIXR,7,10,13,-17
```

Chapter 35

Chapter 36: Multiple Flashes Samples - c:\designii\samples\equipmnt\mulflash

Mulfla1.psd - One Inlet Multiple Outlets

This example determines the properties of the streams leaving Multiple Flashes 2 using a variety of different stream conditions.




```
*TEST ONE INLET, MULTIPLE OUTLETS
HEAt EXChanger 1 = COOL, 1, -2, TEMperature OUT = 120,
MULtiple FLAsh 2 = MULT, 2, -3, -4, -5, -6, -7, -8, -9, -10,
        -11, -12, -13,  -14, -15, -16, -17, -18, -19,-20,
        ADIabatic = 3,
        BUBble point = 4, 15,
        DEW point = 5, 16,
        HEAt ADDed = 6, 5E6, 7,-2.5E6,
        ISEntropic = 8,
        LIQuid = 9,
        LIQuid FRAction = 10, .5, 17, .2,
        DEW WAT = 14
        LIQuid FRAction (MASs) = 11, .5, 18, .75,
        VAPor = 13, 19,
        PREssure OUT = 3, 300, 4, 300, 5, 125, 6, 125, 8, 275,
                10, 200, 11, 200, 12, 14.7, 13,1 00, 14, 80,
        TEMperature OUT = 3, 200, 7, 200, 8, 275, 12, 75, 14,
                100, 15, 100, 16, 125,  17, 200, 18, 250, 19, 250,
GENeral
        COMponents = 40, 41, 42, 43, 44, 10, 12
        FLOw (LB/HR) 1 = 1000, 750, 500, 450, 600, 750, 1200
        TP 1 = 150, 125
        PENK
        BIN PAR = PENG1
END
```

The flash specifications for each stream are summarized in the output as shown below:

```
****MULTI-FLASH          ****
EQUIPMENT NO.            2
EXTERNAL NAME    MULT
STREAM TYPE      TEMP. SPEC.     PRES. SPEC.
NUMBER           DEGF    PSIA
3        ADIABATIC TEMP 200.00  300.00
4        BUBBLE TEMP CALC       NO      300.00
5        DEW TEMP CALC  NO      125.00
6        ADIABATIC TEMP NO      125.00
7        ADIABATIC PRES 200.00  NO
8        ISENTROPIC TEMP        275.00  275.00
9        SET TO VAPOR   NO      NO
10       MOLAR L/F TEMP NO      200.00
11       MASS L/F TEMP  NO      200.00
12       ISOTHERMAL     75.000  14.700
```

```
13      SET TO LIQUID  NO      100.00
14      ISOTHERMAL     100.00  80.000
15      BUBBLE PRES CALC       100.00  NO
16      DEW PRES CALC  125.00  NO
17      MOLAR L/F PRES 200.00  NO
18      MASS L/F PRES  250.00  NO
19      SET TO LIQUID  250.00  NO
20      ISOTHERMAL     NO      NO
```

Mulfla3.in – Multiflash for Bubble and Dewpoint Curves

This is a multiflash simulation input file for obtaining bubble and dew point curves.

Chapter 37: Multiple Phase Flash Samples - c:\designii\samples\equipmnt\mulphase

Meohtold.dat – Multiphase Flash Data File

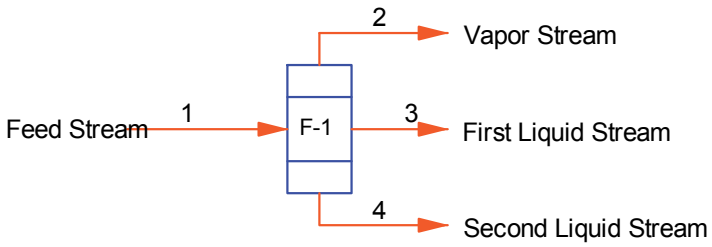
It is a ChemTran created binary data file created by the ChemTran input file METOLCT.IN for use by DESIGN II at runtime of both multiphase files below.

Metolct.in – Methanol, Water, Toluene VLE Data Regression

It is a ChemTran VLE data regression input file for methanol, water and toluene mixture for both multiphase files below.

Mulpha1.psd - Sample

The Multiple phase flash example uses ChemTran LLE data regressed file for methanol/water/toluene**.** The output for this three-phase flash is shown below. Note that the vapor fraction in the stream printout for inlet stream 1 is calculated with the standard two-phase flash technique, and does not reflect the three phase separation. However streams 2, 3, and 4 are after the three phase separator.




```
AB123.
*THREE PHASE SEPARATION
C- EQUIPMENT MODULES
MUL PHA 1 = F-1, 1,-2,-3,-4,
C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,1021,62,41,3,
  REN
  LAT
  STDD
  SUR TEN = STD
  CHE FIL = MEOH,TOL,DATA,
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 50.,5.,20.,25.,0.,
  TP (C,PSIA) 1 = 25.,14.7
  NAM STR 1 = Feed Stream
  NAM STR 2 = Vapor Stream
  NAM STR 3 = First Liquid Stream
  NAM STR 4 = Second Liquid Stream
END
```

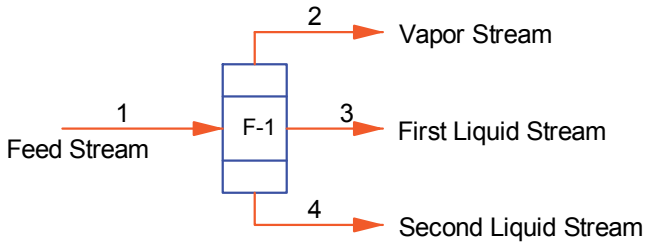
Simulation Results using equilibrium data fitted to RENon (NRTL) equation:

```
                              STREAM SUMMARY
                              --------------
STREAM NUMBER            1              2              3              4
STREAM NAME       Feed Stream    Vapor Stream   First Liquid S Second Liquid
EQUIP CONXION        FEED-F-1       F-1 -PROD      F-1 -PROD      F-1 -PROD
                   (   0)-(   1)  (   1)-(   0)  (   1)-(   0)  (   1)-(   0)
VAPOR FRACTION      0.54958         1.0000        0.00000E+00    0.00000E+00
TEMPERATURE F        77.000         77.000         77.000         77.000
PRESSURE PSIA        14.700         14.700         14.700         14.700
ENTHALPY BTU/HR    -0.72340E+06     19047.       -0.30344E+06   -0.43479E+06
V Cp  BTU/LBMOL/R    7.7493         7.7442
V Cv  BTU/LBMOL/R    5.7581         5.7529
L Cp  BTU/LBMOL/R    27.588                        33.827         22.769
V DEN LB/FT3        0.17418E-01    0.17528E-01
L DEN LB/FT3         56.496                        54.835         57.973
```

```
L S.G.(60F)    STP   0.88121                    0.87282        0.89887
V VIS CP             0.97090E-02    0.97013E-02
L VIS CP             0.61185                    0.55069        0.65934
V THC BTU/FT/HR/F    0.68039E-01    0.67611E-01
L THC BTU/FT/HR/F    0.17856                    0.95767E-01    0.32051
V MMSCF/DAY    STP   0.50046        0.50341
L GAL/MIN      STP    5.7837                     3.7077         2.0595
TOTAL MOLE WEIGHT     29.247         6.8697      82.260         36.985

HYDROGEN              50.000         49.754     0.82111E-01    0.16367
METHANOL              5.0000         2.0791     0.58598         2.3350
WATER                 20.000         1.4146      2.0467         16.539
TOLUENE               25.000         2.0336      16.965         6.0017
ETHANE               0.00000E+00    0.00000E+00  0.00000E+00  0.00000E+00
                     -----------    -----------  -----------  -----------
TOTAL     LBMOL/HR    100.00         55.281      19.680         25.039
TOTAL        LB/HR    2924.7         379.77      1618.8         926.06
```

Mulpha2.psd - Three Phase Flash Using Ternary LLE Data Fitted To RENON Equation

The MULtiple PHAse flash example is shown to rigorously separate multiple phase streams. One inlet and three outlet streams are shown. ChemTran fits a combination of vapor-liquid and liquid-liquid equilibrium data using RENon for this system. The ternary LLE parameters used for this calculation are from regression of literature data via ChemTran for Methanol/Water/Toluene VLE-LLE data.




```
AB123.
*Three Phase Flash Using Ternary LLE Data with RENON equation
C- EQUIPMENT MODULES
MUL PHA 1 = F-1, 1,-2,-3,-4,
C- GENERAL SECTION
GENERAL
  COMPONENTS = 1021,62,41,
  REN
  LAT
  STDD
  SUR TEN = STD
  CHE FIL = MEOH,TOL,DATA,
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 5.,20.,25.,
  TP (C,PSIA) 1 = 25.,14.7
  NAM STR 1 = Feed Stream
  NAM STR 2 = Vapor Stream
  NAM STR 3 = First Liquid Stream
  NAM STR 4 = Second Liquid Stream
END
```

Results using equilibrium data fitted to RENon equation:

```
                          STREAM SUMMARY
                          --------------

STREAM NUMBER            1              3              4
STREAM NAME       Feed Stream    First Liquid S Second Liquid
EQUIP CONXION        FEED-F-1       F-1 -PROD      F-1 -PROD
                   (   0)-(   1)  (   1)-(   0)  (   1)-(   0)
VAPOR FRACTION      0.00000E+00    0.00000E+00    0.00000E+00
TEMPERATURE F        77.000         77.000         77.000
PRESSURE PSIA        14.700         14.700         14.700
ENTHALPY BTU/HR    -0.82838E+06   -0.30296E+06   -0.52542E+06
V Cp  BTU/LBMOL/R
V Cv  BTU/LBMOL/R
L Cp  BTU/LBMOL/R    27.748         33.817         23.853
```

V DEN LB/FT3				
L DEN LB/FT3		56.597	54.994	57.549
L S.G.(60F)	STP	0.88203	0.87375	0.89313
V VIS CP				
L VIS CP		0.61458	0.55339	0.65204
V THC BTU/FT/HR/F				
L THC BTU/FT/HR/F		0.17959	0.96510E-01	0.28978
V MMSCF/DAY	STP			
L GAL/MIN	STP	6.4002	3.6660	2.7342
TOTAL MOLE WEIGHT		56.477	81.982	40.110
METHANOL		5.0000	0.70189	4.2981
WATER		20.000	2.1079	17.892
TOLUENE		25.000	16.735	8.2651
TOTAL	LBMOL/HR	50.000	19.545	30.455
TOTAL	LB/HR	2823.9	1602.3	1221.6

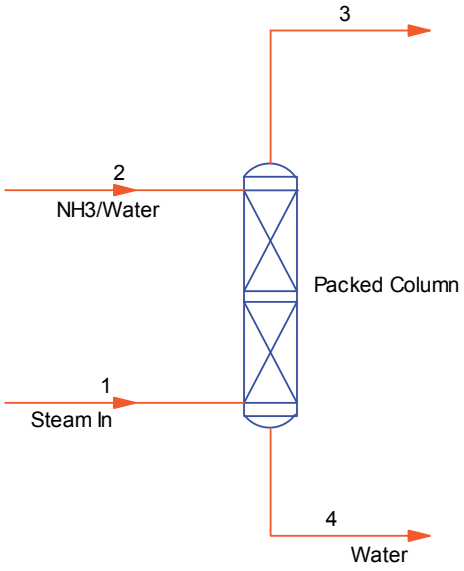
Chapter 37

Chapter 38: Packed Column Samples - c:\designii\samples\equipmnt\packedcolumn

Packed Column.psd - For Natural Gas Stream

Pressures from 50 to 600 PSIA in 100 PSI increments, after 100 PSIA. Bubble point guess is 125 F, dew point guess is 345 F.

The bubble and dew point temperatures are calculated at 50 PSIA using the initial guesses entered. The calculated bubble and dew point temperatures for each pressure are used as initial guesses for subsequent pressures. When a bubble point (dew point) fails to converge because the critical region is approached, DESIGN II continues calculating dew points (bubble points) to try to fill in the top of the envelope. A sample output (including the automatic plot of data points) is shown below and on the following pages:




```
PAC COL 2 = Packed Column, 2,1,-3,-4,
  NUM SEC = 1
  SEC DEP(FT) = 15,
  HETP(FT) = 2,
  TOP PRE(KPA) = 445
  PRE DRO(KPA) = 1,
  MAX ITE = 100
  TOL = 1e-005
```

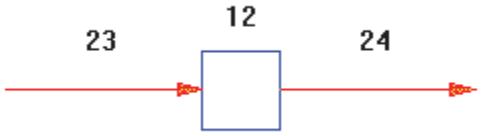
Chapter 34

Chapter 39: Phase Envelope Samples - c:\designii\samples\equipmnt\phaseenv

Phaenv.psd - For Natural Gas Stream

Pressures from 50 to 600 PSIA in 100 PSI increments, after 100 PSIA. Bubble point guess is 125 F, dew point guess is 345 F.

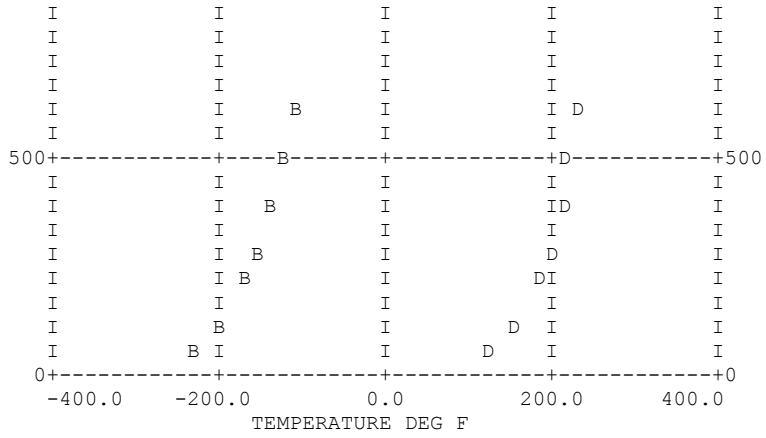
The bubble and dew point temperatures are calculated at 50 PSIA using the initial guesses entered. The calculated bubble and dew point temperatures for each pressure are used as initial guesses for subsequent pressures. When a bubble point (dew point) fails to converge because the critical region is approached, DESIGN II continues calculating dew points (bubble points) to try to fill in the top of the envelope. A sample output (including the automatic plot of data points) is shown below and on the following pages:




```
PHAse envelope 12 = ENVL, 23, -24
       PREssures = 50, 100, 4*0, 600
       TEMperature at BUBble point guess = 125
       TEMperature at DEW point guess = 345
```

DEW POINTS				BUBBLE POINTS			
TEMP F	PRESS PSIA	ENTH BTU/HR	COMP FACTOR	TEMP F	PRESS PSIA	ENTH BTU/HR	COMP FACTOR
124.6	50.00	0.1014E+07	0.9890	-227.8	50.00	-.7885E+07	0.0155
152.7	100.0	0.1315E+07	0.9808	-202.9	100.0	-.7484E+07	0.0290
182.1	200.0	0.1609E+07	0.9671	-172.7	200.0	-.6983E+07	0.0544
199.1	300.0	0.1756E+07	0.9551	-151.9	300.0	-.6624E+07	0.0789
210.5	400.0	0.1836E+07	0.9441	-135.4	400.0	-.6330E+07	0.1030
218.5	500.0	0.1875E+07	0.9337	-121.5	500.0	-.6077E+07	0.1268
224.3	600.0	0.1885E+07	0.9238	-109.3	600.0	-.5850E+07	0.1506

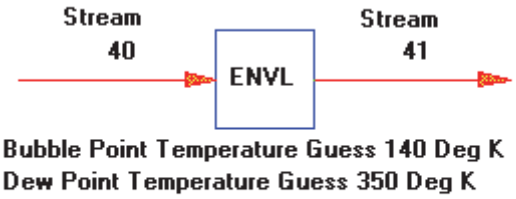
```
          PHASE ENVELOPE FOR EQUIPMENT NUMBER    2

        -400.0    -200.0         0.0         200.0      400.0
    2500+-----------+-----------+-----------+-----------+2500
        I           I           I           I           I
        I           I           I           I           I
        I           I           I           I           I
        I        D DEW POINT    I           I           I
        I        B BUBBLE POINT I           I           I
        I        C CRITICAL POINT I         I           I
        I           I           I C         I           I
        I           I           I           I           I
    2000+-----------+-----------+-----------+-----------+2000
        I           I           I           I           I
        I           I           I           I           I
 P      I           I           I           I           I
 R      I           I           I           I           I
 E      I           I           I           I           I
 S      I           I           I           I           I
 S      I           I           I           I           I
 U      I           I           I           I           I
 R  1500+-----------+-----------+-----------+-----------+1500
 E      I           I           I           I           I
        I           I           I           I           I
        I           I           I           I           I
        I           I           I           I           I
        I           I           I           I           I
 P      I           I           I           I           I
 S      I           I           I           I           I
 I      I           I           I           I           I
 A  1000+-----------+-----------+-----------+-----------+1000
        I           I           I           I           I
        I           I           I           I           I
```


500
0
-400.0
-200.0
0.0
200.0
400.0
TEMPERATURE DEG F

Chapter 40: Phase Map Samples - c:\designii\samples\equipmnt\phasemap

Phamap.psd - For Natural Gas Stream

The PHAse MAP example for calculation of dew points, bubble points and constant liquid fraction lines for hydrocarbon systems. Output includes the critical temperature and pressure of the mixture; temperature, pressure, compressibility factor and enthalpy (in molar units) for each of the bubble and dew points are calculated.



Phase Map

```
*PHASE MAP SAMPLE
PHAse MAP 1 = ENVL,40,-41,
   TEMperature BUBble (K) = 140,
   TEMperature DEW (K) = 350,
   LIQuid FRAction = .05,
GENeral
   COMponents =  46,49,2,3,4,5,6,8,10,11,12
   SOAVEK, SOAVEH, COPED
   TOTal FLOw 40 = 1000
   FLOwrate (FRA) 40 = 008,.022,.74,.094,.053,.018,.013, .011,.008,.02,.013
   TP (K,ATM) 40 = 100.0,10.0
END
```

Partial results for the PHAse MAP calculation are shown on the following pages.

```
 PHASE MAP USES SOAVE        EQ FOR K-VALUES WITHOUT BINARY PARAMETERS

  CRITICAL TEMPERATURE =    37.835F
  CRITICAL PRESSURE =  2137.873PSIA

          RESULTS FOR LIQUID FRACTION = 0.0000
        TEMP          PRES       Z-FACTOR       ENTHALPY
        F             PSIA                      BTU/LBMOLE
     ---------     ---------     -----------   -----------
  1    169.0335      147.0000     0.97431         1481.7
  2    179.9109      189.6567     0.96902         1587.1
  3    187.2051      225.1532     0.96489         1654.1
  4    195.4111      273.8093     0.95953         1724.8
  5    204.4680      342.1071     0.95247         1794.4
  6    214.1089      440.5177     0.94298         1853.3
  7    223.6129      586.3682     0.92991         1881.2
  8    230.5007      775.9407     0.91402         1847.2
  9    232.5438      890.2580     0.90479         1798.5
 10    233.3456     1039.9154     0.89287         1708.4
 11    232.8905     1135.3436     0.88529         1636.8
 12    231.3733     1253.8243     0.87579         1533.5
 13    228.0764     1401.5630     0.86369         1382.5
 14    221.8018     1585.5402     0.84783         1158.3
 15    216.9106     1693.9963     0.83784         1005.3
 16    194.9211     2027.0376     0.80167         405.69
 17    191.2072     2068.2947     0.79618         312.19
 18    186.8404     2112.8799     0.78985         204.09
 19    181.7086     2160.3696     0.78251         79.273
 20    175.6858     2209.9985     0.77401        -64.596
 21    168.6325     2260.5388     0.76413        -229.97
 22    160.4000     2310.1619     0.75262        -419.39
 23    152.0872     2350.9182     0.74094        -607.28
 24    143.7409     2383.1040     0.72908        -792.97
```

25	135.4028	2407.0703	0.71706	-975.91
26	118.8968	2431.9829	0.69252	-1331.5
27	91.0299	2411.3032	0.64782	-1916.2
28	80.7043	2384.8003	0.62995	-2128.7
29	72.5954	2357.1035	0.61540	-2294.3
30	66.8560	2333.8931	0.60478	-2410.9
31	61.4267	2309.2131	0.59447	-2520.7
32	57.0242	2287.2717	0.58592	-2609.4
33	52.9021	2265.1707	0.57776	-2692.3
34	49.3713	2245.0496	0.57064	-2763.0

RESULTS FOR LIQUID FRACTION = 1.0000

	TEMP F	PRES PSIA	Z-FACTOR	ENTHALPY BTU/LBMOLE
1	-185.6555	147.0000	4.09620E-02	-7176.5
2	-175.6548	183.7638	5.02324E-02	-7010.5
3	-169.7760	208.0970	5.62991E-02	-6911.8
4	-162.8710	239.3841	6.40387E-02	-6794.6
5	-155.8245	274.4503	7.26505E-02	-6673.7
6	-148.6529	313.5247	8.21892E-02	-6549.3
7	-141.3757	356.7969	9.27029E-02	-6421.4
8	-134.0161	404.4024	0.10423	-6290.6
9	-126.6016	456.4062	0.11679	-6157.2
10	-119.1642	512.7858	0.13040	-6021.9
11	-111.7410	573.4128	0.14503	-5885.2
12	-104.3746	638.0395	0.16064	-5748.1
13	-97.1109	706.2974	0.17715	-5611.7
14	-89.9945	777.7408	0.19444	-5476.9
15	-83.0517	852.0158	0.21244	-5344.5
16	-76.2476	929.3674	0.23118	-5214.1
17	-69.3788	1012.0061	0.25118	-5082.0
18	-61.8265	1107.4034	0.27426	-4936.5
19	-52.0469	1234.8026	0.30502	-4747.9
20	-36.7444	1432.3774	0.35273	-4452.3
21	-33.6494	1471.1569	0.36214	-4392.5
22	-30.2801	1512.7563	0.37228	-4327.2
23	-26.6112	1557.2634	0.38316	-4256.2
24	-22.6166	1604.7186	0.39481	-4178.7
25	-18.2701	1655.1027	0.40727	-4094.4
26	-13.5451	1708.3236	0.42054	-4002.5
27	-8.4156	1764.2031	0.43464	-3902.7
28	-2.8566	1822.4629	0.44945	-3794.3
29	0.2172	1853.6265	0.45749	-3734.3
30	3.2269	1883.4058	0.46526	-3675.4
31	6.5734	1915.6537	0.47376	-3609.9
32	10.0538	1948.2180	0.48246	-3541.7
33	13.6706	1980.9990	0.49127	-3470.7
34	16.8731	2009.1149	0.49898	-3407.7
35	19.5424	2031.8933	0.50531	-3355.2
36	22.7112	2058.1545	0.51272	-3292.7
37	25.2779	2078.8020	0.51863	-3242.1
38	28.1449	2101.2039	0.52515	-3185.4
39	31.0758	2123.3794	0.53171	-3127.4
40	34.1322	2145.7200	0.53846	-3066.8

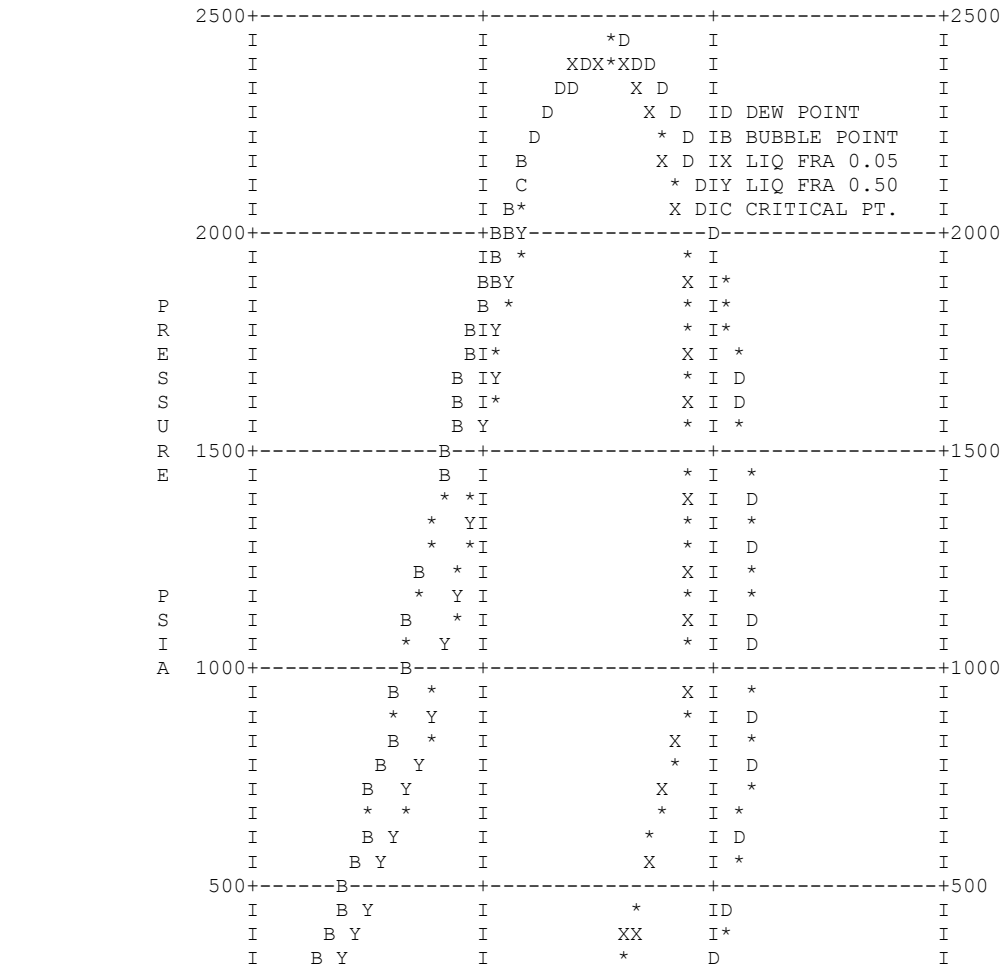
RESULTS FOR LIQUID FRACTION = 0.0500

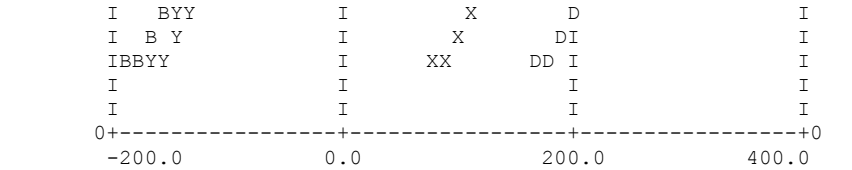
	TEMP F	PRES PSIA	Z-FACTOR VAPOR	Z-FACTOR LIQUID
1	76.8571	147.0000	0.96679	6.14464E-02
2	87.3275	180.4407	0.96169	7.38540E-02
3	95.7304	211.8893	0.95719	8.52505E-02
4	104.1975	248.4204	0.95226	9.82008E-02
5	113.9651	297.6540	0.94601	0.11522
6	125.1469	365.3569	0.93800	0.13793
7	129.9519	398.9363	0.93423	0.14893
8	146.3692	540.7693	0.91942	0.19365
9	160.5667	714.7492	0.90295	0.24522
10	166.5863	813.6807	0.89413	0.27310
11	172.4464	936.6311	0.88350	0.30646
12	177.5945	1088.7322	0.87068	0.34584
13	180.8419	1249.8259	0.85721	0.38536
14	182.0173	1416.4469	0.84309	0.42398
15	181.0038	1584.2883	0.82828	0.46061
16	177.7558	1748.3928	0.81272	0.49421
17	172.3113	1903.4572	0.79633	0.52391

18	164.7954	2044.2218	0.77903	0.54902
19	155.4165	2165.9019	0.76070	0.56909
20	144.4533	2264.5964	0.74126	0.58393
21	132.2349	2337.6162	0.72067	0.59362
22	119.1169	2383.6819	0.69880	0.59821
23	101.6701	2403.7480	0.66936	0.59766
24	82.7578	2381.0706	0.63596	0.59011

RESULTS FOR LIQUID FRACTION = 0.5000

	TEMP F	PRES PSIA	Z-FACTOR VAPOR	Z-FACTOR LIQUID
1	-167.8934	147.0000	0.86999	4.56256E-02
2	-155.6712	188.0867	0.84818	5.67867E-02
3	-148.3788	215.9389	0.83462	6.41971E-02
4	-139.7037	252.5057	0.81800	7.37749E-02
5	-130.7206	294.4921	0.80029	8.45997E-02
6	-121.4334	342.5307	0.78149	9.68037E-02
7	-111.8496	397.2935	0.76163	0.11053
8	-101.9808	459.4902	0.74078	0.12593
9	-91.8431	529.8696	0.71901	0.14316
10	-81.4568	609.2283	0.69646	0.16242
11	-70.8453	698.4398	0.67335	0.18388
12	-60.0313	798.5201	0.64997	0.20779
13	-49.0286	910.7736	0.62680	0.23444
14	-37.8196	1037.0958	0.60462	0.26425
15	-26.3096	1180.5541	0.58464	0.29788
16	-14.2406	1346.2754	0.56880	0.33644
17	-1.1061	1541.6832	0.55976	0.38154
18	6.0504	1652.4158	0.55852	0.40698
19	13.6190	1770.9336	0.55947	0.43432
20	21.4221	1893.2538	0.56197	0.46294
21	28.9804	2010.8372	0.56455	0.49152
22	35.4363	2110.0493	0.56518	0.51803
23	36.8454	2131.5225	0.56464	0.52448



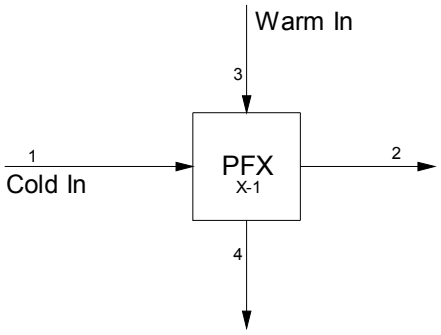


MAX. TEMPERATURE = 233.35 F MAX. PRESSURE = 2433.8 PSIA

Chapter 41: Plate Fin Exchanger Samples - c:\designii\samples\equipmnt\platefin

Pfx1.psd - Simple Gas/ Gas Exchanger

This is a simple case of a gas/gas contact using plate-fin exchanger. The cold side residue gas is used to cool the incoming warm side inlet feed natural gas. In addition to heat exchange calculations, the plate-fin exchanger rating is also performed.



Rating Specifications

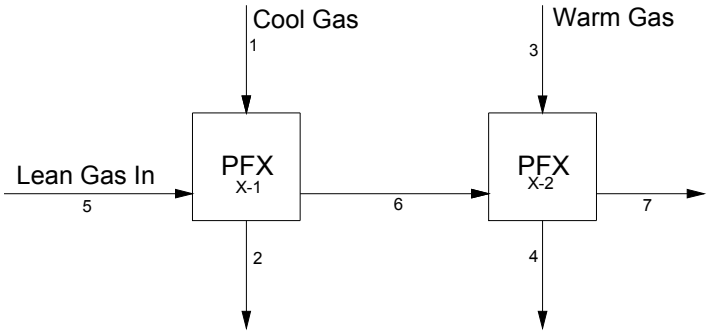
Total Cores	1	
Parallel:	1	
Parting Sheet or Plate Thickness	0.10	
	Warm Side	Cold Side
Number of Layers per Core	44	45
Fin Type	SERR	PERF
Fin Length/ Perforation percent (in/pct)	0.25	3.3
Fin Height, Inches	0.281	0.36
Fin Thickness, Inches	0.016	0.008
Fin Spacing, Fins per Inch	17	25
Effective Passage Width, Inches	39	39
Effective Passage Length, Inches	165.28	155.55
Fin Factor based on Layer Arrangement	1	1

```
AB123.
*Gas/Gas Exchanger

C- EQUIPMENT MODULES

PFX 1 = X-1, 3,1,-4,-2,
  NUM HOT = 1
  TEM APP HOT(F) = 5.
  OUT STR PHA = 0,0,
  Q-T
  DEL PRE(PSI) = 5.,5.,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,1.E-003,
  UFA = 1.,1.,
  PLA THI(IN) = 0.1
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 165.28,155.55,
  EFF WID(IN) = 39.,39.,
  NUM LAY = 44,45,
  FIN TYP = 1,2,
  FIN CON(BTU/HR/FT/F) = 100.
  FIN HEI(IN) = 0.281,0.36,
  FIN LEN(IN) = 0.25,0.25,
```

```
  FIN THI(IN) = 1.65E-002,8.E-003,
  FIN PIT(1/IN) = 17.,17.,

  FIN PER = 5.,3.3,
  FIN FAC = 1.,1.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,46,2,3,4,5,6,7,8,10,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 9.9E-003,5.8E-003,0.9755,8.4E-003,
  4.E-004,5*0.,
  TOT FLO (LBMOL/HR) 1 = 11696.9
  TP (F,PSIA) 1 = -26.63,309.
  NAM STR 1 = Cold In
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 1.37E-002,5.1E-003,0.8671,6.82E-002,
  2.27E-002,5.4E-003,6.9E-003,
  3.3E-003,2.3E-003,5.3E-003,
  TOT FLO (LBMOL/HR) 3 = 8550.25
  TP (F,PSIA) 3 = 100.,910.
  NAM STR 3 = Warm In
  NAM STR 4 = Strm 4
END
```

Pfx2.psd - Gas/Gas Exchangers

This is another example of plate-fin exchanger. The cold side residue gas is used to contact with two warm side inlet feed natural gases. Two plate-fin exchangers are used to represent this scheme with the first plate-fin exchanger cooling the first inlet cool gas followed by a second plate-fin exchanger cooling the warm feed gas. In addition to heat exchange calculations, the plate-fin exchanger rating is also performed.



Rating Specifications

Total Cores	1		
Parallel	1		
Parting Sheet or Plate Thickness, Inches	0.125		
	Warm Gas	Cool Gas	Lean Gas
Number of Layers per Core	90	90	60
Fin Type	PERF	PERF	PERF
Fin Length/ Perforation percent	3.30%	3.30%	3.30%
Fin Height, Inches	0.281	0.281	0.281
Fin Thickness, Inches	0.016	0.016	0.008
Fin Spacing, Fins per Inch	18	18	18
Effective Passage Width, Inches	22	22	22
Effective Passage Length, Inches	49	51	121
Fin Factor based on Layer Arrangement	1	0.5	1

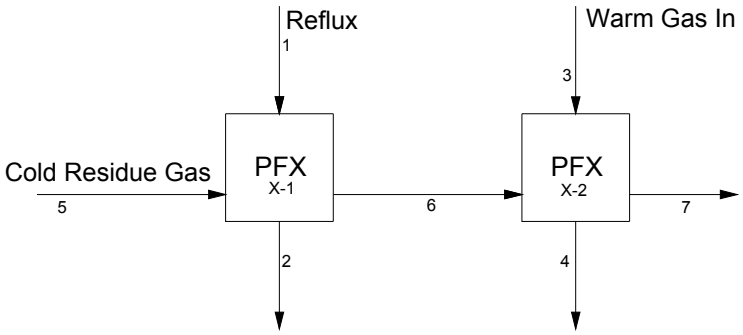
```
AB123.
*Gas/Gas Exchangers

C- EQUIPMENT MODULES
```

```
PFX 1 = X-1, 1,5,-2,-6,
  NUM HOT = 1
  TEM OUT HOT(F) = -5.,
  OUT STR PHA = 0,0,
  Q-T
  DEL PRE(PSI) = 5.,10.,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,1.E-003,
  UFA = 1.,1.,
  PLA THI(IN) = 0.125
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 51.,60.5,
  EFF WID(IN) = 22.,22.,
  NUM LAY = 90,60,
  FIN TYP = 2,2,
  FIN CON(BTU/HR/FT/F) = 100.
  FIN HEI(IN) = 0.281,0.281,
  FIN LEN(IN) = 0.25,0.25,
  FIN THI(IN) = 1.6E-002,1.6E-002,
  FIN PIT(1/IN) = 18.,18.,
  FIN PER = 3.3,3.3,
  FIN FAC = 1.,0.5,

PFX 2 = X-2, 3,6,-4,-7,
  NUM HOT = 1
  TEM OUT HOT(F) = 63.,
  OUT STR PHA = 0,0,
  Q-T
  DEL PRE(PSI) = 5.,10.,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,1.E-003,
  UFA = 1.,1.,
  PLA THI(IN) = 0.125
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 49.,60.5,
  EFF WID(IN) = 22.,22.,
  NUM LAY = 90,60,
  FIN TYP = 2,2,
  FIN CON(BTU/HR/FT/F) = 100.
  FIN HEI(IN) = 0.281,0.281,
  FIN LEN(FT) = 2.083E-002,2.083E-002,
  FIN THI(IN) = 1.6E-002,1.6E-002,
  FIN PIT(1/IN) = 18.,18.,
  FIN PER = 3.3,3.3,
  FIN FAC = 1.,1.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,46,2,3,4,5,6,7,8,10,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 5.4,108.,10499.,1699.,705.,82.,120.,
  19.,15.,3.,
  TP (F,PSIA) 1 = 30.,815.
  NAM STR 1 = Cool Gas
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 6.5,119.8,11983.,2205.,1168.,18.3,
  313.,77.,83.,63.,
  TP (F,PSIA) 3 = 110.,825.
  NAM STR 3 = Warm Gas
  NAM STR 4 = Strm 4
  FLO(LBMOL/HR)5 = 5.4,131.4,11914.5,1398.,343.,23.5,
```

```
  26.9,2.2,1.4,0.1,
  TP (F,PSIA) 5 = -30.,805.
  NAM STR 5 = Lean Gas In
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Pfx3.psd - Gas/Gas and-Reflux Condenser

The sample problem models a gas/gas and reflux condenser using two plate-fin exchangers. The cold side residue gas is used to contact with reflux and warm gas inlet. Two plate-fin exchangers are used to represent this scheme with the first plate-fin cooling the reflux gas followed by second plate-fin cooling the warm feed gas. In addition to heat exchange calculations, the plate-fin exchanger rating is also performed.



Rating Specifications

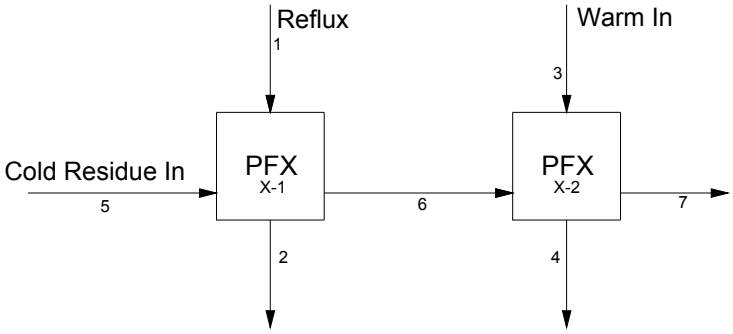
Total Cores	1		
Parallel	2		
Parting Sheet or Plate Thickness, Inches	0.09		
	Warm Gas	Reflux	Residue Gas
Number of Layers per Core	39	39	78
Fin Type	PERF	PERF	SERR
Fin Length/ Perforation percent	3.30%	3.30%	1/8 "
Fin Height, Inches	0.281	0.281	0.380
Fin Thickness, Inches	0.010	0.010	0.010
Fin Spacing, Fins per Inch	32	32	22.5
Effective Passage Width, Inches	39	39	39
Effective Passage Length, Inches	166.4	62	241.1
Fin Factor based on Layer Arrangement	1	1	1

```
AB123.
*Gas/Gas Exchanger-Reflux Condenser

C- EQUIPMENT MODULES

PFX 1 = X-1, 1,5,-2,-6,
  NUM HOT = 1
  TEM OUT HOT(F) = -130.,
  OUT STR PHA = 0,0,
  Q-T
  DEL PRE(PSI) = 5.,6.4,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,1.E-003,
  UFA = 1.,1.,
  PLA THI(IN) = 9.E-002
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 62.,120.55,
  EFF WID(IN) = 39.,39.,
  NUM LAY = 39,78,
  FIN TYP = 2,1,
  FIN CON(BTU/HR/FT/F) = 100.
  FIN HEI(IN) = 0.281,0.38,
  FIN LEN(IN) = 0.125,0.125,
```

```
  FIN THI(IN) = 1.E-002,1.E-002,
  FIN PIT(1/IN) = 32.,22.5,
  FIN PER = 3.3,3.3,
  FIN FAC = 1.,1.,

PFX 2 = X-2, 3,6,-4,-7,
  NUM HOT = 1
  TEM OUT HOT(F) = -28.,
  OUT STR PHA = 0,0,
  Q-T
  DEL PRE(PSI) = 8.4,6.4,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,1.E-003,
  UFA = 1.,1.,
  PLA THI(IN) = 9.E-002
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 166.4,120.5,
  EFF WID(IN) = 39.,39.,
  NUM LAY = 39,78,
  FIN TYP = 2,1,
  FIN CON(BTU/HR/FT/F) = 100.
  FIN HEI(IN) = 0.281,0.38,
  FIN LEN(IN) = 0.125,0.125,
  FIN THI(IN) = 1.E-002,1.E-002,
  FIN PIT(1/IN) = 32.,22.5,
  FIN PER = 3.3,3.3,
  FIN FAC = 1.,1.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 3.8E-003,1.9E-003,0.9178,4.76E-002,
  2.09E-002,2.5E-003,3.7E-003,
  7.E-004,6.E-004,5.E-004,
  TOT FLO (LBMOL/HR) 1 = 11933.15
  TP (F,PSIA) 1 = -27.,1038.
  NAM STR 1 = Reflux
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 3.7E-003,1.9E-003,0.8977,5.11E-002,
  2.7E-002,4.2E-003,7.E-003,
  2.E-003,1.9E-003,3.5E-003,
  TOT FLO (LBMOL/HR) 3 = 20367.05
  TP (F,PSIA) 3 = 74.3,1048.
  NAM STR 3 = Warm Gas In
  NAM STR 4 = Strm 4
  FLO (FRA) 5 = 4.1E-003,9.E-004,0.9879,6.5E-003,
  5.E-004,1.E-004,4*0.,
  TOT FLO (LBMOL/HR) 5 = 30512.52
  TP (F,PSIA) 5 = -135.2,426.5
  NAM STR 5 = Cold Residue Gas
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Pfx4.psd - Gas/Gas and-Reflux Condenser (Rejection)

This simulation models both a gas/gas and reflux condenser with rejection using two plate-fin exchangers. The cold side residue gas is used to contact with reflux and warm gas inlet. Two plate-fin exchangers are used to represent this scheme with the first plate-fin cooling the reflux gas followed by second plate-fin cooling the warm feed gas. In addition to heat exchange calculations, the plate-fin exchanger rating is also performed using required specifications.



Rating Specifications

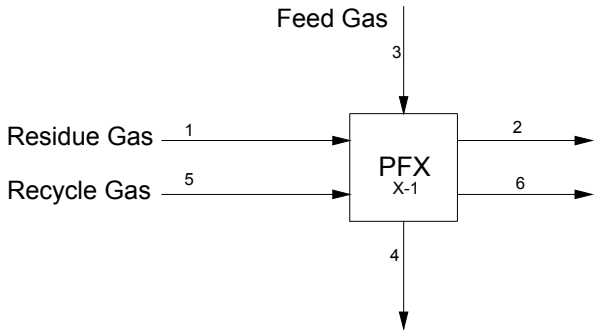
Total Cores	1		
Parallel	2		
Parting Sheet or Plate Thickness, Inches	0.125		
	Warm In	Reflux	Residue Gas
Number of Layers per Core	59	59	60
Fin Type	SERR	SERR	PERF
Fin Length/ Perforation percent	.875	.875	3.3%
Fin Height, Inches	0.281	0.281	0.380
Fin Thickness, Inches	0.016	0.016	0.008
Fin Spacing, Fins per Inch	19.4	19.4	25
Effective Passage Width, Inches	33	33	33
Effective Passage Length, Inches	92	60	201
Fin Factor based on Layer Arrangement	1	.5	.5

```
AB123.
*Gas/Gas Exchanger-Reflux Condenser (Rejection)
C- EQUIPMENT MODULES

PFX 1 = X-1, 1,5,-2,-6,
   NUMber of HOT streams =1
   TEMperature OUT HOT stream (F) = -104
   EFFective passage LENgth (IN) = 60, 100.5,
   EFFective passage WIDth (IN) = 33, 33,
   NUMber in PARallel =1
   NUMber in SERies =1
   NUMber of LAYers = 59,60,
   RATing on
   FOULing FACtor = 0.001, 0.001,
   U-M = 2
   H-M = 2
   FIN TYPe = 1, 2,
   FIN HEIght (IN) = .281, .281,
   FIN offset LENgth (IN) = .875,
   FIN THIckness (IN) = .016,.008,
   FIN PITch (1/IN)= 19.4, 25,
   FIN PERforation = 3.3
   FIN FACtor = 1,0.5
   PLAte THIckness (IN) =0.125
   DELta pressure (PSIA) = 5, 10
   Q-T

PFX 2 = X-2, 3,6,-4,-7,
   NUMber of HOT streams =1
   EFFective passage LENgth (IN) = 92, 100.5,
   EFFective passage WIDth (IN) = 33, 33,
   NUMber in PARallel =1
   NUMber in SERies =1
   NUMber of LAYers = 59,60,
   RATing on
   FOULing FACtor = 0.001, 0.001,
   U-M = 2
   H-M = 2
   FIN TYPe = 1, 2,
   FIN HEIght (IN) = .281, .281,
   FIN offset LENgth (IN) = .875,
   FIN THIckness (IN) = .016,.008,
   FIN PITch (1/IN)= 19.4, 25,
   FIN PERforation = 3.3
```

```
   FIN FACtor = 1, 0.5,
   PLAte THIckness (IN) =0.125
   DELta pressure (PSIA) = 5, 10
   Q-T

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 8.E-004,3.E-004,0.9285,3.94E-002,2.12E-002,4.1E-003,3.4E-003,
  1.E-003,8.E-004,2.E-004,
  TOT FLO (LBMOL/HR) 1 = 7638.25
  TP (F,PSIA) 1 = -15.,953.5
  NAM STR 1 = Reflux
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 8.E-004,3.E-004,0.9125,4.12E-002,2.55E-002,6.E-003,5.7E-003,
  2.4E-003,2.3E-003,3.2E-003,
  TOT FLO (LBMOL/HR) 3 = 29044.4
  TP (F,PSIA) 3 = 79.2,963.5
  NAM STR 3 = Warm In
  NAM STR 4 = Strm 4
  FLO (FRA) 5 = 8.E-004,3.E-004,0.956,4.12E-002,1.6E-003,1.E-004,4*0.,
  TOT FLO (LBMOL/HR) 5 = 31443.09
  TP (F,PSIA) 5 = -113.6,372.
  NAM STR 5 = Cold Residue In
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Pfx5.psd - Brazed Aluminum Exchanger-GPSA Example

This example models a gas/gas exchanger using one plate-fin exchanger. The cold side residue (205 psia & -106 F) gas and recycle (285 psia & -106 F) gas are both used to contact warm feed gas (at 810 psia & 120 F). Warm end temperature approaches are typically designed from 5 to 10 F. In addition to heat exchange calculations the plate-fin exchanger rating is also performed.



Rating Specifications

Total Cores	1		
Parallel	1		
Parting Sheet or Plate Thickness, Inches	0.08		
	Feed Gas	Residue Gas	Recycle Gas
Number of Layers per Core	21	30	12
Fin Type	SERR	SERR	SERR
Fin Length/ Perforation percent	.25	.25	.25
Fin Height, Inches	0.28	0.28	0.28
Fin Thickness, Inches	0.016	0.016	0.016
Fin Spacing, Fins per Inch	17	17	17
Effective Passage Width, Inches	23.375	23.375	23.375
Effective Passage Length, Inches	162.375	162.375	162.375
Fin Factor based on Layer Arrangement	1	.5	.5

```
AB123.
```

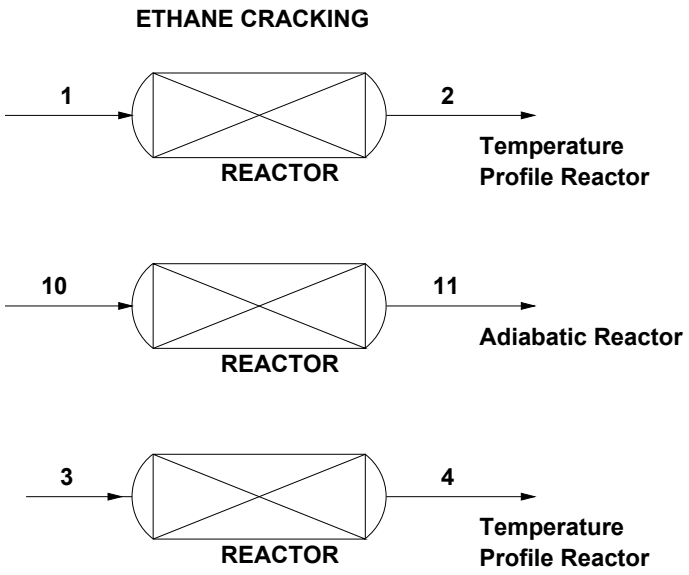
```
*Brazed Aluminum Exchanger-From GPSA

C- EQUIPMENT MODULES

PFX 1 = X-1, 3,1,5,-4,-2,-6,
  NUM HOT = 1
  TEM OUT COLD(F) = 2*113.,
  OUT STR PHA = 0,2*0,
  Q-T
  DEL PRE(PSI) = 10.,2*7.,
  RAT
  NUM INC = 5
  U-M = 2
  H-M = 2
  FOU FAC(1/BTU/FT2/HR/F) = 1.E-003,2*1.E-003,
  UFA = 1.,2*1.,
  PLA THI(IN) = 8.E-002
  PLA CON(BTU/HR/FT/F) = 100.
  NUM PAR = 1
  NUM SER = 1
  EFF LEN(IN) = 162.375,2*162.375,
  EFF WID(IN) = 23.375,2*23.375,
  NUM LAY = 21,30,12,
  FIN TYP = 1,1,1,
  FIN CON(BTU/HR/FT/F) = 100.

  FIN HEI(IN) = 0.28,2*0.28,
  FIN LEN(IN) = 0.25,2*0.25,
  FIN THI(IN) = 1.6E-002,2*1.6E-002,
  FIN PIT(1/IN) = 17.,2*17.,
  FIN PER = 5.,2*5.,
  FIN FAC = 1.,2*0.5,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,
  PENK
  PENH
  STDD
  SUR TEN = STD
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO (FRA) 1 = 4.5E-002,4.E-003,0.937,1.3E-002,
  1.E-003,4*0.,
  TOT FLO (LBMOL/HR) 1 = 1852.83
  TP (F,PSIA) 1 = -106.,205.
  NAM STR 1 = Residue Gas
  NAM STR 2 = Strm 2
  FLO (FRA) 3 = 4.E-002,6.E-003,0.838,6.3E-002,
  3.1E-002,4.E-003,1.E-002,2.E-003,
  6.E-003,

  TOT FLO (LBMOL/HR) 3 = 2086.69
  TP (F,PSIA) 3 = 120.,810.
  NAM STR 3 = Feed Gas
  NAM STR 4 = Strm 4
  FLO (FRA) 5 = 4.E-002,4.E-003,0.925,2.1E-002,1.E-002,4*0.,
  TOT FLO (LBMOL/HR) 5 = 858.88
  TP (F,PSIA) 5 = -106.,285.
  NAM STR 5 = Recycle Gas
  NAM STR 6 = Strm 6
END
```

Chapter 42: Plug Flow Reactor Samples - c:\designii\samples\equipmnt\plugreac

Pfr.psd - Plug Flow Reactor Sample

The PLUg FLOw REActor (PFR) example to calculate single-phase reactions using specified information on the reactor dimensions, stoichiometric coefficients of the reactants and products, and power law rate expressions.

```
AB123.
*ETHANE CRACKING
PLUg REActor 1 = REACTOR,1,-2,
   DIAmeter (IN) = 2.0,
   LENgth (M) = 95,
   TEMperature PROfile (C)=.1,750, .2,775, .3, 780, .4,785, .5,900, .6,800, .8,825, 1.0,850,
   DELta pressure (ATM) = 2.0,
   R 1 = (1*3 = 1*22 + 1*1),
   R 2 = (1.0*22 + 1.*1 = 1*3),
   R 3 = (2*3 = 1*4 + 1*2),
   R 4 = (1*23 = 1*65 + 1*2),
   R 5 = (1*65 + 1*2 = 1*23),
   R 6 = (1*22 + 1*65 = 1*28),
   R 7 = (1*22 + 1*3 = 1*23 + 1*2),
   ORDer 1 = 3,1,
   ORDer 2 = 22,1,1,1,
   ORDer 3 = 3,1,
   ORDer 4 = 23,1,
   ORDer 5 = 65,1,2,1,
   ORDer 7 = 22,1,3,1,
   ORDer 6 = 65,1,22,1,
   NUMber of INCrements = 10,
   FREquency (M3,KGMOL,SEC) = 2,8.75E8,
        1,4.65E13, 5,5.87E4,3,3.85E11,
        4,9.81E8,6,1.03E12,7,7.08E,3,
   ACTivation energy (KCAL/KGMOL) = 6,41260,
                5,7040,  1,65120,
                7,60430, 3,65250,
                2,32690, 4,36920,
   TYPe = TEMperature PROfile

PLUg REActor 2 = REACTOR,10,-11,
       DIAmeter (IN) = 2.0,
       LENgth (M) = 95,
       HEAt added = 4.4E6,
       DELta pressure (ATM) = 2.0,
       R 1 = (1*3 = 1*22 + 1*1),
       R 2 = (1.0*22 + 1.*1 = 1*3),
       R 3 = (2*3 = 1*4 + 1*2),
       R 4 = (1*23 = 1*65 + 1*2),
       R 5 = (1*65 + 1*2 = 1*23),
       R 6 = (1*22 + 1*65 = 1*28),
       R 7 = (1*22 + 1*3 = 1*23 + 1*2),
       ORDer 2 = 22,1,1,1,
       ORDer 3 = 3,1,
       ORDer 1 = 3,1,
       ORDer 4 = 23,1,
       ORDer 5 = 65,1,2,1,
       ORDer 7 = 22,1,3,1,
       ORDer 6 = 65,1,22,1,
       NUMber of INCrements = 10,
       FREquency (M3,KGMOL,SEC) = 2,8.75E8,1,4.65E13,
       5,5.87E4,3,3.85E11, 4,9.81E8,6,1.03E12,7,7.08E13,
       ACTivation energy (KCAL/KGMOL) = 6,41260,
               5,7040, 1,65120,
               7,60430, 3,65250,
               2,32690, 4,36920,
       TYPe = ADIabatic,

PLUg REActor 3 = REACTOR,3,-4,
       DIAmeter (IN) = 2.0,
       LENgth (M) = 95,
       TEMperature PROfile (C) = .1,750,.2,775,.3,780,
```



7 Reactions
2 Inch Diameter
95 Meter Length

```
       .4,785,.5,900, .6,800,.8,825,1.0,850,
       DELta pressure(ATM) = 2.0,
       R 1 = (1*3 = 1*22 + 1*1), R 2 = (1.0*22 + 1.*1 = 1*3),
       R 3 = (2*3 = 1*4 + 1*2), R 4 = (1*23 = 1*65 + 1*2),
       R 5 = (1*65 + 1*2 = 1*23), R 6 = (1*22 + 1*65 = 1*28),
       R 7 = (1*22 + 1*3 = 1*23 + 1*2),
       TYPe = TEMperature PROfile
F-       DURING
F-       RATE1=1.674E17*EXP(-1.1722E5/(1.987*TEM))*CONC3
F-       RATE2=5.0457E13*EXP(-5.8845E4/(1.987*TEM))*CONC22*CONC1
F-       RATE3=1.386E15*EXP(-1.1746E5/(1.987*TEM))*CONC3
F-       RATE4=3.5316E12*EXP(-6.6459E4/(1.987*TEM))*CONC23
F-       RATE5=3.385E9*EXP(-1.2673E4/(1.987*TEM))*CONC65*CONC2
F-       RATE6=5.9395E16*EXP(-7.4271E4/(1.987*TEM))*CONC65*CONC22
F-       RATE7=4.0827E18*EXP(-1.0878E5/(1.987*TEM))*CONC22*CONC3
GENeral
       COMponents = 1,2,3,4,22,23,28,65,62
       FLOw 1 = 0,1,100,1,0,3*1,20
       TP (C,ATM) 1 = 630,3
       FLOw 3 = FLOw 1,TP 3 = TP 1
       FLOw 7 = FLOw 1,TP 7 = TP 1
       FLOw 10 = FLOw 1,TP 10 = TP 1
       PENK,PENH,BIN PAR = PENG1
       METric UNIts OUT
END
```

Details for PLUg REActor 2 are shown below.

```
REACTOR TYPE = ADIABATIC

     DIAMETER   M        =     0.05    LENGTH   M        =  95.000
     HEAT ADDED KCAL/HR  = 1.10873E+06
     NO. OF REACTIONS     =   7


 DISTANCE FROM ENTRANCE M     0.0000E+00   9.500       19.00       28.50
 PROCESS STREAM CONDITIONS
 TEMPERATURE C                 630.00      710.15      769.30      792.56
 PRESSURE    KG/CM2            3.100       2.893       2.686       2.480
 DUTY        KCAL/HR          0.0000E+00  1.1087E+05  2.2175E+05  3.3262E+05
 COMPONENT FLOWRATES  KG/HR
 --------------------
 HYDROGEN                     0.0000E+00  0.2135       2.010       6.603
 METHANE                       7.277       8.087       10.46       13.52
 ETHANE                        1364.       1361.       1333.       1263.
 PROPANE                       20.00       20.04       20.34       21.13
 ETHYLENE                     0.0000E+00   2.964       27.60       88.68
 PROPYLENE                     19.09       17.00       11.39       6.847
 1,3 BUTADIENE                 24.53       24.54       25.04       28.74
 ACETYLENE                     11.81       13.10       16.43       18.31
 WATER                         163.4       163.4       163.4       163.4
 --------------------        ----------- ----------- ----------- -----------
 TOTAL                         1610.       1610.       1610.       1610.

 DISTANCE FROM ENTRANCE M      76.00       85.50       95.00
 PROCESS STREAM CONDITIONS
 TEMPERATURE C                 842.15      853.50      866.08
 PRESSURE    KG/CM2            1.447       1.240       1.033
 DUTY        KCAL/HR          8.8699E+05  9.9786E+05  1.1089E+06
 COMPONENT FLOWRATES  KG/HR
 --------------------
 HYDROGEN                      35.63       41.17       46.56
 METHANE                       33.76       38.14       42.15
 ETHANE                        805.5       717.7       632.6
 PROPANE                       26.36       27.41       28.43
 ETHYLENE                      451.7       521.3       589.7
 PROPYLENE                     6.961       6.819       6.418
 1,3 BUTADIENE                 76.41       84.38       91.31
 ACETYLENE                     10.20       9.657       9.387
 WATER                         163.4       163.4       163.4
 --------------------        ----------- ----------- -----------
 TOTAL                         1610.       1610.       1610.

                     STREAM SUMMARY (CONTINUED)
                     --------------------------

STREAM NUMBER           10             11
STREAM NAME       Strm 10        Strm 11
EQUIP CONXION        FEED-REAC      REAC-PROD
```

```
                   (   0)-(   2)  (   2)-(   0)
VAPOR FRACTION       1.0000         1.0000
TEMPERATURE C        630.00         866.08
PRESSURE KG/CM2      3.0997         1.0332
ENTHALPY KCAL/HR    0.66404E+06    0.10383E+07
V Cp  KCAL/KGMO/C    23.402         16.914
V Cv  KCAL/KGMO/C    21.401         14.926
L Cp  KCAL/KGMO/C
V DEN KG/M3          1.1492        0.21599
L DEN KG/M3
L S.G.(60F)   STP
V VIS CP            0.25877E-01    0.31724E-01
L VIS CP
V THC KCAL/M/HR/C   0.98373E-01    0.16146
L THC KCAL/M/HR/C
V M3(NTP)/HR        0.00000E+00    0.00000E+00
L M3(NTP)/HR
TOTAL MOLE WEIGHT    28.396         20.200

L-LIQUID PHASE (EXCLUDING IMMISCIBLE WATER), V-VAPOR PHASE
Cp,Cv-HEAT CAPACITY, DEN-DENSITY, VIS-VISCOSITY, THC-THERMAL CONDUCTIVITY

FLOW RATES    KGMOL/HR

HYDROGEN            0.00000E+00     23.096
METHANE             0.45359         2.6270
ETHANE               45.359         21.040
PROPANE             0.45359        0.64486
ETHYLENE            0.00000E+00     21.021
PROPYLENE           0.45359        0.15254
1,3 BUTADIENE       0.45359         1.6883
ACETYLENE           0.45359        0.36055
WATER                9.0719         9.0719
                    -----------    -----------
TOTAL    KGMOL/HR    56.699         79.702
TOTAL       KG/HR    1610.0         1610.0

          + EQUIPMENT SUMMARY - INDIVIDUAL DETAILS +  (CONTINUED)

****PLUG FLOW REACTOR   ****
EQUIPMENT NO.           2
EXTERNAL NAME        REACTOR
TYPE OF PFR =        DUTY SPEC.
DIAMETER    M             5.08000E-02
LENGTH  M                  95.000
PRESSURE DROP   KG/CM2      2.0665
U   KCAL/HR/M2/C           244.10
HEAT ADDED  KCAL/HR       1.10873E+06
CALC. DUTY  KCAL/HR       1.10887E+06
NUMBER OF REACTIONS =   7
REACTION NO.   1
  1.00*ETHANE
                    ----------->
  1.00*ETHYLENE         +  1.00*HYDROGEN
REACTION NO.   2
  1.00*ETHYLENE         +  1.00*HYDROGEN
                    ----------->
  1.00*ETHANE
REACTION NO.   3
  2.00*ETHANE
                    ----------->
  1.00*PROPANE          +  1.00*METHANE
REACTION NO.   4
  1.00*PROPYLENE
                    ----------->
  1.00*ACETYLENE        +  1.00*METHANE
REACTION NO.   5
  1.00*ACETYLENE        +  1.00*METHANE
                    ----------->
  1.00*PROPYLENE
REACTION NO.   6
  1.00*ETHYLENE         +  1.00*ACETYLENE
                    ----------->
  1.00*1,3 BUTADIENE
REACTION NO.   7
  1.00*ETHYLENE         +  1.00*ETHANE
                    ----------->
  1.00*PROPYLENE        +  1.00*METHANE
```

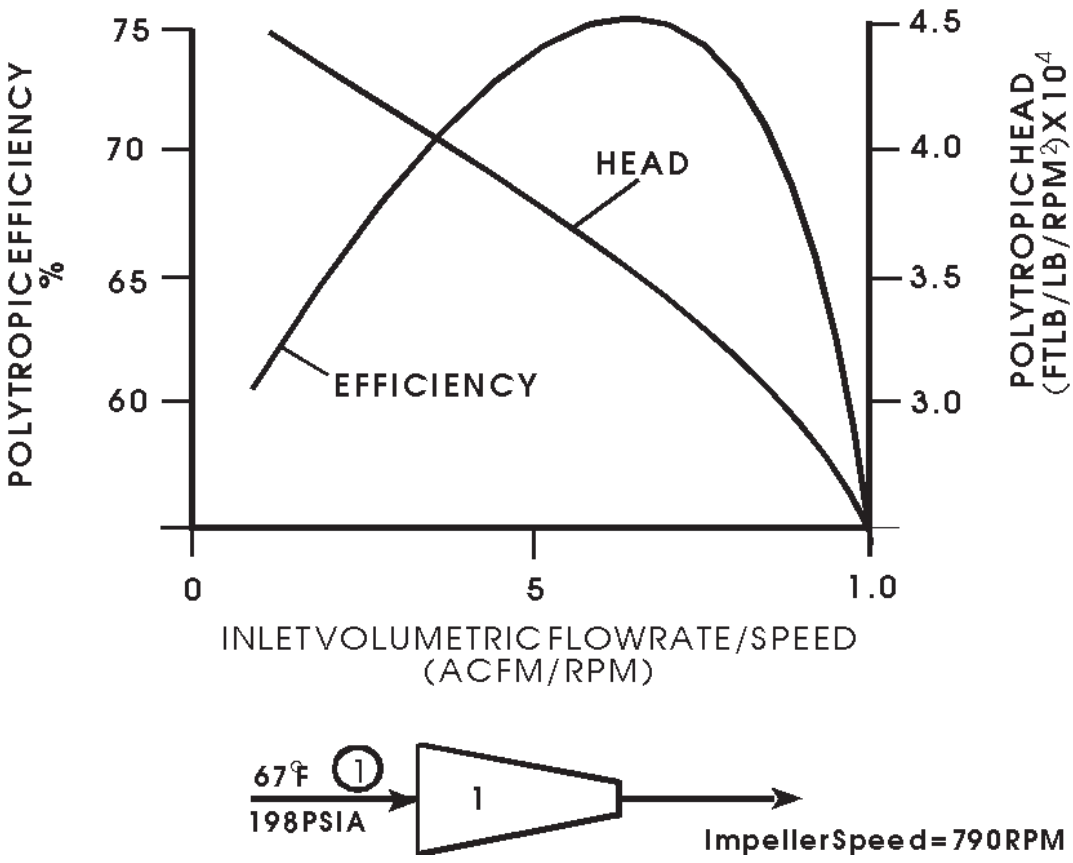
```
 STREAM NUMBER   10     CONNECTING  (   0)  FEED
 NAME: Strm 10                  TO  (   2)  REACTOR
 P.R. K-VALUES    P.R. ENTHALPY    P.R. DENSITY    FLOW RATES /HR
                     TOTAL   LIQUID    VAPOR    TOTAL    TOTAL
 COMPONENT NAME      KGMOL    KGMOL    KGMOL     KG    MOL PCT   KVALUE
 --------- ----    -------- -------- -------- -------- -------- --------
HYDROGEN               0.       0.       0.       0.       0.
METHANE             0.453593    0. 0.453593  7.27698  0.80000
ETHANE               45.3593    0.  45.3593  1363.86   80.000
PROPANE             0.453593    0. 0.453593  20.0007  0.80000
ETHYLENE               0.       0.       0.       0.       0.
PROPYLENE           0.453593    0. 0.453593  19.0863  0.80000
1,3 BUTADIENE       0.453593    0. 0.453593  24.5339  0.80000
ACETYLENE           0.453593    0. 0.453593  11.8097  0.80000
WATER                9.07186    0.  9.07186  163.433   16.000
                    -------- -------- -------- -------- --------
TOTAL                56.6991    0.  56.6991  1610.00  100.000
TOTAL   KG           1610.00    0.  1610.00

STREAM NUMBER   10
TEMPERATURE DEG C      630.00
PRESSURE    KG/CM2     3.0997
ENTHALPY    KCAL/HR   6.64043E+05
VAPOR FRACTION         1.0000

PROPERTIES              TOTAL       VAPOR
----------------      ----------  ----------
FLOWRATE   KGMOL/HR      56.6991     56.6991
MOLECULAR WT.            28.3956     28.3956
ENTHALPY KCAL/KGMOL   11711.6973  11711.6973
ENTROPY  KCAL/KGMO/K     19.7986     19.7986
CP       KCAL/KGMO/C                 23.4016
CV       KCAL/KGMO/C                 21.4014
DENSITY  KG/M3                        1.1492
Z-FACTOR                              1.0004
FLOWRATE M3/HR   T-P                1401.0021
         M3/HR   NTP                1270.6663
VISCOSITY   CP                       0.025877
TH.COND  KCAL/M/HR/C                 0.098373

 STREAM NUMBER   11     CONNECTING  (   2)  REACTOR
 NAME: Strm 11                  TO  (   0)  PRODUCT
 P.R. K-VALUES    P.R. ENTHALPY    P.R. DENSITY    FLOW RATES /HR

                     TOTAL   LIQUID    VAPOR    TOTAL    TOTAL
 COMPONENT NAME      KGMOL    KGMOL    KGMOL     KG    MOL PCT   KVALUE
 --------- ----    -------- -------- -------- -------- -------- --------
HYDROGEN             23.0963    0.  23.0963  46.5607   28.978
METHANE              2.62732    0.  2.62732  42.1501   3.2964
ETHANE               21.0397    0.  21.0397  632.623   26.398
PROPANE             0.644866    0. 0.644866  28.4347  0.80909
ETHYLENE             21.0208    0.  21.0208  589.674   26.374
PROPYLENE           0.152552    0. 0.152552  6.41910  0.19140
1,3 BUTADIENE        1.68842    0.  1.68842  91.3235   2.1184
ACETYLENE           0.360510    0. 0.360510  9.38625  0.45232
WATER                9.07186    0.  9.07186  163.433   11.382
                    -------- -------- -------- -------- --------
TOTAL                79.7023    0.  79.7023  1610.00  100.000
TOTAL   KG           1610.00    0.  1610.00

STREAM NUMBER   11
TEMPERATURE DEG C      866.08
PRESSURE    KG/CM2     1.0332
ENTHALPY    KCAL/HR   1.03833E+06
VAPOR FRACTION         1.0000

PROPERTIES              TOTAL       VAPOR
----------------      ----------  ----------
FLOWRATE   KGMOL/HR      79.7022     79.7022
MOLECULAR WT.            20.2003     20.2003
ENTHALPY KCAL/KGMOL   13027.6777  13027.6777
ENTROPY  KCAL/KGMO/K     23.1039     23.1039
CP       KCAL/KGMO/C                 16.9139
CV       KCAL/KGMO/C                 14.9260
DENSITY  KG/M3                       0.215992
Z-FACTOR                              1.0006
FLOWRATE M3/HR   T-P                7453.8203
         M3/HR   NTP                1786.1803
```

VISCOSITY	CP	0.031724
TH.COND	KCAL/M/HR/C	0.161455

Chapter 43: Polytropic Compressor Samples - c:\designii\samples\equipmnt\polycomp

Polycmp1.psd - Finding Discharge Conditions

To find discharge conditions of the compressor from efficiency curves, head curves, and impeller speed.

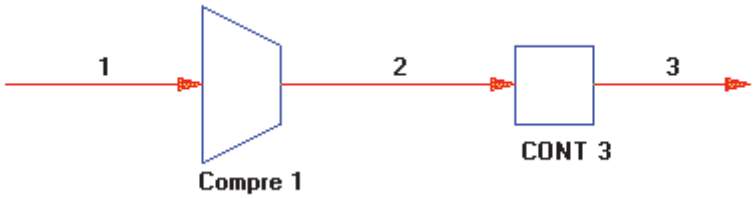



```
AB123.
*POLYTROPIC COMPRESSOR

C- EQUIPMENT MODULES

POL 1 = POLY COMPR., 1,-2,
  POLytropic EFFiciency (FT3/MIN/RPM) = 61.2,.1,
         65.4,.2, 68.9,.3,72.8,.4, 74.5,.5, 75.1,.6,
         73.8,.7, 70.9,.8, 64.5,.9
  POLytropic HEAd (FTLB/LB/RPM,FT3/MIN/RPM) = 4.56E-4, .1,
         4.38E-4, .2,4.16E-4, .3, 4.00E-4, .4, 3.81E-4, .5,
         3.63E-4, .6,3.43E-4, .7, 3.19E-4, .8, 2.88E-4, .9
  SPEed (RPM) = 7990

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 12.,35.7,10.,5.,
  TP (F,PSIA) 1 = 67.,198.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
```

```
END
```

Polycmp2.psd - Vary Speed to Achieve Discharge Pressure

To achieve a discharge pressure of 500 PSIA for the polytropic compressor in the "polycmp1.psd" example via a controller. The POLytropic module calculates the discharge pressure as in Example 1. The CONTroller reads the pressure of stream 2 and compares it to 500 (the setpoint), then changes the SPEed of the impeller. The module is recalculated with the new SPEed, and this process continues until the pressure of stream 2 is within the TOLerance.

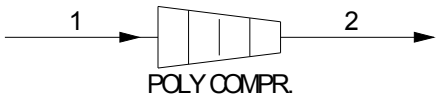



```
CONTroller 2 = SPED, 2, -3
       VARy SPEed (MAX=10000,MIN=5000, STEp = 500) OF POL 1
       UNTil PREssure FROm STReam 2 = 500(BEG=1,MAX=10,TOL=1)
```

Polycmp3.psd - Multiple Curves To Interpolate For One Unknown Speed

This simulation is to show the methodology to enter mutiple curves for head, efficiency, and flowrate in terms of speed. Interpolation at the actual speed point will be performed when a specific speed point is entered between multiple speed points.

POLYTROPIC COMPRESSOR (multiple curves)



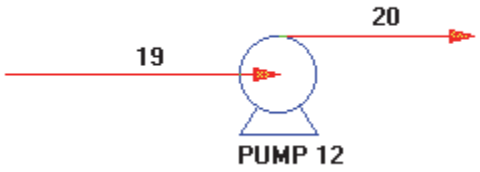

```
AB123.
*POLYTROPIC COMPRESSOR-MULTIPLE CURVES

C- EQUIPMENT MODULES
POL 1 = POLY COMPR., 1,-2,
  POL EFF(FT3/MIN/RPM)7990=
         61.7,.1,65.33,.2,68.98,.3,72.18,.4,74.47,.5,
         75.17,.55,75.34,.6,75.1,.65,74.39,.7,71.08,.8,64.96,.9
  POL HEA(FTLB/LB/RPM2,FT3/MIN/RPM)7990=4.52E-4,.1,4.469E-4,.2,
         4.402E-4,.3,4.306E-4,.4,4.169E-4,.5,3.977E-4,.6,3.718E-4,.7,
         3.379E-4,.8,2.947E-4,.9
  POL EFF(FT3/MIN/RPM)8200=61.7,.097,65.33,.195,68.98,.292,
         72.18,.39,74.47,.487,75.17,.536,75.34,.585,75.1,.633,74.39,.682,
         71.08,.780,64.96,.877
  POL HEA(FTLB/LB/RPM2,FT3/MIN/RPM)8200=4.291E-4,.097,
         4.243E-4,.195,4.179E-4,.292,4.088E-4,.39,3.958E-4,.487,
         3.776E-4,.585,3.530E-4,.682,3.208E-4,.780,2.798E-4,.877,
  POL EFF(FT3/MIN/RPM)8500=61.7,.094,65.33,.188,68.98,.282,
         72.18,.376,74.47,.470,75.17,.517,75.34,.564,75.1,.611,
         74.39,.658,71.08,.752,64.96,.846
  POL HEA(FTLB/LB/RPM2,FT3/MIN/RPM)8500=3.994E-4,.094,
         3.949E-4,.188,3.889E-4,.282,3.805E-4,.376,3.684E-4,.470,
         3.514E-4,.564,3.285E-4,.658,2.986E-4,.752,2.604E-4,.846
  POL EFF(FT3/MIN/RPM)8700=61.7,.094,65.33,.188,68.98,.282,
         72.18,.376,74.47,.470,75.17,.517,75.34,.564,75.1,.611,
         74.39,.658,71.08,.752,64.96,.846
  POL HEA(FTLB/LB/RPM2,FT3/MIN/RPM)8700=3.994E-4,.094,
         3.949E-4,.188,3.889E-4,.282,3.805E-4,.376,3.684E-4,.470,
         3.514E-4,.564,3.285E-4,.658,2.986E-4,.752,2.604E-4,.846
  SPE(RPM)=8400

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,3,4,
  APISOAVEK
  APISOAVEH
  STDD
```

```
  FLO(LBMOL/HR)1 = 12.,35.7,10.,5.,
  TP (F,PSIA) 1 = 67.,198.
END
```

Chapter 44: Pump Samples - c:\designii\samples\equipmnt\pump

Pump1.psd - Required Horsepower Calculated

A pump is used to increase the pressure of a liquid stream to 800 PSIA. The horsepower required and kilowatt usages are to be calculated.



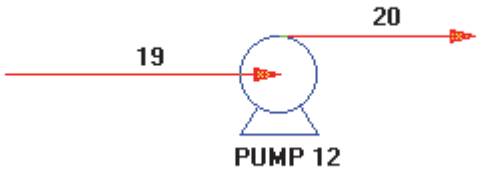

```
PUMp 12 = L.O., 19, -20,
     PREssure OUT = 800
```

Method of Calculation

All properties of stream 19 are known. The default volumetric efficiency of 0.9 will be used as the efficiency in calculating the work required to increase the pressure to 800 PSIA. The temperature out is calculated along with the outlet enthalpy. The calculated work required should be divided by the mechanical efficiency to get the true work required.

Pump2.psd - Outlet Pressure Calculated

A 25-HP pump is to be used to increase the pressure of a liquid stream as much as possible. The driver for the pump uses 950-BTU/SCF fuel gas. The mechanical efficiency is 90% and the volumetric efficiency is 95%. The pressure out (approximately 800 PSIA) and fuel requirements for the driver are to be calculated.




```
PUMp 12 = LOPM, 19, -20,
       WORk AVAilable = 22.5
       PREssure OUT = 1000, EFFiciency = .95
```

Method of Calculation

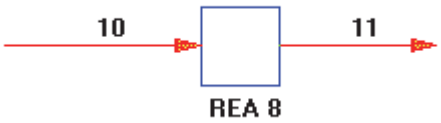
All properties of stream 19 are known. The mechanical efficiency is taken into account by adjusting the 25-HP (WORk AVAilable = 22.5). Since the pressure out must be larger than can be obtained with the WORk AVAilable, PREssure OUT = 1000 was specified. The work required to achieve the PREssure OUT is calculated and compared with the WORk AVAilable. The PREssure OUT is adjusted until convergence is obtained. The fuel gas required is calculated, assuming 980 BTU/SCF and 8000 BTU/HR/HP. Since the example fuel is 950 BTU/SCF, the fuel consumption should be increased by 980/950.

Chapter 45: Reactor Samples - c:\designii\samples\equipmnt\reactor

Reactr1.psd - Conversion Of Ethylene To N-Octane Coding

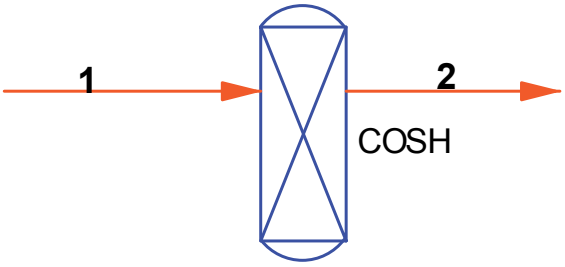
In this reaction, 98% of the ethylene (C2=) in the feed is consumed. For every mole of ethylene, one-half mole of i-butane is consumed and one-half mole of n-octane is created. All other components pass through the reactor as inerts. The feed temperature and pressure (since no pressure drop was specified) ,along with the calculated product composition, are used in calculating the product enthalpy.

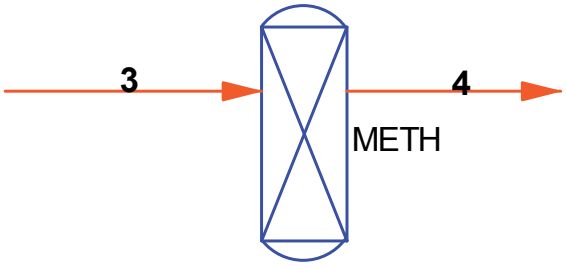
```
REActor 8 = CBRX, 10, -11,
        ISOthermal reactor,
        CONversion = .98,
        LIMiting reactant = 22,
        STOichiometric COEfficient = 0,0,0,-2,0,-1,1,
GEN, COM = 46,3,4,22,6,5,12
END
```



Reactr2.psd - COshift And Methanation Reactors

The phase of the feed(s) to each reactor is checked. If the feeds are vapor, the calculation continues. For ADIabatic reactor, the enthalpy of the feed is determined and the heat of reaction is calculated after meeting the desired approach to equilibrium. The product temperature is calculated from the feed enthalpy, product composition and enthalpy.

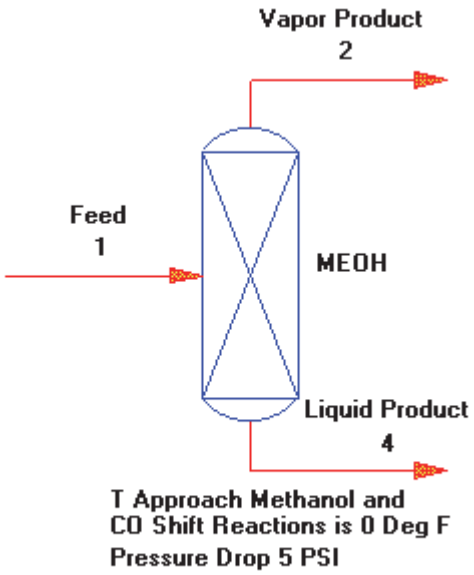





```
*EQUILIBRIUM REACTORS EXAMPLE
REActor 1 = COSH, 1, -2,
        COSHIFt reaction, NATional,
        TEMperature APProach COSHIFt = -25,
        ADIabatic, DELta pressure = 10,
REActor 2 = METH, 3, -4,
        METHANation,
        TEMperature APProach METHANation = -30,
        NATional,
        TEMperature APProach COSHIFt = -25,
        ADIabatic, DELta pressure = 10,
GENeral,
                                    COMponents = 1, 48, 49, 2, 62, 3
        FLOw 1 = 200, 200, 200, 50, 200, 500
        FLOw 3 = 200, 200, 200, 50, 200, 500
        TP 1 = 500, 300
        TP 3 = 500, 300
        SOAVEK, SOAVEH, STDD,
END
```

Reactr3.psd - Methanol Synthesis Reactor

The feed phase to the reactor is checked to be sure the feed is all vapor. If it is, the calculation proceeds until the heat and mass balance solutions meet the desired approach to equilibrium temperature. Methane and argon are treated as chemically inert. The outlet temperature is calculated from the product composition, feed enthalpy and the heat of reaction.

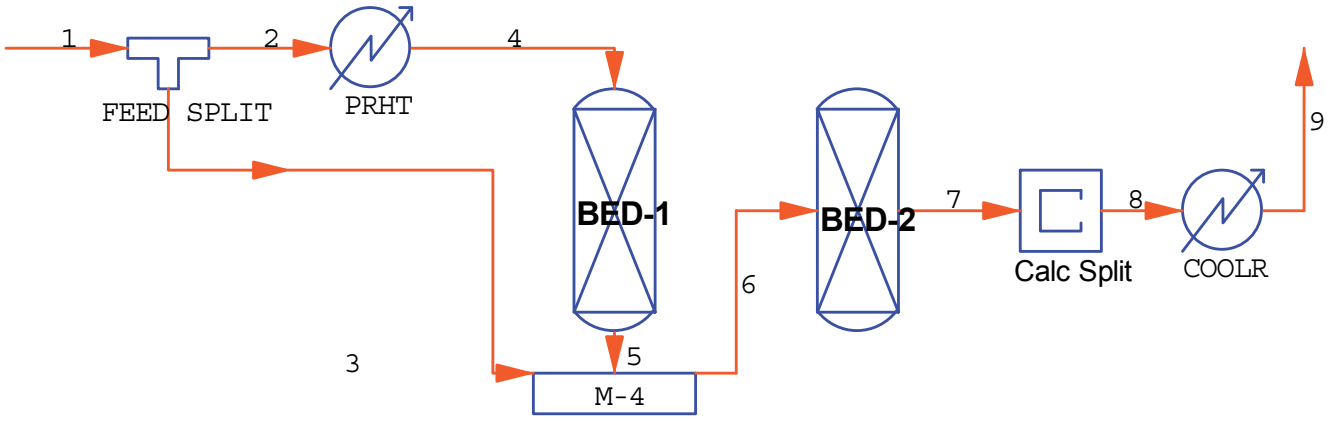


Methanol Synthesis Reactor

```
*EXAMPLE BASED ON CATALYST MANUFACTURER MATERIAL
        REActor 1 = MEOH, 1, -2, -4,
        METhanol SYNthesis reaction,
        TEMperature APProach METhanol SYNthesis = 0,
        NATional,DELta pressure = 5,
        TEMperature APProach COSHIFt = 0
GENeral
        COMponents = 48, 49, 1, 2, 46, 62, 1021, 1171
        FLOw 1 = 11, 57, 3, 22, 69.17, 2.9, 10.96, 0.02,
                 .52, 1.64
        TP 1 = 500, 725
        PENK, PENH, PEND
END
```

Reactr4.psd - Ammonia Synthesis Reactor

The program checks the feed phase to verify that it is vapor. The heat and mass balances are then calculated to satisfy the approach to equilibrium temperature. The product temperature is calculated from product composition, feed enthalpy and heat of reaction. The CONTroller checks the temperatures of streams 4 and 6 to determine if they are equal. If not, the split for DIVider 1 is adjusted, and all modules are recalculated until the temperatures are equal or the tolerance is satisfied.

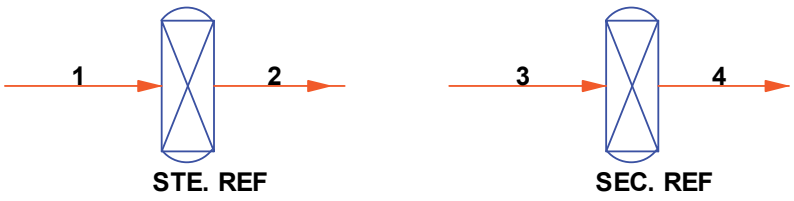



```
AB123.
*AMMONIA SYNTHESIS LOOP - CALCULATE BED SPLIT

DIV 1 = FEED SPLIT, 1,-2,-3,
  FLO(FRA) = 0.7

HEA EXC 2 = PRHT, 2,-4,
  DUT(BTU/HR) = 2147000
```

```
  U(BTU/HR/FT2/F) = 50
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

REA 3 = BED-1, 4,-5,
  ADI
  DEL(PSI) = 20
  NAT
  AMM SYN
  TEM APP AMM SYN (F) = 0

MIX 4 = M-4, 5,3,-6,

REA 5 = BED-2, 6,-7,
  ADI
  DEL(PSI) = 20
  NAT
  AMM SYN
  TEM APP AMM SYN (F) = 0

CONT 6 = Calc Split, 7,-8,
  VARY FLO RAT (MIN=0.01,MAX=0.99,STE=0.08,FRA) OF DIV 1
  UNTIL TEM FROM STREAM 4 / TEM FROM STREAM 6 =
  1 (BEG=1,LOO=10,TOL=1e-005)

HEA EXC 7 = COOLR, 8,-9,
  DUT(BTU/HR) = -2147000
  U(BTU/HR/FT2/F) = 50
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0


C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,46,63,1171,2,
  PENK
  PENH
  PEND
  DEN LIQ = PEND
  SUR TEN = STD
  BINPAR = PENG1
  FLO (FRA) 1 = 0.65,0.216,0.034,0.03,0.07,
  TOT FLO (LBMOL/HR) 1 = 1000
  TP (F,PSIA) 1 = 380,3400
END
```

Reactr5.psd - Steam Reformer Reactor

Steam reformer example for using hydrocarbons (including CO, CO2, H2, H2O, and O2). The components will react and achieve thermal and mass equilibrium. Any oxygen found in the input stream is totally consumed.

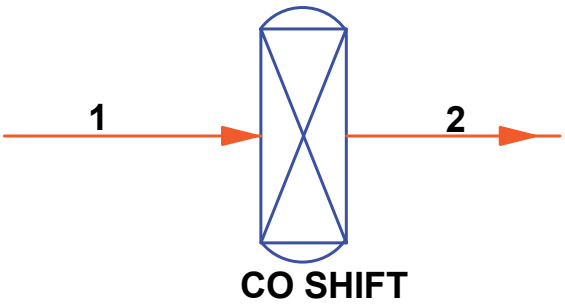



```
AB123.
*EXAMPLE FOR STEAM/ SECONDARY REFORMER

C- EQUIPMENT MODULES
```

```
REA 4 = STE. REF, 1,-2,
  ADI
  DEL(PSI) = 9.
  NAT
  STE REF
  TEM APP COSHIF(F) = 15.
  TEM APP METHAN(F) = 8.

REA 5 = SEC. REF, 3,-4,
  TEM OUT(F) = 1157.6
  DEL(PSI) = 6.
  NAT
  SEC REF
  TEM APP COSHIF(F) = 40.
  TEM APP METHAN(F) = -19.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 48,49,1,2,62,46,1021,1171,63,47,3,22,100,
  NAM 100 = 243_ABP
  AMW 100 = 109.54
  AMB (F) 100 = 243.
  API (API) 100 = 57.56
  SOAVEK
  SOAVEH
  SOAVED
  SUR TEN = STD
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 1.14,0.31,25.97,3.3,1110.2,10.96,0.52,1.64,5.32,8.6,6.23,
  2.83,82.4,
  TP (F,PSIA) 1 = 850.,550.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  FLO(LBMOL/HR)3 = 67.57,14.28,69.17,2.9,4.19,10.96,0.52,1.64,5.32,8.6,7.2,2.6,
  67.1,
  TP (F,PSIA) 3 = 500.,725.
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
END
```

Reactr6.psd – CO Shift Using Product Rate Specification

Reactor flowsheet showing CO Shift using product rate specification.



Chapter 46: Refine Column Samples - c:\designii\samples\equipmnt\refine

Refi1.psd - Total Condenser (TOT)

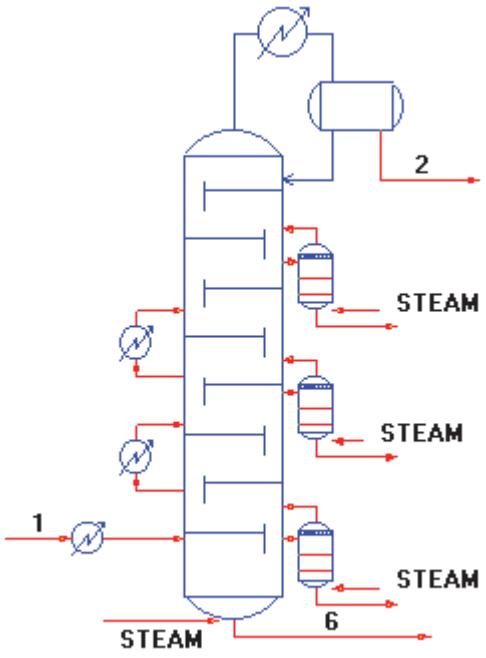
Example with an externally refluxed TOTal condenser column. Reflux returning to the top tray includes any water dissolved in the hydrocarbons. Any remaining water in the condenser is automatically decanted.

Feed Flowrate = 63750 BPD
Temperature = 650 F Pressure = 29.7 psia

Real Components	Volume percents
water	-na-
ethane	0
propane	0.008
i-butane	0.012
n-butane	0.058
i-pentane	0.099
n-pentane	0.130

Crude Characterization:

Volume percent	TBP (F)	API Gravity
0	-40	65.9
5	245	55.2
10	294	50.8
20	380	44.3
30	463	38.5
40	565	33.1
50	637	29.1
60	749	26.5
70	848	24.1
80	942	19.2
90	1127	9.6
95	1253	-0.7
100	1343	-1.3



Guesses: Temperature of Top Tray = 330 F
Temperature of Bottom Product = 700 F
Bottom Product Flow Rate = 26875 BPD
Molar Flowrate of Liquid from Top Tray = 130 lbmoles/hr

Specification: Bottom Product Flowrate = 26875 BPD

Coding

```
AB123.
* TOTAL CONDENSER EXAMPLE
REFI 33 = FRAC, 1, -2, -3, -4, -5, -6
       TOTal CONdenser
       STAges = 17, 3, 3, 3
       TEMperature of TOP = 330
       TEMperature of BOTtom = 700
       PREssure of TOP = 24.7, 25.2, 26.2, 27
       DELta pressure = 4, 3*0
       PREssure of CONdenser = 19.7
       LOCation of FEEd = 16
       PROduct rates (BBL/DAY) = 0, 26875
       FIX LIQuid (BBL/DAY) 1 = 3187.5, 15, H1
       FIX GUEss = 130
C-
C- SIDE STRIPPER DESCRIPTION
C-
       LOCation of VAPor returns = 4, 9, 14
       LOCation of DRAws = 4, 9, 14
       PROduct from STRippers (BBL/DAY) = 9100, 7875, 9200
C-
C- PUMPAROUNDS
C-
       LOCation of PUMparounds = 9, 8, 14, 13
```

```
       PROduct of PUMparounds = 2200, 1900
       HEAt from PUMparounds = 35.E6, 40.E6 C-
C- HEAT SOURCES
C-
       LOCation of STEam = 18, 24, 27
       TEMperature of STEam = 3*370
       PREssure of STEam = 3*164.7
       STEam rates (LB/HR) = 3000, 1000, 1000
       REBoiler = 0, 1.9E6
C-
C- PRINT CONTROLS
C-
       CONdensing CURve
       REPorts = 1
       PROduct COOling CURves = 1, 0, 0, 0, 0
       PROduct PUMparound cooling CURves = 0, 0
       MAX MAT = 3
C-
C- CRUDE FEED DESCRIPTION
C-
CRUde
       FEEd RATe (BBL/DAY) 1 = 63750
       FEEd REAl 1 = 0, 0.008, 0.012, 0.058, 0.099, 0.130
       FEEd TBP 1 = -40, 245, 294, 380, 463, 565, 637, 749,
       848, 942, 1127,1253, 1343
       FEEd VOLume PERcents 1 = 0, 5, 10, 20, 30, 40, 50, 60,
       70, 80, 90,95, 100
       FEEd GRAvity (API) 1 = 65.9, 55.2, 50.8, 44.3, 38.5,
       33.1,29.1, 26.5, 24.1, 19.2, 9.6, -0.7, -1.3
       TP 1 = 650, 29.7
       TEMperature INCrements = 30, 60, 120
       TEMperature OF BREaks = 600, 900
C- GENERAL SECTION
C-
GENeral
       COMponents = 62, 3, 4, 5, 6, 7, 8
       STDK , API , STDD
       HEAt CURve 1
       API ENThalpy
END
```

Refi2.psd - Partial Condenser (PAR)

Refine column with an externally refluxed PARtial condenser column which decants water (if it is present in sufficient quantity) as a separate liquid phase. Solubility of water in liquid hydrocarbons is taken into account for both the liquid distillate and the reflux returning to the column. The vapor distillate will be saturated.

Feed Flowrate = 60000 BPD
Temperature = 600 F Pressure = 32.5 psia
Crude Characterization :

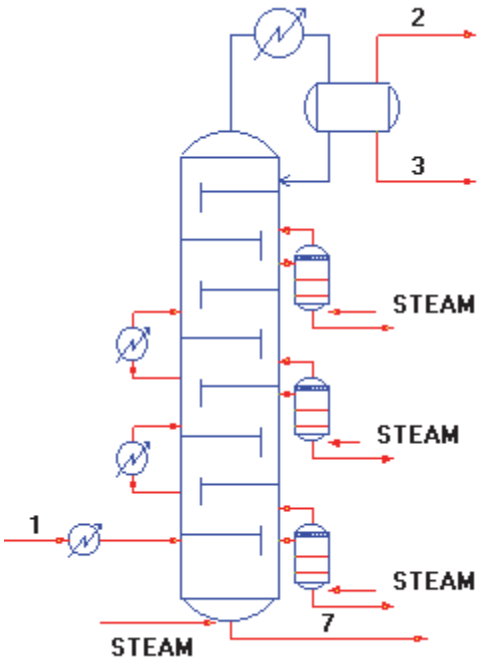
Volume percent	TBP(F)
0	0
5	125
10	200
20	301
30	405
40	499
50	615
60	718
70	819
80	954
90	1157
100	1450

Bulk API Gravity = 33.4

Guesses: Temperature of Top Tray = 250 F
Temperature of Bottom Product = 64
Bottom Product = 785 lbmoles/hr 0 F
Product Flow Rates : Vapor Distillate = 25 lbmoles/hr
Liquid Distillate = 590 lbmoles/hr

Specifications: Temperature of Condenser = 120 F
Bottoms Product Flowrate = 5400 BPD

```
AB123.
EXAMPLE REFINE PROBLEM WITH PARTIAL CONDENSER
C-
REFI 27 = FRAC, 1, -2, -3, -4, -5, -6, -7
        PARtial CONdenser ,
        STAges = 15, 2, 2, 2
        LOCation of FEEd = 13
        PROduct NAMes = VENTGAS,LT NAPHTHA,HVY NAPHTHA,KERO,RESID
C-
C- MAIN COLUMN DESCRIPTION
C-
        PREssure of CONdenser = 19.7
        TEMperature of TOP GUEss = 250,
        TEMperature of BOTtom GUEss = 640,
        PREssure of TOP = 28
        DELta pressure = 5
        PROduct rates (BBL/DAY) = 0, 5400, 0,
        PROduct rate GUEsses = 25, 590, 785,
        FIX LIQuid (BBL/DAY) 1 = 1500, 12, H1,
        FIX GUEss = 75
        TEMperature of CONdenser = 120
C-
C- SIDE STRIPPERS
C-
        LOCation of DRAws = 3, 7, 11,
        LOCation of VAPor RETurns = 3, 7, 11
        PROduct from STRippers (BBL/DAY) = 5040, 9000, 11880
        PROduct from STRippers GUEsses = 440, 625, 615
C-
C- PUMPAROUNDS
C-
        LOCation of PUMparounds = 7, 6, 11, 10
        PROduct from PUMparounds = 1200, 2100
        HEAt from PUMparounds = 15E6, 40E6
C-
C- HEAT SOURCE DESCRIPTION
C-
        LOCation of STEam = 15, 17, 19, 21
        TEMperature of STEam = 4*400
        PREssure of STEam = 4*45
        STEam RATes (LB/HR) = 8000, 600, 800, 1500
C-
C- PRINT OPTIONS
C-
        PROduct CURves = 1, 5*0
        PROduct from PUMparound CURves = 0, 0
        CONdensing CURve
C- GENERAL SECTION
GENeral
               COMponents = 62, 2, 3, 4, 5
               ESSO TAB , GRAYSON JOHNSON
               HEAt CURve 1
C-
C- CRUDE FEED DESCRIPTION
CRUde
        FEEd TBP 1 = 0, 125, 200, 301, 405, 499, 615, 718,
                819, 954, 1157, 1450
        FEEd GRAvity 1 = 33.4
        FEEd VOLume 1 = 0, 5, 10, 20, 30, 40, 50, 60, 70,
                80, 90, 100
        FEEd RATe (BBL/DAY) 1 = 60000
        TP 1 = 600, 32.5
END
```



Refi3.psd - Pumparound Internal Reflux without Condenser (PUM REF)

Example showing internally refluxed column with no overhead condenser to provide reflux. Instead, a pumparound on tray 1 generates internal reflux. With this column type, there is no automatic decant of liquid water.

Feed Flowrate = 5000 BPD
Temperature = 650 F Pressure = 95 psia
Crude Characterization :

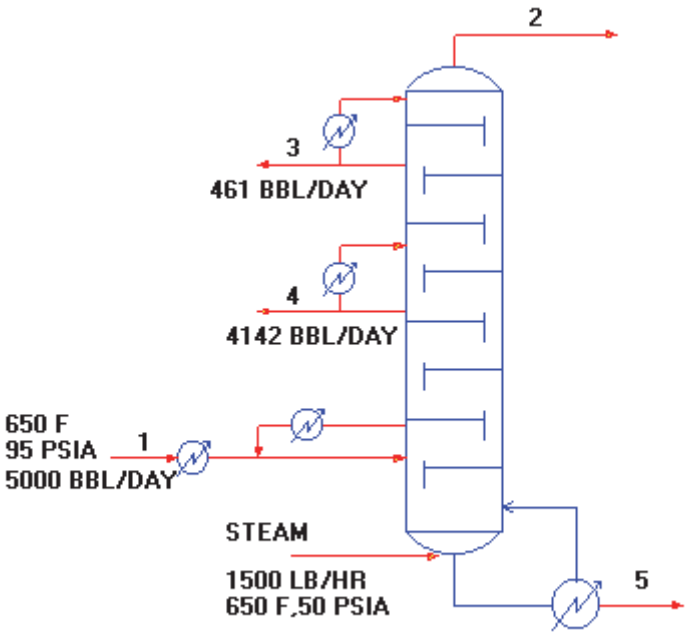
Volume percent	TBP (F)
2	471

5	530
10	584
20	651
30	685
40	714
50	741
60	771
70	802
80	855
90	943
95	977
98	1059

Bulk API Gravity = 1.7

Guesses: Top Product Temperature = 250 F
Bottom Product Temperature = 600 F
Vapor Rate Tray Profile = 3.3,160,230,250,300,350,94,83 lbmoles/hr
Liquid Molar Flow Rate from Tray 6 = 44 lbmoles/hr

Specification: Top Tray Temperature = 160 F

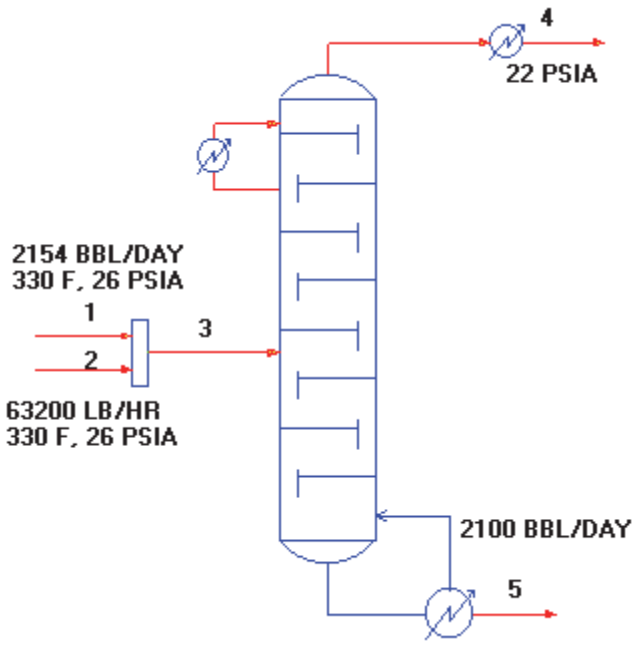



```
AB123.
* EXAMPLE REFINE WITH PUMPED REFLUX
REFI 100 = VAC, 1, -2, -3, -4, -5
        PUMped REFlux STAges = 8
        TEMperature of TOP = 250
        TEMperature of BOTtom = 600
        PREssure of TOP (MMHG) = 75
        PREssure PROfile (MMHG)=75,76, 80, 85, 90, 90, 110,100
        VAPor PROfile = 3.3, 160, 230, 250, 300, 350, 94, 83
        LOCation of FEEd = 7
        LOCation of SIDedraw = 2, 4
        PROduct rates (BBL/DAY) = 0, 461, 4142, 300
        PROduct rate GUEsses = 3, 28, 221, 15
C-
C- SPECIFY THE TOP TRAY TEMPERATURE
C-
        FIX TEMperature 1 = 160, 1, 1
C-
C- FIX THE LIQUID FLOW FROM TRAY 6 (OVERFLASH)
C- ADJUST FEED NUMBER 1 FURNACE DUTY TO ACHIEVE THIS C-
        FIX LIQuid (BBL/DAY) 2 = 933, 6, H1 C-
C- ESTIMATE OF LIQUID MOLAR FLOW RATE FOR FIX LIQ COMMAND C-
        FIX GUEss = 44
C-
C- PUMPAROUND DESCRIPTION
C-
```

```
        LOCation of PUMparounds = 2, 1, 4, 3, 6, -1
        PROduct of PUMparound (BBL/DAY) = 3000, 19300, 933
        PROduct of PUMparound GUEsses = 228, 1038, 44
        HEAt from PUMparounds = 1E6, 15.85E6, 0.5E5
C-
C- DESCRIBE THE STEAM SOURCES
C-
        LOCation of STEam = 9
        TEMperature of STEam = 650
        PREssure of STEam = 50 STEam (LB/HR) = 1500
C-
C- PRINT CONTROLS
C-
        PROduct cooling CURves = 0, 0, 0, 0
        PROduct of PUMparound cooling CURves = 2*0
        PROduct NAMes = OVHD, LVGO, HVGO, RESID
        MAXimum MATrices = 3
C-
C- GENERAL SECTION
C-
GENeral
        COMponents = 62
        MODESSO, APIMOD
        API ENThalpy
        HEAt CURve 1
C-
C- CRUDE FEED DESCRIPTION
C-
CRUde
        TP 1 = 650, 95
        FEEd ASTM 1 = 471, 530, 584, 651, 685, 714, 741,
                771, 802, 855, 943, 977,1059
        FEEd VOLume percents 1 = 2, 5, 10, 20, 30 , 40 ,
                50, 60, 70, 80, 90, 95,98
        FEEd GRAvity 1 = 1.7 FEEd (BBL/DAY) 1 = 5000
END
```

Refi4.psd - Pumparound Internal Reflux with Total Condenser (PUM TOT)

Example of refine with internally refluxed column having a top tray pumparound to generate reflux. There is no liquid returned from the external condenser to the main column. Water, if present, is decanted automatically from the external condenser.



Feed Stream 1 Component Flowrates :

Water	0
Toluene	63200 lb/hr

Temperature = 330 F Pressure = 26 psia

Feed Stream 2 Flowrate = 63200 LB/HR

Temperature = 330 F Pressure = 26 psia

Crude Characterization :

Volume percent	TBP(F)
0	400
5	475
10	609
20	785
30	885
40	956
50	1006
60	1044
70	1083
80	1122
90	1161
95	1181
100	1200

Bulk API Gravity = 8.1

Guesses: Liquid Distillate Flowrate = 691 lbmoles/hr
Bottom Product Flowrate = 69 lbmoles/hr
Temperature of Condenser = 259 F
Temperature Profile = 265,277,297,316,400,7*0,766 F
Vapor Profile = 689,850,824,787,87,120,202,240,4*0,140 lbmoles/hr

Specification: Top Pumparound Duty = 2.7 MMBtu/hr

```
AB123.
* BITUMEN RECOVERY ; REFINE PUM TOT EXAMPLE
MIXer 1 = MIXR, 1, 2, -3
REFI 2 = COAL, 3, -4, -5
        PUMped TOTal condenser
        STAges = 13
        PREssure of TOP = 25 ,
        DELta pressure = 1.5
        PREssure of CONdenser = 22
        LOCation of FEEd = 5
        PROduct rates GUEss = 691, 69
        TEMperature of CONdenser GUEss = 259
        LOCation of PUMparound = 2, 1
        PROduct of PUMparound = 400
        HEAt from PUMparounds = 2.7E6
        REBoiler (BTU/HR) = 9.4E6
        TEMperature PROfile =265,277,297,316,400,7*0,766
        VAPor PROfile =689,850,824,787,87,120,202,240,4*0,140
        REPort = 1, 1, 1, 0, 0, 1
        PROduct NAMes = SOLVENT, BITUMEN
        COLumn sizing OFF
C-
C- CRUDE FEED DESCRIPTION
C-
CRUde
        FEEd rate (BBL/DAY) 1 = 2154
        FEEd TBP 1 = 400, 475, 609, 785, 885, 956, 1006, 1044,
                1083, 1122, 1161, 1181, 1200
        FEEd GRAvity 1 = 9.1
        FEEd NAMe 1 = BITUMEN FEED
        FEEd NAMe 2 = TOLUENE FEED
        FEEd rate (LB/HR) 2 = 0, 63200
        TP 1 = 330 , 26 , TP 2 = 330, 26
C-
C- GENERAL SECTION
C-
GENeral
        COMponents = 62, 41 , STDH , BRAUN
END
```

Refi5.psd - Pumparound Internal Reflux with Partial Condenser (PUM PAR)

This refine column has a top tray pumparound to generate reflux, and an external partial condenser. There is no liquid returned to the column from the partial condenser. Water (if present) is decanted automatically from the condenser.

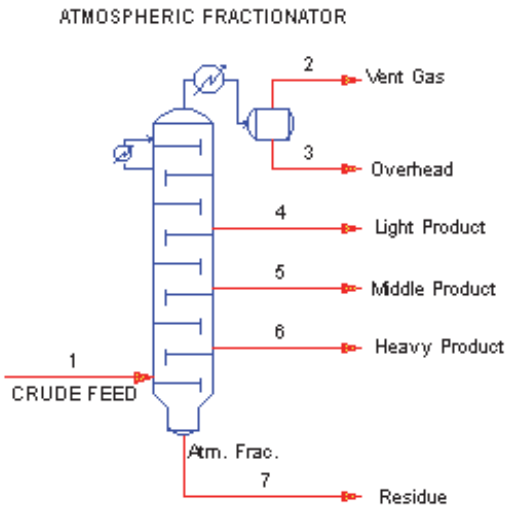
Feed Flowrate = 63750 BPD

Temperature = 650 F Pressure = 29.7 psia

Real Components	Volume percents
water	-NA-
ethane	0
propane	0.008
i-butane	0.012
n-butane	0.058
i-pentane	0.099
n-pentane	0.130

Crude Characterization :

Volume percent	TBP(F)	API Gravity
0	-40	65.9
5	245	55.2
10	294	50.8
20	380	44.3
30	463	38.5
40	565	33.1
50	637	29.1
60	749	26.5
70	848	24.1
80	942	19.2
90	1127	9.6
95	1253	-0.7
100	1343	-1.3



Guesses: Temperature of Top Tray = 330 F
Temperature of Bottom Product = 700 F

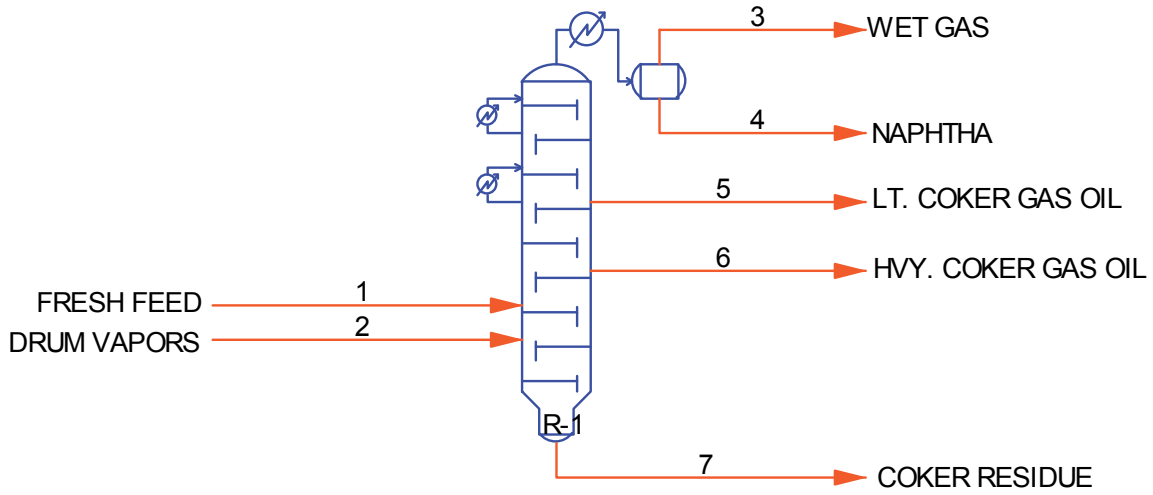
Specifications: Temperature of Condenser = 200 F Bottom Product Rate = 26100 BPD

```
AB123.
* PUM PAR EXAMPLE
REFI 1 = FCC, 17, -25, -16, -7, -12, -3, -99
       PUMped with PARtial condenser
       STAges = 18, 3, 3, 3
       TEMperature of TOP = 330 ,
       TEMperature of BOTtom = 700
       PREssure of TOP = 24.7, 25.2, 26.2, 27 ,
       DELta pressure = 4, 3 * 0
       PREssure of CONdenser = 19.7 ,
       LOCation of FEEd = 16
       PROduct rates (BBL/DAY) = 0, 0, 26100
       TEMperature of CONdenser = 200
C- SPECIFY THE OVERFLASH
C-
       FIX LIQuid (BBL/DAY) = 3187.5, 15, H1
       FIX GUEss = 130
C-
C- SIDE STRIPPER DESCRIPTION
C-
       LOCation of DRAws to strippers = 4, 10, 14
       LOCation of VAPor returns = 4, 10, 14
       PROducts from STRippers (BBL/DAY) = 9100, 7875, 9200
C-
C- PUMPAROUND DESCRIPTION
C-
       LOCation of PUMparound = 4, 1, 10, 9, 14, 13
       PROduct of PUMparounds = 1000, 2200, 1900
       HEAt from PUMparounds = 10.E6, 35.E6, 40.E6
C-
C- HEAT SOURCES
C-
       LOCation of STEam = 18, 24, 27
       TEMperature of STEam = 3 * 370
       PREssure of STEam = 3 * 164.7
       STEam rates (LB/HR) = 3000, 1000, 1000
       REBoiler = 0, 1.9E6
C-
C- PRINT CONTROLS
C-
       CONdensing CURve
       PROduct of PUMparound cooling
       CURve = 3 * 0
       MAXimum MATrices = 3
```

```
C-
C- GENERAL SECTION
C-
GENeral
        STDK , APIMOD , STDD
        HEAting CURve 17 = 350 , 0, 7
        COMponents = 62, 3, 4, 5, 6, 7, 8
        API ENThalpy
C-
C- CRUDE FEED DESCRIPTION
C-
CRUde
        FEEd rate (BBL/DAY) 17 = 63750
        FEEd REAl 17 = 0, 0.008, 0.012, 0.058, 0.099, 0.130 FEEd TBP 17 = -40, 245, 294, 380, 463, 565,
        637, 749,
                848, 942, 1127, 1253, 1343
        FEEd VOLume percents 17 = 0, 5, 10, 20, 30, 40, 50,
                60, 70, 80, 90, 95, 100
        FEEd GRAvity (API) 17 = 65.9, 55.2, 50.8, 44.3, 38.5,
                33.1,29.1, 26.5, 24.1, 19.2, 9.6, -0.7, -1.3
        TP 17 = 650, 29.7
        TEMperature INCrements = 30, 60, 120
        TEMperature of BREaks = 600, 900
END
```

Refi6.psd - Coker Fractionator Column

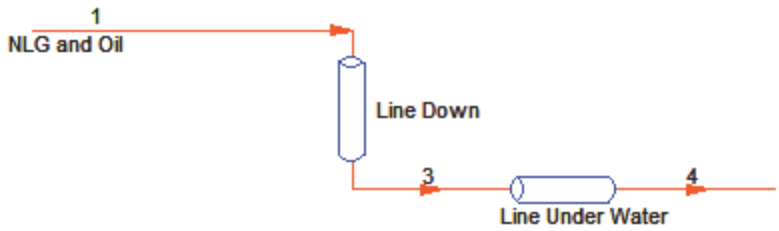
An example of a coker fractionator using REFine module. The column has pumparounds to generate reflux, and an external partial condenser to produce products such as Naptha, Light Coker Gas Oil, Heavy Coker Gas Oil, and Residue.



Chapter 47: Reverse Calculation Samples - c:\designii\samples\ reverse_calc

reverse1.psd - Phase Map and Line with Surroundings Heat Transfer

Phase Map and Line with Surroundings Heat Transfer



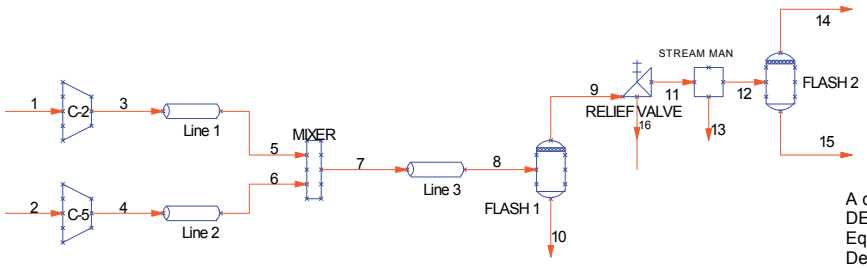
c:\designii\samples\reverse_calc\reverse1.psd

```
AB123.
*Line Pressure Drop with Heat Transfer to Surroundings

C- EQUIPMENT MODULES

LINE 2 = Line Down, 1,-3,
  DIA (INCH) = 16
  PIP WAL = STD
  DOWN
C- ENTRIES IN LEN COMMAND TOTAL TO 45.72 M
  LEN (M) = 45.72,
  SEG LEN (M) = 4.572
  MET HOL = DUN
  TEM SUR(F) = 70,
  PIP MAT = STA
  SUR AIR
  VEL SUR(FT/SEC) = 4,

LINE 3 = Line Under Water, 3,-4,
  DIA (INCH) = 16
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 16000 M
  LEN (M) = 10*1600,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (M) = 0,3.048,3.048,1.524,0.609,6.097,9.144,3.049,3.047,3.047
  SEG LEN (M) = 1600
  MET HOL = MUK
  TEM SUR(F) = 10*60,
  PIP MAT = CAR
  SUR WAT CON(KCAL/HR/M/C) = 10*0.1709,
  VEL SUR(M/MIN) = 10*0.051,


C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,46,2,3,4,5,6,7,8,10,11,12,100
  NAM POS 1 = 'CARBON DIOXIDE'
  NAM POS 2 = 'NITROGEN'
  NAM POS 3 = 'METHANE'
  NAM POS 4 = 'ETHANE'
  NAM POS 5 = 'PROPANE'
  NAM POS 6 = 'I-BUTANE'
  NAM POS 7 = 'N-BUTANE'
  NAM POS 8 = 'I-PENTANE'
  NAM POS 9 = 'N-PENTANE'
  NAM POS 10 = 'N-HEXANE'
  NAM POS 11 = 'N-HEPTANE'
  NAM POS 12 = 'N-OCTANE'
  NAM POS 13 = 'C9+'
  AMW 100 = 120.797
  AMB (C) 100 = 137.44
```

```
  API (SPG) 100 = 0.7138
C-  Thermo Comments
  PENK
  PENH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  REV CAL
  REV TOL = 1e-007
  REV MAX = 1000
  NAM STR 1 = NLG and Oil
  NAM STR 3 = Strm 3
  FLO(LBMOL/HR)4 = 5.00008,3.00049,2000.01,999.995,99.9995,70.0012,50.0008,120,
           160.001,110,200.001,207.001,500.008,
  TP (F,PSIA) 4 = 60.915,1595.945
  NAM STR 4 = Strm 4
  PLOT VELOCITY AND DELTA PRES FRIC VS LENGTH = 2,3
  PLOT DEL P FRIC AND PRESSURE VS LENGTH = 2,3
  PLOT DEL P ELEV AND ELEV VS LENGTH = 2,3
  PLOT TEMPERATURE AND DEL P ELEV AGAINST LENGTH = 2,3
  PLOT HOLDUP VS LENGTH=2,3
END
```

Flrehdr_reverse.psd - Flare Header Model with Depressurizer

FLARE HEADER MODEL WITH DEPRESURRIZER - Reverse Calculation



A description of this simulation is under HELP...
DESIGN II Reference Guide...Equipment Modules...
Equipment Modules by Name...Depressuring...
Depressuring Analysis Examples: Depressuring:
Flrehdr.psd - Flare Header Simulation

c:\designii\samples\reverse_calc\Flrehdr_reverse.psd

```
MYXXX.CGR.EXAMPLE
*Gas Flare Header Sample Problem

C- EQUIPMENT MODULES

STR 1 = STREAM MAN, 11,-13,-12,
  MUL
  SAM FAC, FAC = 1

COMPRE 2 = C-2, 1,-3,
  PRE OUT(PSIA) = 650
  EFF = 0.76
  CAL POL
  ELE DRI
  STA = 1
  PRE INL(PSIA) = 350

DEP 3 = RELIEF VALVE, 9,-11,-16,
  CRO(IN2) = 3.2648
  COE = 0.975
  DIA(FT) = 15
  VER
  LEN(FT) = 120
  LIQ HEI(MM) = 0.01
  RED(PSIA) = 300
  BAC(PSIA) = 25
```

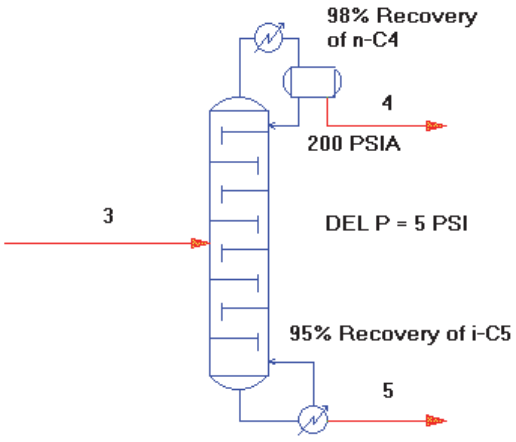
```
  SET(PSIA) = 650
  RES(PSIA) = 0
  TIME STEP(MIN) = 2
  MIN TIME(MIN) = 15
  MAX TIME(MIN) = 60
  ENV = 1
  PRI = 2
  WET = 34500
  RUP = 0
  BAL = 0
  BAC FAC = 0
  VAP(FRA) = 0.1

FLA 4 = FLASH 1, 8,-9,-10,
  TEM FEE
  PRE FEE

COMPRE 5 = C-5, 2,-4,
  PRE OUT(PSIA) = 650
  EFF = 0.76
  CAL POL
  ELE DRI
  STA = 1
  PRE INL(PSIA) = 350

LINE 6 = Line 1, 3,-5,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 100 FT
  LEN (FT) = 100,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0
  ROU (IN) = 2
  MET ACC
  ADI
      PREssure OUT (PSIA) =600

FLA 8 = FLASH 2, 12,-14,-15,
  TEM FEE
  PRE FEE

LINE 9 = Line 2, 4,-6,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 100 FT
  LEN (FT) = 100,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0
  SEG LEN (FT) = 10
  ROU (IN) = 2
  ADI
      PREssure OUT (PSIA) =600

MIX 10 = MIXER, 5,6,-7,
  FLO(LBMOL/HR) = 100,

LINE 11 = Line 3, 7,-8,
  DIA (INCH) = 4
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 1000 FT
  LEN (FT) = 2*500,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (FT) = 0,0
  SEG LEN (FT) = 100
  ROU (IN) = 2
  ADI
      PREssure OUT (PSIA) =580

C- FLOWSHEET LINE SIZING
SIZ LIN
  USE streams = 5, 6, 8

  MAXimum SONic fraction =0.75
```

```
C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,6,62
  NAM POS 1 = 'NITROGEN'
  NAM POS 2 = 'CO2'
  NAM POS 3 = 'METHANE'
  NAM POS 4 = 'ETHANE'
  NAM POS 5 = 'PROPANE'
  NAM POS 6 = 'N-BUTANE'
  NAM POS 7 = 'WATER'
C-  Thermo Comments
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  AME UNI OUT
  MAX = 25
  CON TOL = 0.001
  REV CAL
  REV TOL = 1e-007
  REV MAX = 1000
  FLO (FRA) 1 = 0.009,0.006,0.85,0.12,0.01,0.005,
  TOT FLO (LBMOL/HR) 1 = 100
  TP (F,PSIA) 1 = -35,350
  NAM STR 1 = Strm 1
  FLO (FRA) 2 = 0.009,0.006,0.79,0.18,0.01,0.005,5e-005,
  TOT FLO (LBMOL/HR) 2 = 50
  TP (F,PSIA) 2 = -35,350
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  FLO(LBMOL/HR)14 = 1.34998,0.899985,124.498,20.9996,1.49998,0.749988,0.0025,
  TP (F,PSIA) 14 = 52.16085,646.6143
  NAM STR 14 = Strm 14
  FLO(LBMOL/HR)15 = 0,
  TP (F,PSIA) 15 = 100,100
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
END
```

Chapter 48: Shortcut Fractionator Column Samples - c:\designii\samples\equipmnt\shortcut

Shortct.psd - Debutanizer

Columns can be shortcut with the SHOrtcut fractionator module if they have both a condenser and reboiler and only a top and bottom product. Other columns can be shortcut with the COMponent SPLitter Module.



Components	ID No.	Flowrates ,LBMOL/HR
C3	4	2000
i-C4	5	400
n-C4	6	600
i-C5	7	150
n-C5	8	200
n-C6	10	50
C7+	100	200

```
SHOrtcut 4 = DEC4,3,-4,-5, TOTal,
       RECovery to TOP 6 =.98,
       RECovery to BOTtom 7 = .95,
       DELta pressure = 5,
       PREssure of TOP product = 200,
       PERcent of MINimum reflux = 500

RESULTS FROM SHORT CUT FRACTIONATION FOR A TOTAL CONDENSER

+++DISTILLATE LIQUID PRODUCT+++

                     LBMOL/HR    MOL PERCENT      LB/HR
                     --------     --------     --------
 1 PROPANE           2000.0       66.741       88188.
 2 I-BUTANE          399.23       13.322       23203.
 3 N-BUTANE          588.00       19.622       34175.
 4 I-PENTANE         7.5000      0.25028       541.10
 5 N-PENTANE         1.9419      6.48011E-02   140.10
 6 N-HEXANE         1.22791E-04  4.09760E-06  1.05812E-02
 7 C7+              2.00000E-07  6.67408E-09  2.72000E-05
                      ------       ------       ------
      TOTALS         2996.7       100.00      1.46247E+05

TEMPERATURE   DEG F    126.58
PRESSURE  PSIA         200.00
ENTHALPY  BTU/HR     -1.56489E+07
MOLECULAR WT           48.803

+++BOTTOM PRODUCT+++

                     LBMOL/HR    MOL PERCENT      LB/HR
```

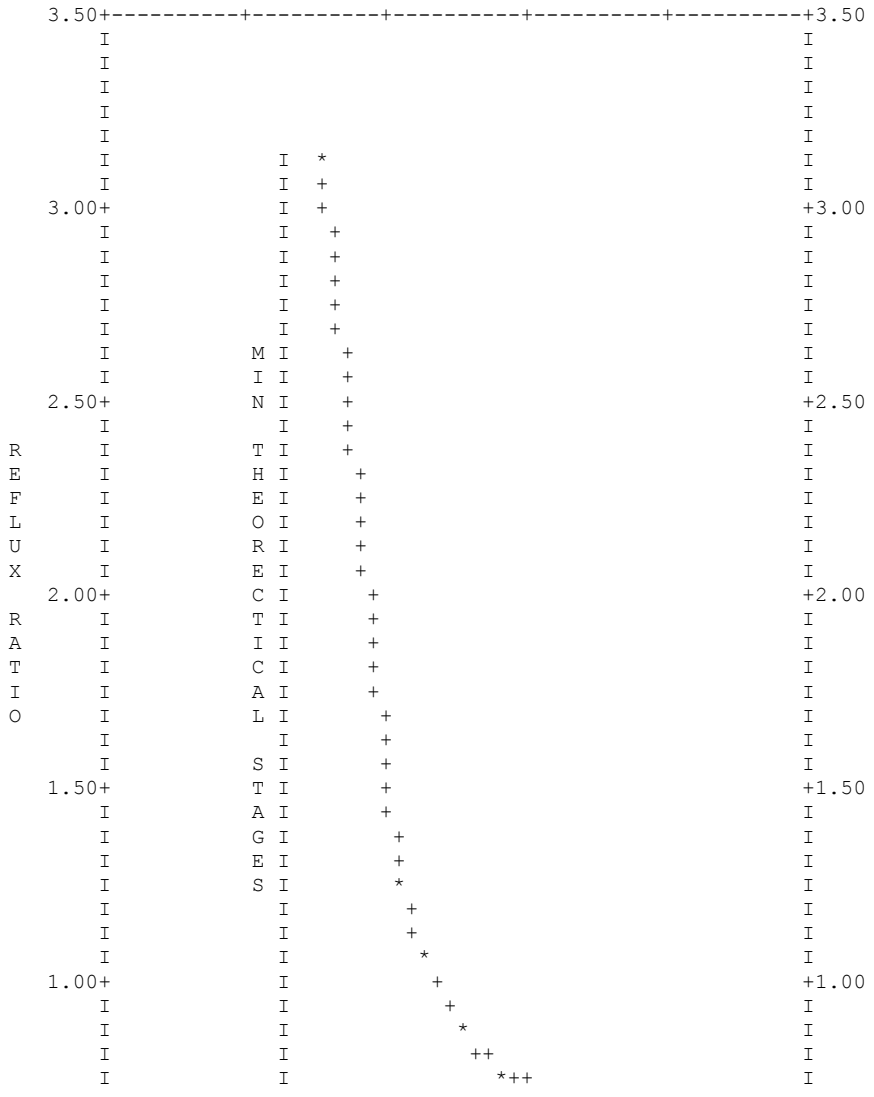
1 PROPANE	5.73730E-03	9.50934E-04	0.25298
2 I-BUTANE	0.76999	0.12762	44.752
3 N-BUTANE	12.000	1.9889	697.44
4 I-PENTANE	142.50	23.619	10281.
5 N-PENTANE	198.06	32.827	14289.
6 N-HEXANE	50.000	8.2873	4308.6
7 C7+	200.00	33.149	27200.
TOTALS	603.33	100.00	56821.

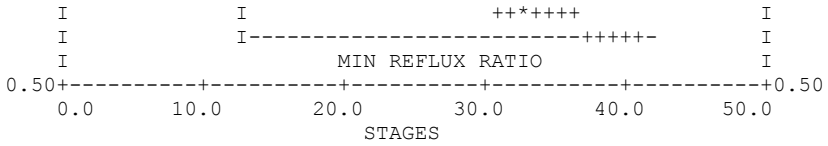
TEMPERATURE DEG F 338.05
PRESSURE PSIA 205.00
ENTHALPY BTU/HR 4.77982E+05
MOLECULAR WT 94.178

+++COLUMN PARAMETERS BY SHORT CUT METHOD+++

% OF MIN REFLUX	REFLUX RATIO	STAGES	FEED TRAY LOCATION	CONDENSER DUTY BTU/HR	REBOILER DUTY BTU/HR
105.0	0.654	41.38	23.58	0.3288E+08	0.4635E+08
110.0	0.686	32.71	18.74	0.3350E+08	0.4697E+08
120.0	0.748	28.59	16.43	0.3474E+08	0.4821E+08
140.0	0.872	25.45	14.67	0.3721E+08	0.5069E+08
175.0	1.091	22.60	13.08	0.4155E+08	0.5502E+08
200.0	1.246	21.27	12.34	0.4464E+08	0.5812E+08
500.0	3.116	15.77	9.26	0.8180E+08	0.9527E+08

REFLUX VERSUS STAGES FOR SHORTCUT FRACTIONATOR 1



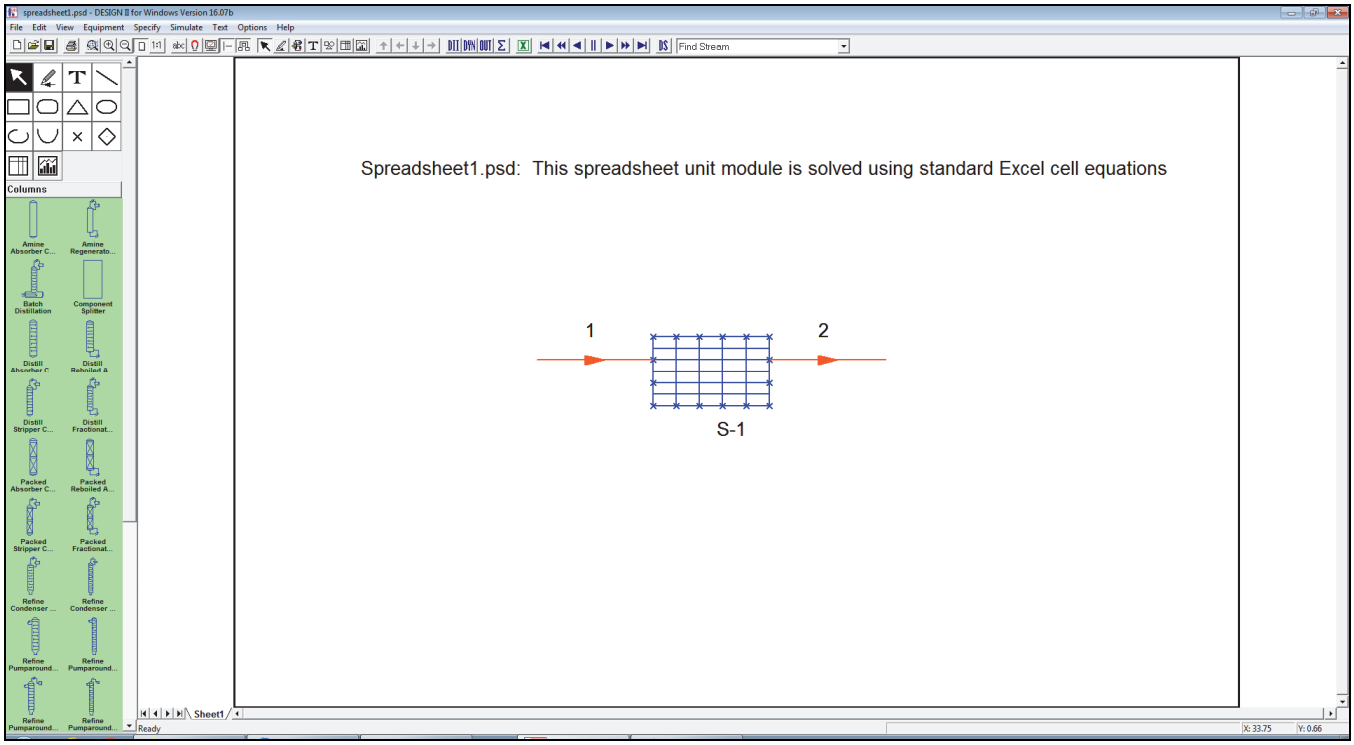

MIN REFLUX RATIO
0.50
0.50
0.0
10.0
20.0
30.0
40.0
50.0
STAGES

Chapter 48

Chapter 49: Spreadsheet Samples - c:\designii\samples\equipmnt\spreadsheet

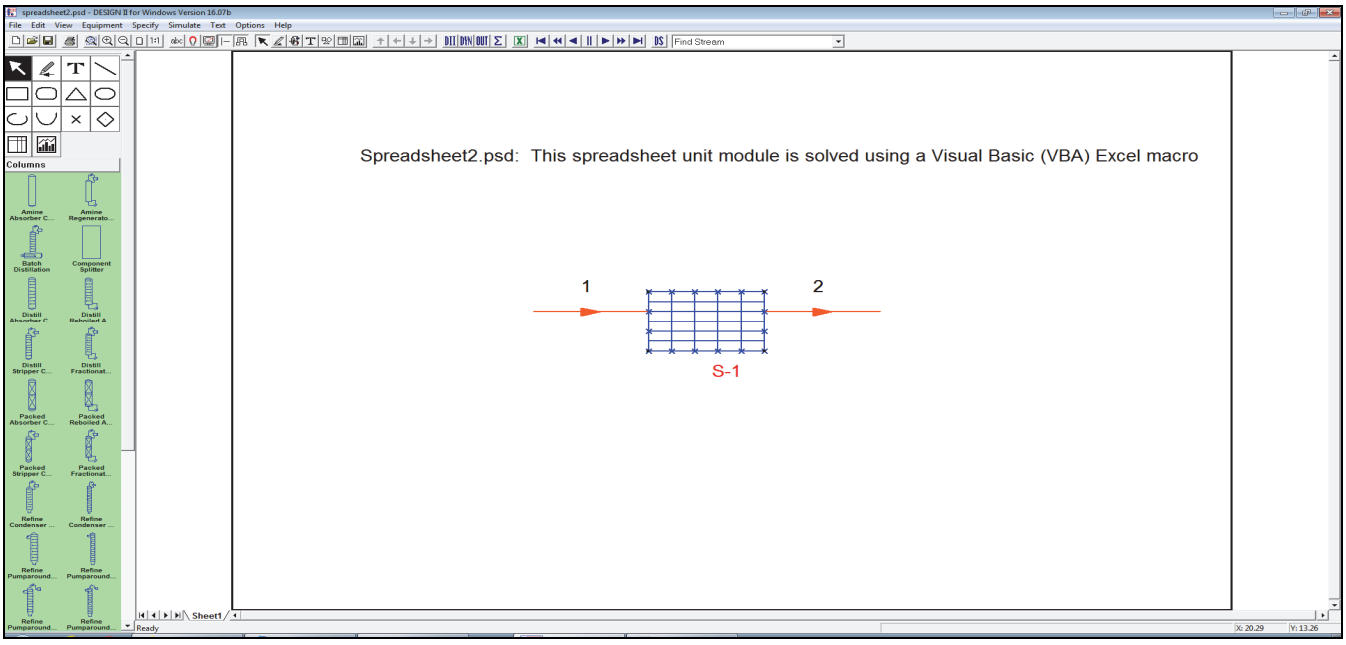
Spreadsheet1.psd

This spreadsheet unit module is solved using standard Excel cell equations.



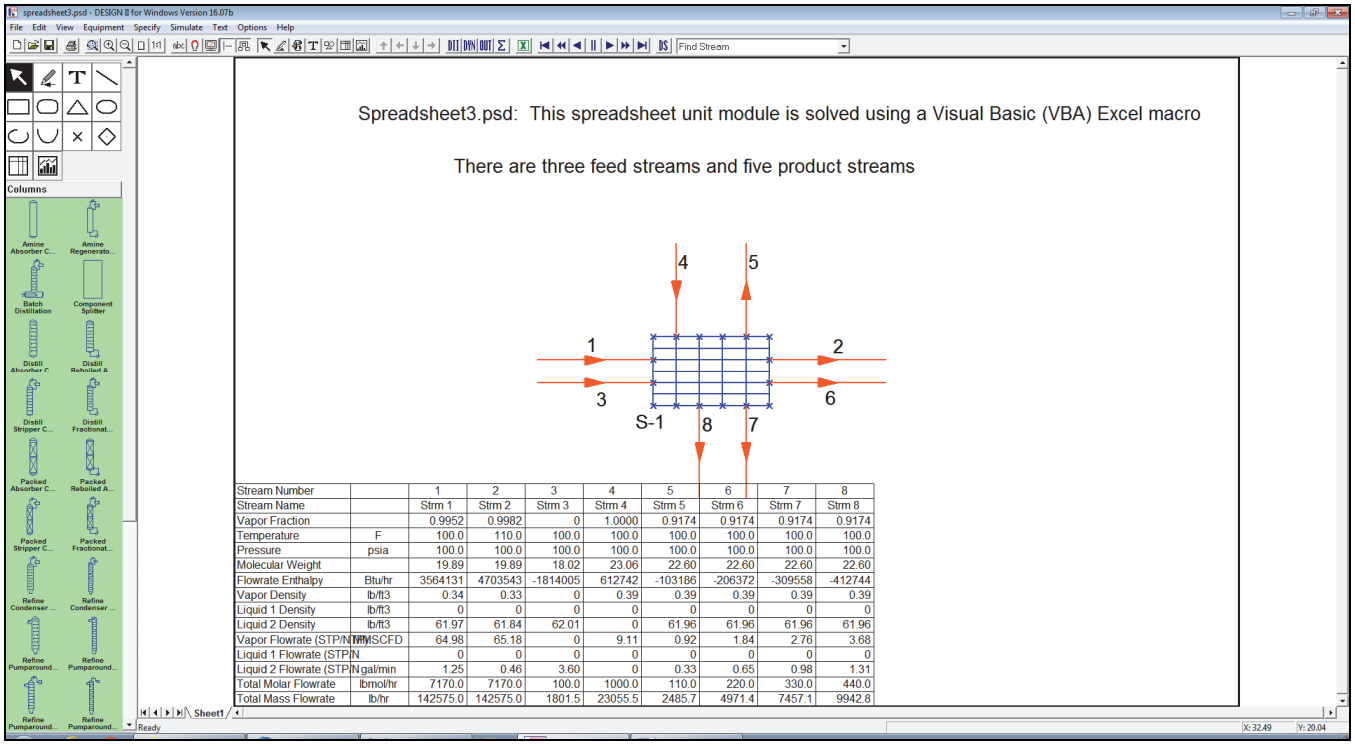
Spreadsheet2.psd

This spreadsheet unit module is solved using a Visual Basic (VBA) Excel macro.



Spreadsheet3.psd

This spreadsheet unit module is solved using a Visual Basic (VBA) Excel macro. There are threefeed streams and five product streams.

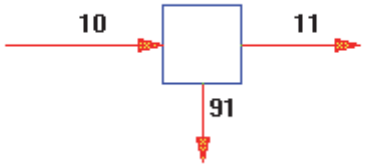


Stream Number		1	2	3	4	5	6	7	8
Stream Name		Strm 1	Strm 2	Strm 3	Strm 4	Strm 5	Strm 6	Strm 7	Strm 8
Vapor Fraction		0.9952	0.9982	0	1.0000	0.9174	0.9174	0.9174	0.9174
Temperature	F	100.0	110.0	100.0	100.0	100.0	100.0	100.0	100.0
Pressure	psia	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Molecular Weight		19.89	19.89	18.02	23.06	22.60	22.60	22.60	22.60
Flowrate Enthalpy	Btu/hr	3564131	4703543	-1814005	612742	-103186	-206372	-309558	-412744
Vapor Density	lb/ft3	0.34	0.33	0	0.39	0.39	0.39	0.39	0.39
Liquid 1 Density	lb/ft3	0	0	0	0	0	0	0	0
Liquid 2 Density	lb/ft3	61.97	61.84	62.01	0	61.96	61.96	61.96	61.96
Vapor Flowrate (STP/N)	MMSCFD	64.98	65.18	0	9.11	0.92	1.84	2.76	3.68
Liquid 1 Flowrate (STP/N		0	0	0	0	0	0	0	0
Liquid 2 Flowrate (STP/N	gal/min	1.25	0.46	3.60	0	0.33	0.65	0.98	1.31
Total Molar Flowrate	lbmol/hr	7170.0	7170.0	100.0	1000.0	110.0	220.0	330.0	440.0
Total Mass Flowrate	lb/hr	142575.0	142575.0	1801.5	23055.5	2485.7	4971.4	7457.1	9942.8

Chapter 50: Stream Manipulator Samples - c:\designii\samples\equipmnt\strman

Strman1.psd - Calculate Heating Value Of Stream For Fuel

Each component flowrate is multiplied by its heat of combustion (gross) in calculating the BTU/HR for the total stream. The BTU/LBMOL is then calculated by dividing the total BTU/HR by the total flow in LBMOL/HR. The BTU/SCF is then calculated by dividing the BTU/LBMOL by a SCF/LBMOL conversion factor.



Coding

```
STReam manipulator 5 = FUEL,10,-91,-11,
     HEAting VALue
GENeral,COMponents = 46,49,2,3,4,5,6,7,8,10
```

Method of Calculation

Both the BTU/LBMOL and the BTU/SCF are reported in the equipment summary. Consistent units are reported for metric and SI units. The heating value (in BTU/SCF) is reported as the equipment calculations proceed so the value can be used in a CONTroller operation.

Alternatively, you can enter your own values for the heats of combustion and choose the calculation option MULtiply to calculate the heating value of a stream. This is shown on the following page.

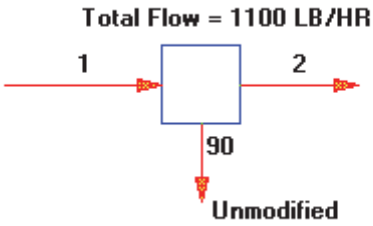
Components	Heat of Combustion BTU/LBMOL
N2	0
CO2	0
CH4	344994
C2	613939
C3	878937
i-C4	1138887
n-C4	1142417
i-C5	1403468
n-C5	1406959
n-C6	1671160

```
STReam manipulator 5 = FUEL,10,-91,-11,
       MULtiply,
       FACtor=0,0,344994,613939,878937,1138887,
              1142417,1403468,1406959,1671160
GENeral,
       COMponents = 46,49,2,3,4,5,6,7,8,10
```

NOTE: Stream 91 will be identical to the feed stream (stream 10). Stream 11 will contain "flowrates" for each component which are actually the Btu/hr available for that component flow, if burned.

Strman2.psd - Specifying by Composition and Total Flow

Feed stream composition can be entered on a mole, weight, or volume fraction or percent basis with a FLO command in the GENeral command section, and the STReam manipulator module can then be used to enter the total flow. This is also a useful feature when you want to use a CONTroller to vary the total feedstream flowrate. If the feed is to be in weight or volume fraction, weight or volume units must be specified with the FLO command in the GENeral section.




```
STReam manipulator 40 = TOTF,1,-90,-2
        MULtiply,FACtor = 11.10,SAMe FACtor
        GENeral, FLOw(LB/HR) 1 = 90,9,1
```

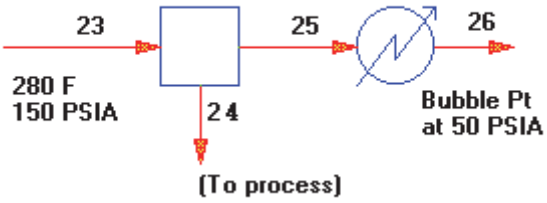
Method of Calculation

The weight percents of the components are converted to mass flowrates using a 100 LB/HR basis. These converted FLOws are entered in LB/HR with the FLOw command. Since a 100 LB/HR flow basis was used, the STReam manipulator then multiplies these flows by 11.10 to obtain the desired flow of 1110 LB/HR.

NOTE: Check SPECIFYING FEEDS Section for the TOTal FLOw option which is an alternative method for entering a specified flow with mass, molar, or volume fractions for the individual components.

Strman3.psd - Creating Two Identical Streams

Sometimes it is useful to create a stream identical to one of your process streams. For example, you may want to look at two different ways of processing the same stream. Or you may want to know a property of a process flow. To calculate the bubble point temperature of an intermediate product stream at the storage pressure:



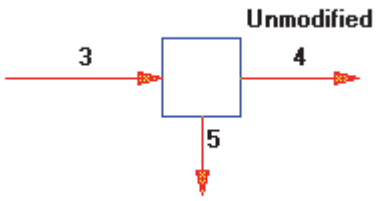

```
STReam manipulator 17 = PROD,23,-24,-25,
        MULtiply, FACtor = 1,SAMe FACtor
HEAt EXChanger 18 = STOR,25,-26,
        DELta = 100,
        TEMperature OUT is BUBble point
```

Method of Calculation

Stream 24 is created as the unmodified stream, and stream 25 is created by multiplying all component flows by a factor of one, to create a second identical stream. Stream 25 then goes to a HEAt EXChanger where the pressure is dropped from 150 PSIA and the bubble point temperature is calculated. Stream 24, the unmodified stream, would go on to the next equipment in the process.

Strman4.psd - Black Box Material Balance

Since the STReam manipulator can be used to add, subtract, multiply, or divide each component flowrate by its own factor, complicated reactors or entire sections of plants can be black boxed for material balance purposes. The dehydration and treating sections of a gas processing plant are to be black boxed for material balance.



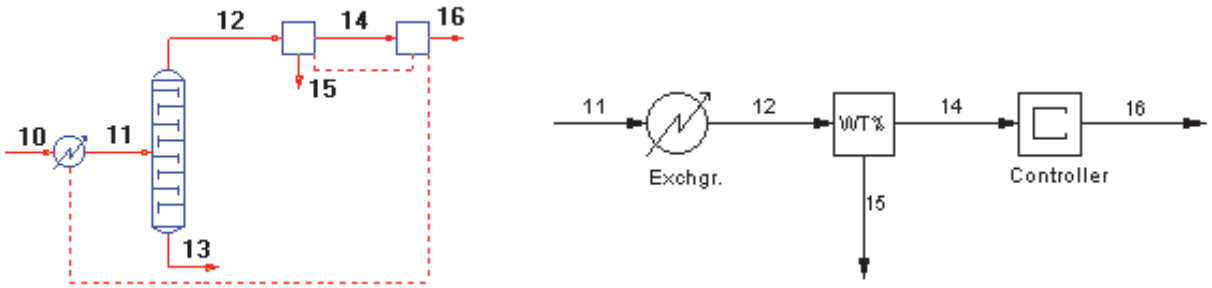
```
STReam manipulator 3 = DH+T,3,-4,-5,
       MULtiply, FACtor = .25,0,9*1,.98,0
GEN, COM = 49,50,2,3,4,5,6,7,8,9,10,11,62
```

Method of Calculation

Twenty-five percent of the first component (CO2), none of the H2S, and 100% of the next nine components (N2 through n-C5) are sent to stream 5. None of the water and 98% of the C6+ (to allow for 2% absorption of the hydrocarbons in the glycol or treating solvent) is sent to stream 5. The component flowrates removed from stream 3 in creating stream 5 do not appear anywhere in the output.

Strman5.psd - Mass Flowrate For Access By The Controller Module

Normally, the CONTroller module will only work with internal flowrates that are molar. This example shows how you can get CONTroller to access mass flowrates.



This example shows how you can get CONTroller to access mass flowrates.

The individual flowrates for each component in stream 12 will be varied by the controller via the multiplication FACtor of the stream manipulator to achieve a flow of 37000 lb/hr of the component ID# 6 (n-Butane) in stream 14.

Coding

```
AB123.
*Mass flowrate for access by the CONTroller module

C- EQUIPMENT MODULES

STR 1 = WT%, 12,-15,-14,
  MUL
  SAM FAC, FAC = 1.

CONT 2 = Controller, 14,-16,
  VARY FAC (MIN=0.,MAX=100.,STE=5.) OF STR 1
  UNTIL FLO COM 6 (SCA = 1.) FROM STREAM 14 =
  37000. (BEG=1,LOO=30,TOL=0.1,LB/HR)

HEA EXC 3 = Exchgr., 11,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 80.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 2.

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,6,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'N-BUTANE'
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
```

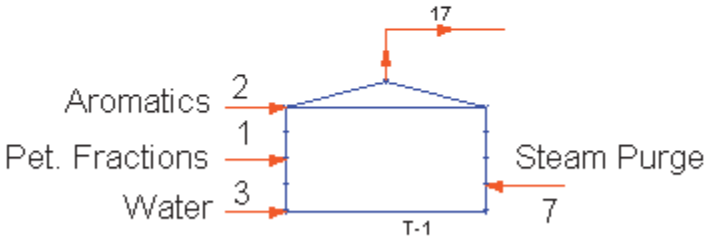
```
  CON TOL = 1.E-003
  FLO(LBMOL/HR)11 = 100.,26.,38.,11.,
  TP (F,PSIA) 11 = 70.,120.
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
END
```

Chapter 51: Tank Samples - c:\designii\samples\equipmnt\tank

Tank1.psd - Simulation of Heating a Tank

(From the DESIN II Menu Bar, please select...File...Open... c:\designii\samples\equipmnt\tank\tank1.psd to simulate this problem)

BASIS: 1 hr.



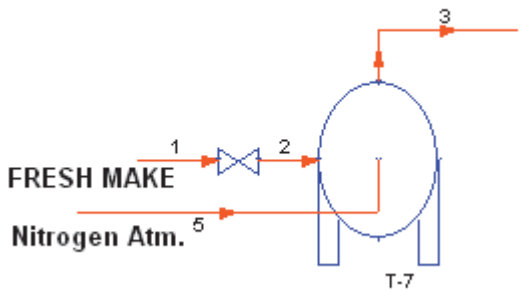
Heating a tank to simulate a dynamic system.

```
AB123.
*Simulation of Heating a Tank - dynamic

C- EQUIPMENT MODULES

TANK 1 = T-1, 7,3,1,2,-17,
  VEN PRE(PSIA) = 14.696
  VER
  LEN(FT) = 6
  DIA(FT) = 20
  INI CHA(FRA) = 18*0,0.79,0.21,
  INI TEM(F) = 70
  INI PRE(PSIA) = 14.696
  PRO TYP 1 = VAP
  STA TIM 1(HR) = 0
  STO TIM 1(HR) = 0
  PRO TYP 2 = VAP
  STA TIM 2(HR) = 0
  STO TIM 2(HR) = 0
  PRO TYP 3 = VAP
  STA TIM 3(HR) = 0
  STO TIM 3(HR) = 0
  PRO TYP 4 = VAP
  STA TIM 4(HR) = 0
  STO TIM 4(HR) = 0
  PRO TYP 5 = VAP
  STA TIM 5(HR) = 0
  STO TIM 5(HR) = 0


C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,1176,3111,3050,1140,3058,40,41,43,45,38,5003,10,8,11,12,13,14,
            46,47,100,101
  NAM POS 1 = 'WATER'
  NAM POS 2 = 'CCL2F2'
  NAM POS 3 = '1,1,1-CL3 ETHANE'
  NAM POS 4 = 'CL2HC-CH2CL'
  NAM POS 5 = 'CL3 ETHYLENE'
```

```
  NAM POS 6 = 'CL4 ETHYLENE'
  NAM POS 7 = 'BENZENE'
  NAM POS 8 = 'TOLUENE'
  NAM POS 9 = 'M-XYLENE'
  NAM POS 10 = 'ETHYLBENZENE'
  NAM POS 11 = 'CYCLOHEXANE'
  NAM POS 12 = 'NAPHTHALENE'
  NAM POS 13 = 'N-HEXANE'
  NAM POS 14 = 'N-PENTANE'
  NAM POS 15 = 'N-HEPTANE'
  NAM POS 16 = 'N-OCTANE'
  NAM POS 17 = 'N-NONANE'
  NAM POS 18 = 'N-DECANE'
  NAM POS 19 = 'NITROGEN'
  NAM POS 20 = 'OXYGEN'
  NAM POS 21 = 'C7PLUS'
  AMW 100 = 97
  AMB (F) 100 = 187
  API (API) 100 = 70
  NAM POS 22 = 'C8PLUS'
  AMW 101 = 107
  AMB (F) 101 = 225
  API (API) 101 = 61.6
C-  Thermo Comments
  STDK
  STDH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  FLO(LB/HR)1 = 20*0,50,75,
  TP (F,PSIA) 1 = 70,14.7
  NAM STR 1 = Pet. Fractions
  FLO(LBMOL/HR)2 = 0,1.5,20,1,18,27,90,102,75,5,3,40,2,15,10,7,14,5,2*0,
  TP (F,PSIA) 2 = 80,14.7
  NAM STR 2 = Aromatics
  FLO(GAL/DAY)3 = 200,
  TP (F,PSIA) 3 = 80,14.7
  NAM STR 3 = Water
  FLO(LB/HR)7 = 150,
  TP (F,PSIG) 7 = 600,150
  NAM STR 7 = Steam Purge
  NAM STR 17 = Tank Vent
END
```

Tank2.psd - Tank Vent with Nitrogen Purge

(From the DESIGN II Menu Bar, please select...File...Open... c:\designii\samples\equipmnt\tank\tank2.psd to simulate this problem)



Tank Vent with Nitrogen Purge

```
AB123.
```

```
*Tank Vent with Nitrogen Purge

C- EQUIPMENT MODULES

VAL 1 = FRESH MAKE, 1,-2,
  PRE OUT(BAR) = 1.03

TANK 7 = T-7, 2,5,-3,
  VEN PRE(BAR) = 1.03
  SPH
  DIA(M) = 32.961
  HEA(KCAL/HR) = 2601500
  INI CHA(LBMOL) = 0,240000,129600,28800,12000,9600,7200,4800,
  INI TEM(F) = -260
  INI PRE(BAR) = 1.03
  PRO TYP 1 = VAP
  STA TIM 1(HR) = 0
  STO TIM 1(HR) = 0
  PRO TYP 2 = VAP
  STA TIM 2(HR) = 0
  STO TIM 2(HR) = 0
  PRO TYP 3 = VAP
  STA TIM 3(HR) = 0
  STO TIM 3(HR) = 0
  PRO TYP 4 = VAP
  STA TIM 4(HR) = 0
  STO TIM 4(HR) = 0
  PRO TYP 5 = VAP
  STA TIM 5(HR) = 0
  STO TIM 5(HR) = 0


C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,2,3,4,5,6,7,8
  NAM POS 1 = 'NITROGEN'
  NAM POS 2 = 'METHANE'
  NAM POS 3 = 'ETHANE'
  NAM POS 4 = 'PROPANE'
  NAM POS 5 = 'I-BUTANE'
  NAM POS 6 = 'N-BUTANE'
  NAM POS 7 = 'I-PENTANE'
  NAM POS 8 = 'N-PENTANE'
C-  Thermo Comments
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  AME UNI OUT
  MAX = 20
  CON TOL = 0.001
  FLO(LBMOL/HR)1 = 0,1000,540,120,50,40,30,20,
  TP (F,BAR) 1 = -260,1.05
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Vent Flow
  FLO(LBMOL/HR)5 = 1e-005,
  TP (F,BAR) 5 = -100,1.05
  NAM STR 5 = Nitrogen Atm.
END
```

Chapter 52: Tank Sample/Dynamic - c:\designii\samples\dynamic

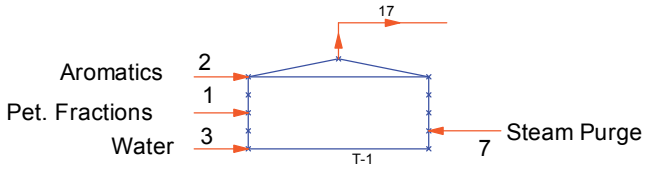
C:\designii\samples\dynamic\Tank1.psd

Heating of a tank to simulate a dynamic case for an adiabatic flash.

Simulation of Heating a Tank

All flashes are adiabatic (no heat exchanged)
Constant Feed Steam rate as Heat Source

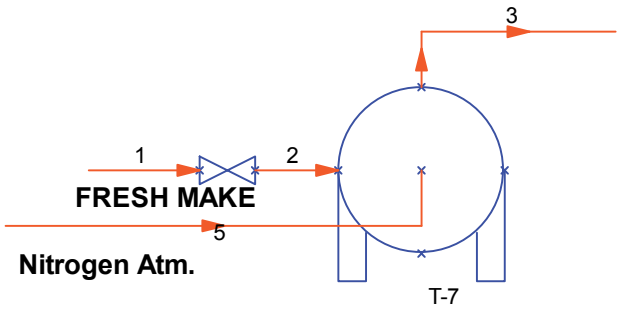
BASIS: 1 hr.



C:\designii\samples\dynamic\Tank2.psd

Simulation of a tank vent with a dynamic nitrogen purge system.

TANK Vent with Nitrogen Purge



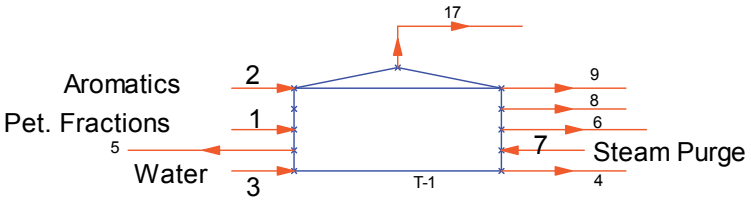
C:\designii\samples\dynamic\Tank3.psd

Simulation of a tank with various timed multiple product draws.

Tank with Timed Multiple Product Draws

All flashes are adiabatic (no heat exchanged)
Constant Feed Steam rate as Heat Source

BASIS: 1 hr.



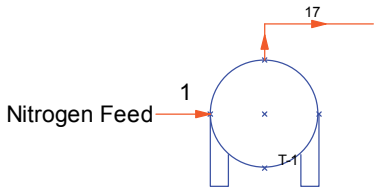
C:\designii\samples\dynamic\Tank7.psd

Simulation of pressurizing a tank for two hours.

Simulation of Pressurizing a Tank

All flashes are adiabatic (no heat exchanged)
Constant Feed Steam rate as Heat Source

BASIS: 24 hr.



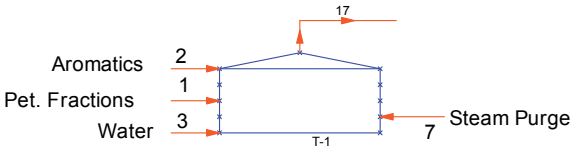
C:\designii\samples\dynamic\Tank8.psd

Heating of a tank to simulate an unequal time duration.

Simulation of Heating a Tank

All flashes are adiabatic (no heat exchanged)
Constant Feed Steam rate as Heat Source

BASIS: 1 hr with unequal time durations



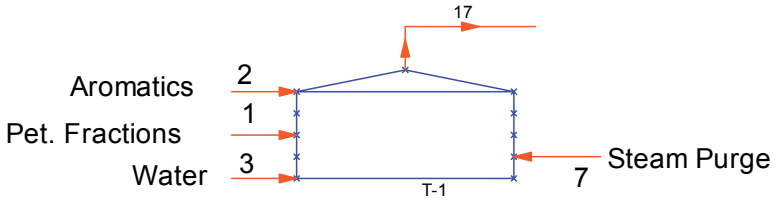
C:\designii\samples\dynamic\Tank9.psd

Heating of a tank to simulate an isothermal case with calculated cooling duty.

Simulation of Heating a Tank

All flashes are isothermal (constant temperature at 80 F)
Constant Feed Steam rate as Heat Source and
the cooling duty will be calculated

BASIS: 1 hr.



Chapter 52

Chapter 53: Valve Samples - c:\designii\samples\equipmnt\valve

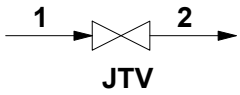
Valve1.psd - Specified Pressure Out with Phase Separation

All properties of stream 1 are known. A constant enthalpy (adiabatic) flash is performed to 200 PSIA and the outlet temperature is calculated. The vapor and liquid portions of the resulting stream are separated into streams 5 and 6.


```
VALve 3 =FLSH,4,-5,-6, PREssure OUT = 200
```

Valve2.psd - Joule-Thompson Pressure Drop

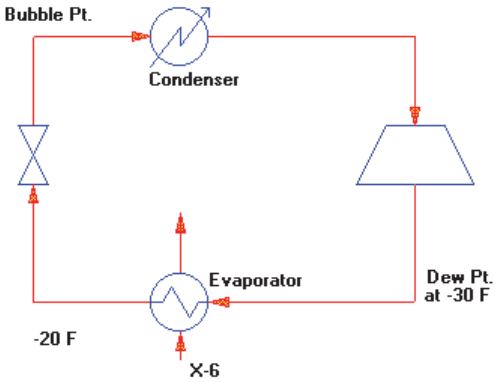
All properties of stream 1 are known. The pressure drop (50 PSIA) is subtracted from the feed pressure and an adiabatic flash performed. The outlet temperature is calculated.




```
VALve 2 = JTV,3,-4, DELta pressure = 50
```

Valve3.psd - Refrigeration Cycle J-T Valve

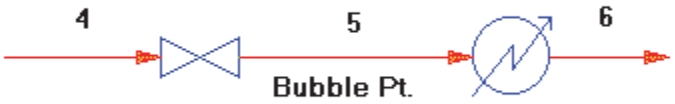
All properties of stream 3 are known. The dew point pressure is calculated at the specified temperature (-30F). An adiabatic (constant enthalpy) flash is performed to this pressure and the resulting temperature calculated. Stream 4 will be two-phase; the liquid portion will be evaporated to achieve chilling of the process stream.




```
VALve 3 = J-T,3,-4,
       PREssure OUT is DEWpoint,
       TEMperature DEWpoint = -30
```

Valve4.psd - Bubble Point Pressure at Specified Temperature

All properties of stream 4 are known. The bubble point is calculated at the specified temperature (100 F). An adiabatic (constant enthalpy) flash is performed to this pressure and the resulting temperature calculated.




```
VALve 3 = BUBl,4,-5,
        PREssure OUT is BUBble point,
        TEMperature BUBble point = 100,
HEAt EXChanger 4 = BUBP,5,-6,
        TEMperature OUT = 100
```

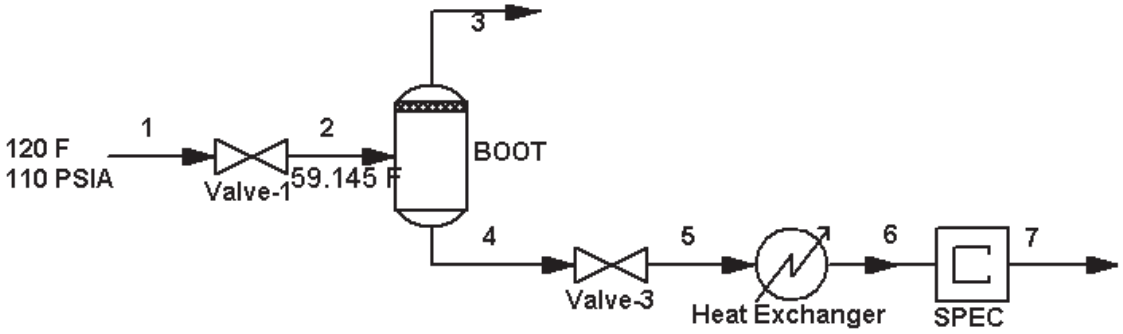
Method of Calculation

The HEAt EXChanger module will be used to adjust the enthalpy in stream 5 so that stream 6 will be single phase at the bubble point pressure. Alternatively, this may be done with a single module:

```
FLAsh 3 = BUB, 4, -6, TEMperature OUT = 100, BUBble = 6
```

Valve5.psd - Using Controller To Meet Desired Bubble Point Pressure

Using the CONTroller module, with a slightly modified flowsheet, one can change the composition of the condensate to attempt to meet a specified bubble point pressure at a specified temperature.



Using the CONTroller module, with a slightly modified flowsheet, one can change the composition of the condensate to attempt to meet a specified bubble point pressure at a specified temperature.

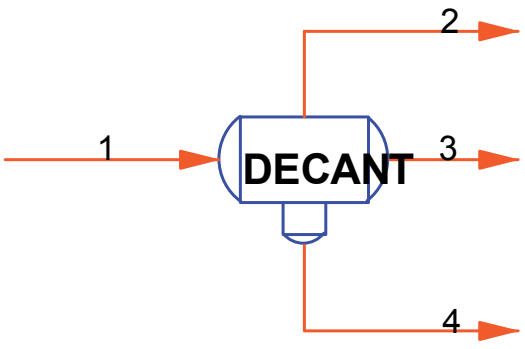
All properties of stream 1 are known. The pressure is adjusted and a new enthalpy is calculated for stream 2. Stream 2 is separated into a vapor and liquid phase. The bubble point pressure for stream 5 is calculated at the specified temperature (50F). An adiabatic (constant enthalpy) flash is performed to this pressure and the resulting temperature calculated. Stream 5 will be two-phase. The HEAt EXChanger module will be used to adjust the enthalpy in stream 5 so that stream 6 will be single phase at the bubble point pressure. The CONTroller module will vary the pressure out of VALve 1 until the PRE OUT BUB required in valve 3 is calculated to be 9.5 PSIA.

```
AB123.
*Using Controller To Meet Desired Bubble Point Pressure

C- EQUIPMENT MODULES

VAL 1 = Valve-1, 1,-2,
  PRE OUT(PSIA) = 5.

FLA 2 = BOOT, 2,-3,-4,

VAL 3 = Valve-3, 4,-5,
  PRE OUT BUB, TEM BUB(F) = 50.

HEA EXC 4 = Heat Exchanger, 5,-6,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT BUB
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
```

```
  HOR
  DEL(PSI) = 0.

CONT 5 = SPEC, 6,-7,
  VARY PRE OUT (MIN=5.,MAX=50.,STE=1.,PSIA) OF VAL 1
  UNTIL PRE (SCA = 1.) FROM STREAM 5 =
  9.5 (BEG=1,LOO=30,TOL=1.e-004,PSIA)

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,2,3,49,4,5,6,7,8,10,11,12,13,14,15,
  NAM 62 = 'WATER'
  NAM 46 = 'NITROGEN'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 49 = 'CO2'
  NAM 4 = 'PROPANE'
  NAM 5 = 'I-BUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 7 = 'I-PENTANE'
  NAM 8 = 'N-PENTANE'
  NAM 10 = 'N-HEXANE'
  NAM 11 = 'N-HEPTANE'
  NAM 12 = 'N-OCTANE'
  NAM 13 = 'N-NONANE'
  NAM 14 = 'N-DECANE'
  NAM 15 = 'N-UNDECANE'
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 0.,2.05,8.55,17.7,30.52,48.75,15.015,64.35,21.54,26.4,32.97,
  20.11,6.81,1.44,1.905,11.7,
  TP (F,PSIA) 1 = 120.,110.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Valve6.psd – Hydrocarbon – Water Decant

All properties of stream 1 are known. If there is any “free water” present, the vapor phase (stream 2) will be saturated. Stream 3 will contain liquid hydrocarbons and soluble water. Stream 4 will contain “free water” and soluble hydrocarbons. If CO2, H2S, and/or NH3 are present, the APISOUR correlation will be used to estimate their solubility in the water phase. Enthalpies and properties of the liquid streams will be recalculated.



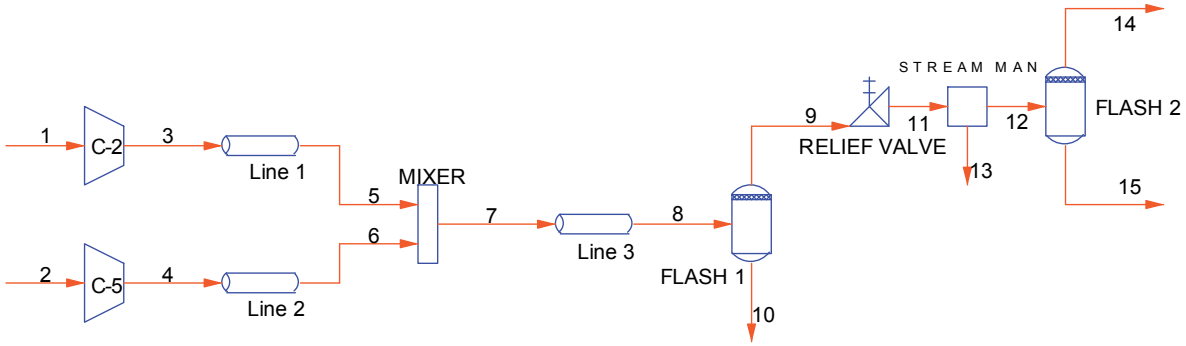

```
AB123.
*Hydrocarbon Water Decant

C- EQUIPMENT MODULES

VAL 1 = DECANT, 1,-2,-3,-4,
```

```
  PRE OUT(BAR) = 70

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,4,62,5,6,49,50,63,
  NAM 2 = 'METHANE'
  NAM 4 = 'PROPANE'
  NAM 62 = 'WATER'
  NAM 5 = 'I-BUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 49 = 'CO2'
  NAM 50 = 'H2S'
  NAM 63 = 'AMMONIA'
  SKDK
  SKDH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  IMM = 62
  QUA UNI OUT = LB
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  FLO(LB/HR)1 = 3,1,2*2,4*1,
  TP (F,BAR) 1 = 70,70
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
END
```

Chapter 54: Flareheader Samples-c:\designii\samples\flare\Flrehdr.psd

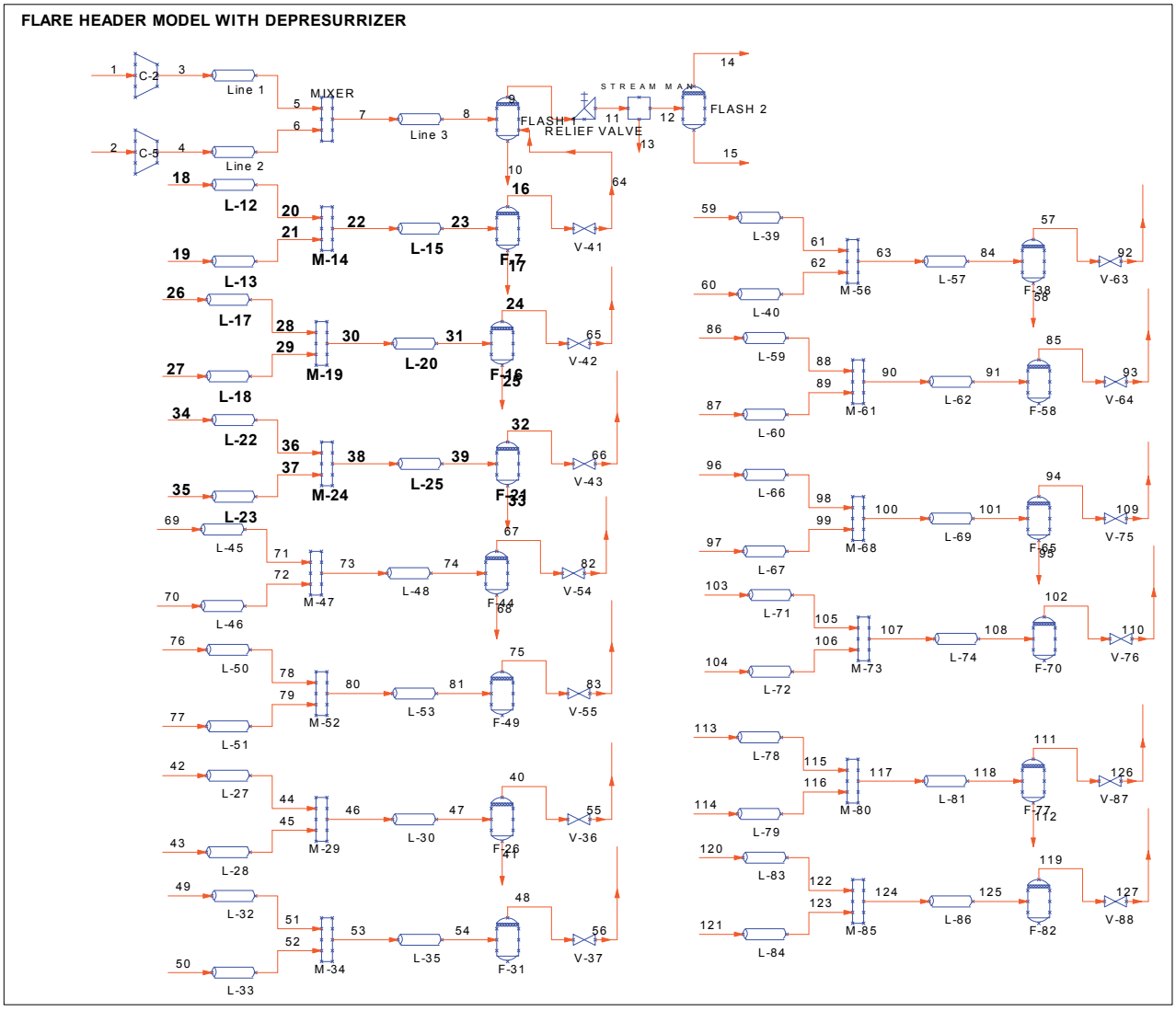
C:designii\samples\flare\Flrehdr.psd

The sample simulation shows two different lines being compressed for a typical flare header relief valve sizing calculation. Methods such as API 520 and API521 relief valve sizing can be performed via the depressurizer module. The depressurizer module produces the vapor and liquid compositions relieved from the set pressure to a pre-defined reduced pressure. Ambient pressure is the default for the backpressure.



C:designii\samples\flare\Flrehdr.psd

Flareheader scheme illustration for using depressurizer.



C:designii\samples\flare\PII Flare Header.psd

Design of Flare/Vent system simulated with line module, valve and respective controllers.

USE DESIGN II FOR PROCESS FLARE/VENT SYSTEM DESIGN & RATING

TO DESIGN A FLARE/VENT SYSTEM, THE RELIEF LOADS MUST BE KNOWN. THESE ARE NORMALLY DETERMINED BY ANALYSIS OF THE CREDIBLE RELIEF SCENARIOS (BLOCKED OUTLET, FIRE, ETC.) FOR THE SUBJECT EQUIPMENT..THE SCENARIO THAT GENERATES THE HIGHEST RELIEF RATE IS THE "WORST CASE" FOR THE RELIEVING DEVICE (PSV, RUPTURE DISK, ETC) AND PROVIDES THE BASIS FOR DEVICE "SIZING". THIS WORK IS PERFORMED IN ACCORDANCE WITH API 520 (OR OTHER APPLICABLE PROCEDURE). DURING THIS WORK, DESIGN II IS USED TO GENERATE PHYSICAL PROPERTY DATA FOR THE RELIEVING FLUID (UPSTREAM OF THE PSV). THE RELIEF DEVICE DEVICE MAY BE SIZED WHEN THIS WORK IS COMPLETE. SIZING WILL PROVIDE THE PSV INLET/OUTLET PIPING SIZE AND, IF "LOCAL VENTING" IS ADEQUATE, THE DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) WILL ALSO BE KNOWN.

ENVIRONMENTAL CONCERNS, HOWEVER, ARE REQUIRING THAT MORE FACILITIES ROUTE MOST - IF NOT ALL - OF THEIR RELIEF STREAMS TO A FLARE/VENT SYSTEM. IT IS NO LONGER SAFE TO ASSUME THAT, IF A RELIEVING FLUID IS "LIGHTER THAN AIR", THEN A "LOCAL VENT" WILL SUFFICE.

THERE ARE A FEW "SPECIALTY" FLARE/VENT HEADER DESIGN PROGRAMS, AND SOME DO THIS JOB VERY WELL BUT ARE ALSO QUITE EXPENSIVE. OTHERS MAY BE LESS EXPENSIVE, BUT ARE LIMITED IN SCOPE. DESIGN II PROVIDES A VERY SATISFACTORY SOLUTION WITH HEADER DESIGN PERFORMED USING THE PROCESS SIMULATOR THAT PROVIDED THE ORIGINAL PROCESS DESIGN AS WELL AS THE RELIEF LOAD PROPERTIES.

THE STEPS INVOLVED IN FLARE/VENT HEADER DESIGN ARE:

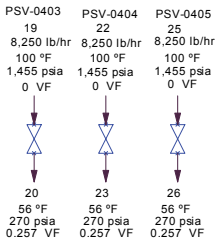
1. FINALIZE THE PHYSICAL LAYOUT OF THE EQUIPMENT THAT WILL RELIEVE TO THE HEADER AS WELL AS THE LOCATION OF THE FLARE/VENT STACK (I.E., PLOT PLAN REQUIRED).

2. PERFORM A PRELIMINARY HEADER PIPING DESIGN TO GENERATE HEADER PIPE SPOOL LENGTHS. ENSURE THAT THE EQUIVALENT PIPING LENGTHS OF FITTINGS ARE INCLUDED. FOR PIPE DIAMETERS, USE PSV OUTLET SIZES - AS DETERMINED ABOVE - FOR THE PSV "TAILPIPES" AND YOUR BEST ESTIMATE FOR THE MAIN HEADER.

3. GENERATE CREDIBLE FLARE/VENT HEADER LOADING SCENARIOS. THIS WORK IS NORMALLY PERFORMED BY A "TEAM". THE TEAM WILL DETERMINE THE AREA(S) AND SITUATIONS FOR ALL FLARE/VENT EVENTS THAT CAN OCCUR. NOTE THAT THE DESIGN II OUTPUT PROVIDES AN EXCELLENT TOOL FOR DOCUMENTING THIS WORK.

4. USING THE PHYSICAL PROPERTY AND PIPING DATA GENERATED FROM THE ABOVE PROCEDURES, CREATE A DESIGN II "CASE" FOR EACH FLARE/VENT HEADER LOADING SCENARIO. THIS WORK WILL REQUIRE SEVERAL ITERATIONS BEFORE AN OPTIMUM DESIGN IS ACHIEVED. AFTER EACH ITERATION, CHECK THE BACK-PRESSURE ON EACH DEVICE THAT "FEEDS" THE HEADER SYSTEM THEN ADJUST THE DESIGN II PIPE SIZES AS REQUIRED TO ACHIEVE A MINIMUM "SYSTEM COST". THE SYSTEM COST INCLUDES THE COSTS FOR ALL COMPONENTS IN THE FLARE/VENT SYSTEM - INCLUDING RELIEF DEVICES AND THE FLARE STACK.

WHEN COMPLETE, THE DESIGN II OUTPUT IS PASSED TO PIPING DESIGN FOR FINAL FLARE/VENT HEADER WORK. ALSO NOTE THAT THE RELIEF DEVICE "TYPE" (CONVENTIONAL, PILOT-OPERATED, ETC.) HAS BEEN DETERMINED AND THAT RELIEF DEVICE SPECIFICATIONS CAN BE PASSED TO QUALIFIED VENDORS FOR PRICING. FINALLY, THE FLARE/VENT STACK INLET STREAM DATA FOR EACH SCENARIO SHOULD BE TRANSMITTED TO THE STACK FABRICATOR FOR DESIGN AND PRICING.



DISCONNECT PSV-0403/4/5 TAILPIPES FROM FLARE/VENT HEADER AND ROUTE TO PRODUCT SURGE DRUM

THIS IS A FLARE/VENT SYSTEM RATING CASE

THE ABOVE-NOTED PSV-0403/0404/0405 WERE RE-ROUTED TO THE NGL PRODUCT SURGE DRUM AS NOTED ABOVE. NO OTHER CHANGES TO THE SYSTEM WERE REQUIRED.

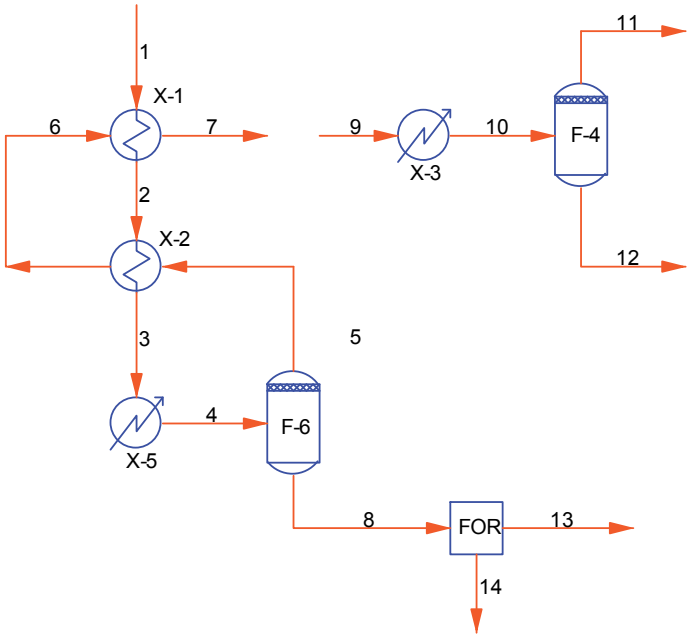
DATE:	DRAWN BY:	CHECKED BY:	REVISION:
28 Nov 12	Young		
Process Tech, Inc.			
PROJECT: Panola II Flare/Vent Header Rating			

Chapter 54

Chapter 55: Inline Fortran Samples- C:\designii\samples\fortran

Inlfrn1.psd - Gas Plant Feed Chilldown

Two feed streams to a gas plant are to be chilled to -30 F using propane refrigeration. The two feed streams are shown in the figure above. The feed streams and equipments are simulated on the same flowsheet.) Inline Fortran is used to retrieve calculated values from the simulation run and manipulate them to calculate TONS OF REFRIGERATION REQUIRED, SALES GAS REVENUE, and REFRIGERATION COST.



FEED STREAMS	1	9
Component	Flow (Lbmol/Hr)	
N2	98 1	
CO2	45015	
C1	27400	600
C2	2130	75
C3	71321	
i-C4	1102	
n-C4	1453	
i-C5	560.5	
n-C5	430.4	
n-C6	800.2	
C7+	10 --	
Temperature (F)	6020	
Pressure (PSIA)	1000	500

Also provide flow and TP data for assigned recycle streams 2 and 4. Estimate for recycle stream 2 temperature is 30 F.

```
AB123.
*REFRIGERATION AND SALES REVENUE CALCULATION
HEAt EXChanger 1 = EXC1, 6,1,-7,-2,
       SIMple, AREA = 5000, DELta = 5,10
HEAt EXChanger 2 =EXC2, 5,2,-6,-3,
       SIMple, AREa = 5000, DELta = 5,10
HEAt EXChanger 3 =EXC3, 3,-4,
       TEMperature OUT = -30, DELta = 10
FLAsh 4 = FLA4, 4,-5,-8
HEAt EXChanger 5 =EXC5, 9,-10,
       TEMperature OUT = -30, DELta = 10
FLAsh 6 = FLA6, 10,-11,-12
STReam manipulator 7 = FOR, 8,-13,
```

```
       MULtiply, FACtor = 1.0, SAMe FACtor
C-*****************IN-LINE FORTRAN SECTION***************
C-****TAKE SIMULATION RESULTS AND CALCULATE REFRIGERATION***
C-****REQUIREMENTS AND SALES GAS REVENUE*******************
F-      POST
F-      C3DUT1 = GET(HEA EXC(3),CAL DUT)
F-      C3DUT2 = GET(HEA EXC(5),CAL DUT)
F-      SUMDUT = C3DUT1 + C3DUT2
F-      TONS = -SUMDUT/12000
F-      WRITE(6,10)TONS
F- 10   FORMAT('1',///,1X,'**** TONS REFRIG REQUIRED = ',F8.2)
F-      GAS1 = GET(STR(7),FLO)*379.5
F-      GAS2 = GET(STR(11),FLO)*379.5
F-      GASDOL = ((GAS1 + GAS2)/1000.)*2.5
F-      WRITE(6,20)GASDOL
F- 20   FORMAT(1X,'********** SALES GAS REVENUE    = ',F8.2)
F-      REFDOL = -5.0*SUMDUT/1E6
F-      WRITE(6,30)REFDOL
F- 30   FORMAT(1X,'******** REFRIGERATION COST   = ',F8.2,///)
F-      CALL LINECK(9)
C-*******************************
GENeral,
       COMponents = 46,49,2,3,4,5,6,7,8,10,100
       AMB  100 = 250, AMW 100 = 140, API 100 = 75,
       NAMe 100 = C7+
       FLOw 1 = 98,450,27400,2130,713,110,145,56,43,80,10
       TP 1 = 60,1000
       RECycle STReam = 4,2,
       FLOw 4 = FLOw 1, TP 4 = -30,970
       CONvergence on TEMperature = 4,2,
       FLOw 2 = FLOw 1, TP 2 = 30,990
       FLOw 9 = 1,15,600,75,21,2,3,.5,.4,.2, TP 9 = 20,500
       MAXimum iterations = 10,
       APISOAVEK, APISOAVEH, STDD
END
```

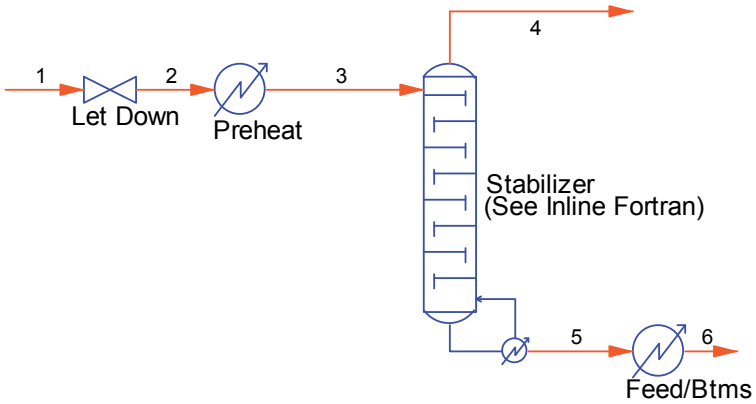
NOTE: A dummy equipment module (STR 7 = FOR) was inserted containing the Inline FORTRAN statements. Only the final results would be printed since this equipment module is not a part of the recycle loop. The Inline FORTRAN results are printed in the NOW CALLING section of the output. The results from the printout are shown below:

```
********** TONS REFRIG REQUIRED =  1915.91
********** SALES GAS REVENUE    = 28657.61
********* REFRIGERATION COST   =   114.95
```

Inlfrn2.psd - Reid Vapor Pressure Column Specification

This example creates a Reid vapor pressure specification for the DIStillation column using Inline FORTRAN. It calls the Design II subroutine REIDVP to calculate the Reid vapor pressure of the bottoms product. Note the use of the FLOWL(x,y) utility to retrieve component flowrates for liquid leaving the bottom stage of the column.

NOTE: There are three separate Reid vapor pressure calculation options the user can choose: REIDVP, REICRU, REIPRO. See ***Inline Fortran: Table 6. Thermodynamic Library Subroutines*** for details.



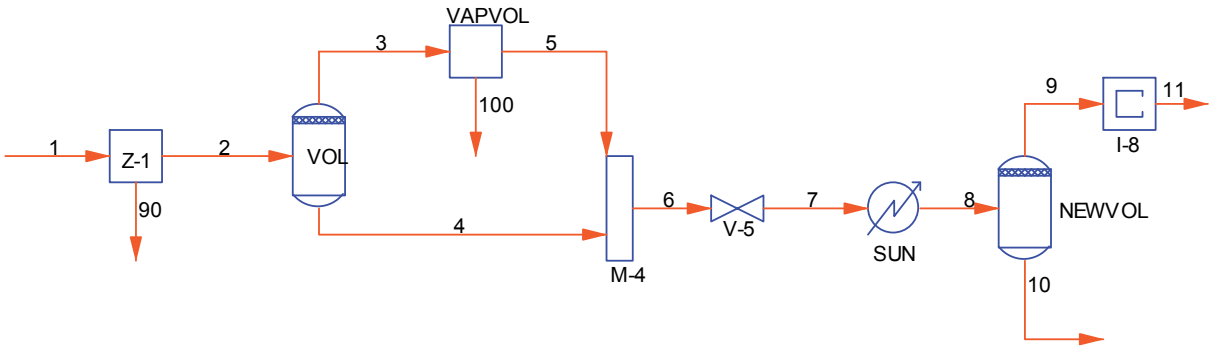

```
AB123.
* SAMPLE INLINE FORTRAN PROBLEM
VALve 1 = LETDOWN, 1, -2,
       PREssure OUT = 120
HEAt EXChanger 2 = FEED PREHEAT, 2, -3,
       TEMperature OUT = 120, DEL = 5
```

```
DIStillation 3 = STABILIZER, 3, -4, -5,
       ABSorbed REBoiler
       TRAys = 12,
       PREssure of TOP = 105, DELta pressure = 5
       TEMperature of TOP = 140, TEMperature of BOTtom = 275
       LOCate FEEd = 1, PROducts = 0, 116
       SUPer, PRInt SWItch = 0
C-
C-PURITY SPEC WILL BE REPLACED BY INLINE FORTRAN EPS(1) SPEC
C-FOR REID VAPOR PRESSURE -- BOTTOM PRODUCT RVP = 10
C-
       PURity of BOTtom 2 =.015
C-
F-     DURING
F-     CALL REIDVP(FLOWL(1,1),RVP,IC,ERR)
F-     EPS(1) = (RVP - 10.)/10.
C-
HEAt EXChanger 4 = BTMS, 5, -6,
C-
C-  INLINE FORTRAN WILL RETRIEVE CALCULATED DUTY FROM FEED
C-  PREHEAT AND SET SPECIFICATION FOR BOTTOMS EXCHANGER
C-
F-     PRE
F-     FPREHT = GET(HEAEXC(2),CAL DUT)
F-     SET (HEAEXC(4),DUT) = -FPREHT
C-
       DELta pressure=5
C-
GENeral,
       COMponents = 46, 49, 2, 3, 4, 5, 6, 7, 8, 10, 11, 12, 13, 14, 15
       FLOw 1 = 2.05, 30.52, 8.55, 17.7, 48.75, 15.015, 64.35, 21.54, 26.4, 32.97, 20.11, 6.81, 1.44,
       1.905, 11.7
       TP 1 = 85, 310
       APISOAVEK, APISOAVEH, STDD REID vapor pressure = 5
END
```

NOTE: The EPS(1) statement allows the user to define his own distillation column specifications, in this case, a bottom product Reid vapor pressure of 10. This specification replaces the original PURity of BOTtom for methane (see ***Inline Fortran Table 3. DIStillation Function*** for more details). At least one EPS specification **must** be defined in DIStillation for DURing operations to be performed.

Inlfrn3.psd - Railroad Tank Car Volume Calculation

In this problem, an olefin stream at 25 F and 75 PSIA is fed to a tank car. Initially the stream contains 10% by volume of vapor. Along its trip, the tank car is warmed by the sun to 115 F. The objective of this example is to use Inline Fortran to determine the new tank car pressure, assuming no venting.



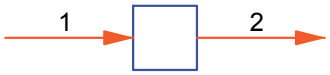

```
AB123.
*RAILROAD TANK CAR
C-
STReam manipulator 1 = VOLFLO,1,-2,
       FACtor = 1, 1
C-
C-  CALC ACTUAL VOLUMES FOR VAPOR AND LIQUID STREAMS OUT OF
C-  FLASH 2, CALC FACTOR FOR STREAM MANIPULATOR 3 TO ADJUST
C-  VAPOR FLOW SO THAT IT IS 10 VOLUME % OF THE MIXTURE.
C-
FLAsh 2 = VOL,2,-3,-4
F-     POST
```

```
F-      INCLUDE STRCOM
F-      DIMENSION FLO(5)
F-      CALL STRCOM(3,FLO,5)
F-      T = GET(STR(2),TEM)
F-      P = GET(STR(2),PRE)
F-      CALL DENSTY(FLO(1),T,P,0,D)
F-      VAPD = D
F-      VAP = GET(STR(3),FLO)
F-      VAPVOL = VAP/VAPD
F-      CALL STRCOM(4,FLO,5)
F-      CALL DENSTY(FLO(1),T,P,1,D)
F-      DLIQ = D
F-      FLOLIQ = GET(STR(4),FLO)
F-      VLIQ = FLOLIQ/DLIQ
F-      NEWVAP = VLIQ/9
F-      VRATIO = NEWVAP/VAPVOL
F-      WRITE(6,500)VAPVOL,VLIQ,VRATIO
F- 500  FORMAT(1X,8HVAPVOL=,F15.5,/1X,8HVLIQ=,F15.5
     *  /1X,8HVRATIO=,F15.5)
F-      SET(STRMAN(3),FAC(1)) = VRATIO
F-      SET(STRMAN(1),FAC(1)) = VLIQ + (VAPVOL*VRATIO)
STReam 3 = VAPVOL,3,-100,-5,
        FACtor = 1, SAMe FACtor, MULtiply
MIXer 4 = TANK, 4, 5, -6
VALve 5 = PRESSURE, 6, -7,
        PREssure OUT = 800
HEAt EXChanger 6 = SUN, 7, -8,
        TEMperature OUT = 115
C-
C-  CALCULATE ACTUAL VOLUME AT 115 F AND NEW PRESSURE
C-
FLAsh 7 = NEWVOL, 8, -9, -10
F-     POST
F-      INCLUDE STRCOM
F-      DIMENSION FLO(5)
F-      CALL STRCOM(9,FLO,5)
F-      T = GET(STR(8),TEM)
F-      P = GET(STR(8),PRE)
F-      CALL DENSTY(FLO(1),T,P,0,D)
F-      VAPD = D
F-      VAP = GET(STR(9),FLO)
F-      VAPVOL = VAP/VAPD
F-      CALL STRCOM(10,FLO,5)
F-      CALL DENSTY(FLO(1),T,P,1,D)
F-      DLIQ = D
F-      FLOLIQ = GET(STR(10),FLO)
F-      VLIQ = FLOLIQ/DLIQ
F-      WRITE(6,501)VAPVOL,VLIQ
F-501   FORMAT(1X,8HVAPVOL=,F15.5/1X,8HVLIQ=,F15.5)
F-      SET(STRMAN(1),FAC(2)) = VLIQ + VAPVOL
C-
C-USE CONTROLLER TO COMPARE THE ACTUAL VOLUMES CALCULATED AT
C-25F AND 115F, ADJUST PRESSURE IN VALVE 5 UNTIL THE VOLUMES
C-ARE EQUAL.
C-
CONTroller 8 = SETP,9,-11,
        VARy PRE OUT(MIN=75,MAX=1200,STE=40)OF VAL5
        UNTil FAC2 FROm STR1 = FAC1 FROm STR1
C-
GENeral,
        COMponents = 46,4,23,28,24
        FLOw (LB/HR) 1 = 750,5000,7500,10000,1585
        TP 1 = 25,75
        PENK,PENH,BIN PAR = PENG1
C-
C-FORTRAN LIBRARY IS WRITTEN TO GET THE VECTOR OF COMPONENT
C-FLOWS FOR SPECIFIED STREAM #. DESIGN II DENSTY ROUTINE
C-USED ABOVE REQUIRES THIS INFORMATION.
C-
FORTRAN
        STArt LIBrary
               SUBROUTINE STRCOM(NSTR,FLO,NCP)
               DIMENSION FLO(1)
               DO 100 I = 1,NCP
               FLO(I) = GET(STR(NSTR),FLO COM(I))
    100        CONTINUE
               RETURN
```

```
            END
        STOp LIBrary
END
```

Inlfrn4.psd - Air-cooled Exchanger ADD Block

This is a user written equipment module, or ADD block, of a simple air cooled exchanger. Use the EQPar keyword to specify the outlet temperature of the process stream, the ambient air temperature, and the extended surface heat transfer coefficient. The module performs a simple heat and mass balance to calculate the outlet air's temperature, flow (lb/hr), and actual cubic feet per minute.



AIRCOOLER

```
AB123.
*SIMPLE AIR COOLED EXCHANGER CALCULATION WITH INLINE FORTRAN
ADD 8 =AIRCOOLER,1,-2
USE BLOCK 3,
NEQP =10
EQP = 8, 0, 150, 100, 4.2
GENeral
        COMponents = 100
        AMB 100 = 350, API 100 = 37, AMW 100 = 175,
        NAMe 100 = PETFRAC
        FLO (LB/HR) 1 = 273000
        TP 1 = 250, 50
FORTRAN
BLOck 3 = 1,AIRCLR
STArt LIBrary
        SUBROUTINE AIRCLR ( NIN,  NOUT,  NCP,  NEQP,  NDSP,
    &   SIVPFR, SITEMP, SIPRES, SIENTH, SIMOLE, SICOMP, SIKV,
    &   SOVPFR,  SOTEMP, SOPRES, SOENTH, SOMOLE, SOCOMP,
    &   SOKV,    EQPAR,  DESPAR, AMW,    ATB,    IDCOMP  ,
    &   ISTOP,  KTRACE, NCP)
C--
        DIMENSION SIVPFR(NIN),SITEMP(NIN),SIPRES(NIN),
    &   SIENTH(NIN),SIMOLE(NIN),SICOMP(NCP,NIN),SIKV(NCP,NIN)

        DIMENSION SOVPFR(NOUT),SOTEMP(NOUT),SOPRES(NOUT),
    &   SOENTH(NOUT),SOMOLE(NOUT),SOCOMP(NCP,NOUT),
    &   SOKV(NCP,NOUT)

        DIMENSION EQPAR(NEQP),DESPAR(NDSP),AMW(NCP),ATB(NCP),
    &   IDCOMP(NCP)

        DIMENSION DECANT(2)
C--
C---------------------------------------
C--
C--   THIS IS A SIMPLE AIR COOLED EXCHANGER ADAPTED FROM
C--   THE GPSA'S ENGINEERING DATA BOOK.
C--
C-ONE INLET STREAM AND 1, 2, OR 3 OUTLET STREAMS ARE ALLOWED
C-IF 2 OR 3 OUTLET STREAM CODED, PHASES SEPARATED AS IN THE
C-VALVE MODULE.
C--
C--   INPUT PARAMETERS
C--   ----------------
C-   EQPAR(1) =NE  ...EQUIPMENT NUMBER IN FLOWSHEET
C-   EQPAR(2) =DUMMY SET TO ZERO, USED FOR ISOTHERMAL FLASH
C-   EQPAR(3) =TOUT  TEM OUT SPEC ON PROCESS STREAM (F)
C-   EQPAR(4) =TINAIR  TEMP OF AIR (F), DEFAULT IS 90 F
C-   EQPAR(5) =UX      EXTENDED SURFACE HEAT TRANSFER COEFFICIENT
C-           DEFAULT IS 4.0 BTU/HR*FT2*F
C--
C--    OUTPUT PARAMETERS
C--    -----------------
C--    EQPAR(6) =Q    DUTY (BTU/HR)
C--    EQPAR(7) =DELAIR        AIR TEMPERATURE CHANGE (F)
C--    EQPAR(8) =TOUTA TEM OUT OF AIR (F)
C--    EQPAR(9) =WTAIR WEIGHT OF AIR FLOW (LB/HR)
```

```
C--     EQPAR(10)=ACFM  ACTUAL FLOW OF AIR (FT3/MIN)
C--
C-------------------------------------------------------------
C--
C--     COLLECT INPUT DATA AND SET DEFAULTS
C--
        TIN = SITEMP(1) - 459.67
        TOUT = EQPAR(3)
        HIN = SIENTH(1)
        TINAIR = EQPAR(4)
        UX = EQPAR(5)

        IF(TINAIR.EQ.0.)TINAIR = 90.
        IF(UX.EQ.0.)UX = 4.
C--
C--ISOTHERMAL FLASH TO CALCULATE OUTLET STREAMS
C--
        EQPAR(2) = 1.
        SITEMP(1) = TOUT + 459.67
        CALL FLASH1 ( NIN,    NOUT,   NCP,    NEQP,   NDSP,
   &            SIVPFR,SITEMP,SIPRES,SIENTH,SIMOLE,SICOMP,SIKV,
   &            SOVPFR, SOTEMP, SOPRES, SOENTH, SOMOLE, SOCOMP,
   &            SOKV,   DECANT, EQPAR,  DESPAR)
C--
C-- CALCULATE DUTY, AIR TEMP OUT, AND AIR FLOW
C--
                HOUT = 0.
                DO 10 I = 1,NOUT
        10      HOUT = HOUT + SOENTH(I)
                Q = HOUT - HIN
                F1 = (UX+1.)/10.
                F2 = ((TIN+TOUT)/2)-TINAIR
                DELAIR = F1*F2
                TOUTA = TINAIR+DELAIR
                WTAIR = ABS(Q)/(0.24*DELAIR)
                ACFM = 0.222*WTAIR/0.94
C--
C-- PUT RESULTS IN EQPAR VECTOR
C--
                EQPAR(3) = TOUT
                EQPAR(4) = TINAIR
                EQPAR(5) = UX
                EQPAR(6) = Q
                EQPAR(7) = DELAIR
                EQPAR(8) = TOUTA
                EQPAR(9) = WTAIR
                EQPAR(10) = ACFM

        RETURN
        END
STOp LIBrary
END
```

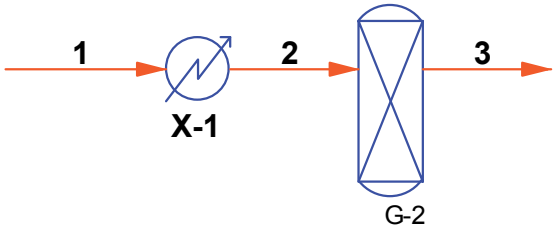
A portion of the output file is shown below.

```
        STREAM SUMMARY
        --------------
STREAM NUMBER              1               2
EQUIP CONXION           FEED-AIRC        AIRC-PROD
                       (  0)-(  8)      (  8)-(  0)
VAPOR FRACTION           0.00000          0.00000
TEMPERATURE F             250.00           150.00
PRESSURE PSIA             50.000           50.000
ENTHALPY BTU/HR       -0.47369E 07     -0.17918E 08
LB/FT3 T-P                49.007           50.979
S.G.(60F) STP            0.83965          0.83965
GAL/MIN STP               649.97           649.97
MMSCF/DAY STP
MOLECULAR WT              175.00           175.00
FLOW RATES LBMOL/HR
PETFRAC 1560.0            1560.0
        -----------     -----------
TOTAL LBMOL/HR            1560.0           1560.0
 TOTAL LB/HR           0.27300E 06      0.27300E 06
+ EQUIPMENT SUMMARY - INDIVIDUAL DETAILS +

 ****ADD MODULE               ****
```

```
 +++BLO3+++

 EQUIPMENT NO.                          8

EXTERNAL NAME              AIRC
PARAMETERS :               1.000
       3                   150.0
       4                   100.0
       5                   4.200
       6                -0.1318E 08
       7                   52.00
       8                   152.0
       9                 0.1716E 07
                        10    0.4053E 06
```

Inlfrn5.psd - Print Extra Flowsheet Information

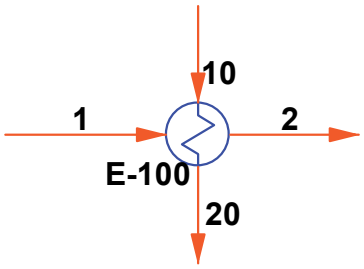
Inline FORTRAN can write calculated information to the output file or DISPLAY information on the screen in interactive versions. In this steam reformer example, Inline FORTRAN GETs product component flowrates, then calculates and prints the H2 to CO ratio.




```
AB123.
*REFORMER MODEL
HEAt EXChanger 1 = PREHEAT, 1, -2
        TEMperature OUT(c) = 483.8
REActor 2 = REFORMER, 2, -3
        STEam REFormer
        TEMperature of APProach METHANation = -30
        TEMperature of APProach COSHIFt = -20
        TEMperature OUT (C) = 808.5
        NATional data

C- USE INLINE FORTRAN TO CALCULATE METHANE
C- IN DRY REFORMED GAS MOLE %
C- AND H2/CO RATIO

F-              POST
F-              TOTAL = GET(STR(3),FLO)
F-              DRY = TOTAL - GET(STR(3),FLO COM(9))
F-              CH4 = GET(STR(3),FLO COM(2))
F-              CH4DRY = (CH4/DRY)*100.
F-              H2    = GET(STR(3),FLO COM(1))
F-              CO    = GET(STR(3),FLO COM(6))
F-              RATIO = H2/CO
F-              WRITE(6,99) CH4DRY, RATIO
F-     99       FORMAT(//,1X,'DRY MOLE PERCENT CH4 = ',F8.4,
F-     $        /,1X,'H2/CO MOL RATIO     = ',F8.4,//)

GENeral
        COMponents = 1,2,3,4,6,48,49,46,62
        PENK, PENH, PEND, BIN PAR = PENG1
        TP 1 = 360,150
        FLOw 1 = 300, 2000, 80, 20, 10, 20, 500, 20, 4000
END
```

Inlfrn6.psd - Adding Auxiliary Equipment Calculations

Inline Fortran can modify input equipment specifications or do stand-alone calculations for an equipment module. Tying a SIMple heat exchanger pressure drop to feed flow rates is a useful application. This example calculates a new heat exchanger pressure drop based on the old pressure drop and new and old flow rates. Next, it resets the original DELta pressure specification with the new value.



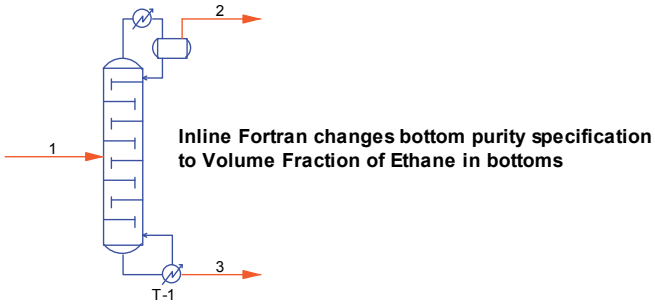

```
AB123.
*HEAT EXCHANGER CALCULATIONS WITH ADJUSTED PRESSURE DROP
HEAt EXChanger 1 = E-100, 1, 10, -2, -20,
        SIMple,
        U = 50, AREa = 2000,
        DELta pressure = 5, 15
F-              PRE
F-              BSFLO = 500
F-              BSDP = 5
F-              DPNEW = (GET(STR(1),FLO)/BSFLO)**2 * BSDP
F-              SET(HEA EXC(1),DEL(1)) = DPNEW
GENeral
        COMponents= 62, 5, 6, 7, 8, 10
        FLOw 1= 0, 20, 40, 50, 80, 350
        TP 1= 300, 100
        FLOw 10= 13500, 5*0
        TP 10= 80, 100
        STDK, STDH, STDD
END
```

More complex calculations can also be added to an existing equipment module. This example uses the HEAt EXChanger's routines to change the temperature of a stream while Inline FORTRAN performs a user-written set of air cooler calculations.

```
AB123.
*SIMPLE AIR COOLED EXCHANGER CALCULATION WITH INLINE FORTRAN
HEAt EXChanger 1 = AIRCOOLER, 1,-2,
        TEMperature OUT = 150,
C- ADD FORTRAN TO HEAT EXCHANGER FOR AIR FLOW PREDICTION.
C- CALCS ADAPTED FROM GPSA'S ENGINEERING DATA BOOK.
F-              POST
F-              TINAIR = 100.
F-              UX = 4.2
F-              Q = -GET(HEAEXC(1),CAL DUTY)
F-              TIN = GET(STR(1),TEM)-459.67
F-              TOUT = GET(STR(2),TEM)-459.67
F-              DELAIR = (UX+1.)/10. * (((TIN+TOUT)/2.)-TINAIR)
F-              TOUTA = TINAIR + DELAIR
F-              WTAIR = ABS(Q)/(0.24*DELAIR)
F-              ACFM = 0.222*WTAIR/0.94
C-
F-              WRITE(6,100) UX
F-              WRITE(6,101) TINAIR,TOUTA
F-              WRITE(6,102) Q,WTAIR
F-              WRITE(6,103) ACFM
F-   100 FORMAT(///,1X,'** AIR COOLER RESULTS **',//,
F-      &       1X,'U FINNED (BTU/HR*FT2*F)=',G15.5)
F-   101 FORMAT( 1X,'AIR IN    (F)          =',G15.5,/,
F-      &              1X,'AIR OUT   (F)             =',G15.5)
F-   102 FORMAT(   1X,'DUTY     (BTU/HR)       =',G15.5,/,
F-      &              1X,'AIR FLOW (LB/HR)          =',G15.5)
F-   103 FORMAT(1X,'AIR FLOW(ACTUAL FT3/MIN)=',G15.5,///)

GENeral
        COMponents = 100
        AMB 100 = 350
        API 100 = 37
        AMW 100 = 175
        NAMe 100 = PETFRAC
        FLOw (LB/HR) 1 = 273000
        TP 1 = 250, 50
END
```

Inlfrn7.psd - Creating Custom Distillation Specifications

Inline Fortran can replace "hard-coded" DIStillation column specs with anything the user calculates. Product true or Reid vapor pressure, purity of a range of components, or physical properties are just a few of the possible specifications that can be made with Inline Fortran. This example sets the liquid volume percent of ethane in the column bottoms.

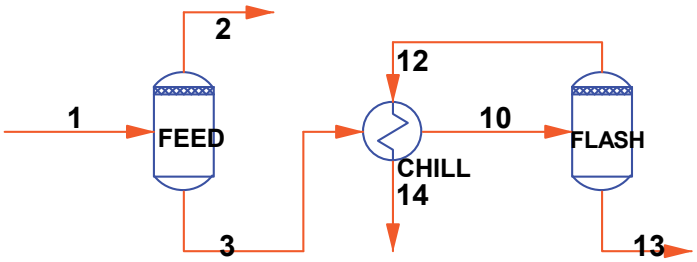



```
AB123.
*DEETHANIZER
DIStillation 1 = DEC2, 1,-2,-3
        PARtial, TRAy = 20, LOCation of FEEd = 10
        PREssure TOP = 345, DELta pressure = 5,
        PREssure of CONdenser = 340
        TEMperature of TOP guess = 30,
        TEMperature of  BOTtom guess = 200
        PROduct guess = 150, 0
        REFlux ratio= 3
        PURity BOTtom 3 = .002

C-  FORTRAN BELOW WILL REPLACE THE PURITY SPEC ABOVE
C-  WITH THE "EPS(1)" SPEC SETTING ETHANE'S LIQUID
C-  VOLUME FRACTION IN THE BOTTOM PRODUCT = .002
F-      DURING
F-      VOLTOT = 0.
F-      DO 100 I = 1,9
F-      VOLTOT = VOLTOT + (FLOWL(I,1)*GET(BANK(GPMO),POS(I)))
F- 100 CONTINUE
F-      VOLFRA  = FLOWL(3,1)*GET(BANK(GPMO),POS(3))/VOLTOT
F-      PRINT VOLFRA
F-      EPS(1) = (VOLFRA-0.002)/0.002

GENeral
        COMponents = 49,2,3,4,5,6,7,8,10
        TP 1 = 121.77, 350
        FLOw(LBMOL/HR)1=0.2,3.2,156.7,113.8,38.9,46.4,
                20.0,16.0,30.0
END
```

Inlfrn8.psd - Referencing Internally Calculated Streams

The example below shows the use of Inline Fortran's COPSTR and INISTR commands to guess a recycle stream composition. COPSTR copies stream 3 to stream 10 and INISTR initializes stream 10 under its new conditions. The operation is performed only for the first iteration of the recycle loop calculations.

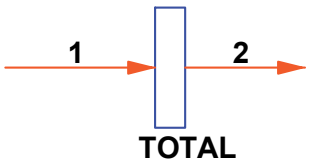
Sometimes a user needs to make specifications based on a stream whose properties are unknown at the start of the run. Inline FORTRAN can access that information and send it where it needs to go. Inline FORTRAN can even make recycle guesses based on internally-calculated streams. The example below shows the use of Inline FORTRAN's COPSTR and INISTR commands to guess a recycle stream composition. COPSTR copies stream 3 to stream 10 and INISTR initializes stream 10 under its new conditions. The operation is performed only for the first iteration of the recycle loop calculations.



```
AB123.
*INTERNAL VARIABLES SAMPLE PROBLEM
FLAsh 1 = FEED, 1, -2, -3
F-      POST
F-      COMMON ITER
F-      ITER = ITER + 1.
F-      IF (ITER .GT. 1) GO TO 100
F-      CALL COPSTR(3,10)
F-      SET(STR(10),TEM)= 555
F-      CALL INISTR(10)
F- 100 CONTINUE

HEAt EXChanger 2 = CHILL, 3, 12, -10, -14,
        SIMple, U = 50, AREa = 400

VALve 3 = FLASH, 10, -12, -13, PREssure OUT = 100

GENeral,
        COMponents = 2,3,4,5,6,7,8
        FLOw 1 = 2000, 1000, 500, 400, 300, 200, 100
        TP 1 = 100, 750
        SOAVEK, SOAVEH, STDD
        MAX = 10
        RECycle STReam = 10
END
```

Inlfrn9.psd - Point Properties and Name Retrieval

Flowsheet showing Inline Fortran code to retrieve database component properties such as name, molecular weight, critical properties, (Tc, Pc & Vc), acentric factor, boiling point, solubility, and characteristic volume.

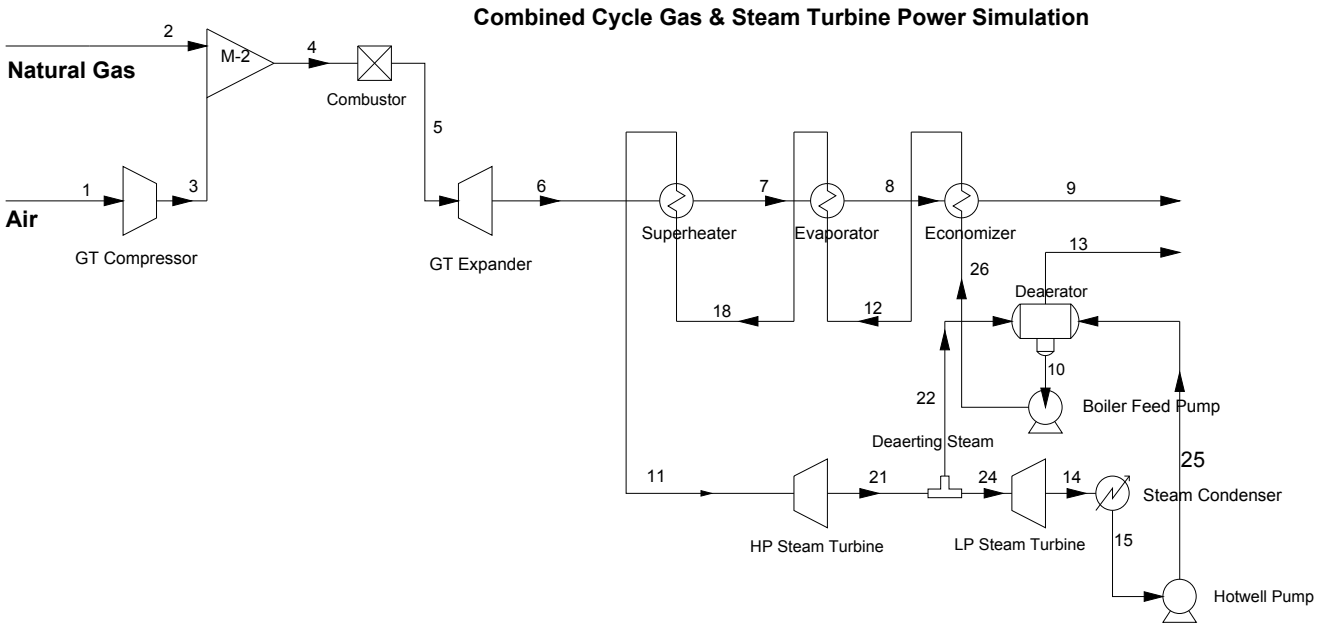



```
AB123.
MIX 1 = TOTAL,1, -2
F-              PRE
C-------------------------------------------------------------
C-              RETRIEVE THE NAMES OF THE 5 COMPONENTS
C-------------------------------------------------------------
F-              DIMENSION INAME(4)
F-              DO 120 L = 1,5
F-              DO 130 M = 1,4
F-                      N= (L-1)*4+M
F-                      INAME(M) =GET(IBANK(KOMN),POS(N))
F- 130  CONTINUE
F-              WRITE (6,40) (INAME(M),M=1,4)
C-   ------------------------------------------------------
C-                      RETRIEVE THE PROPERTIES FOR EACH COMPONENT
C-   ------------------------------------------------------
F-              AMW = GET(BANK (AMW),POS(L))
F-              ATB = GET (BANK (ATB),POS (L))
F-              ATC = GET (BANK(ATC),POS(L))
F-              APC = GET(BANK(APC),POS(L))
F-              AVC = GET(BANK(AVC),POS(L))
F-              ADEL = GET(BANK(ADEL),POS(L))
F-              AOMA = GET(BANK(AOMA),POS(L))
F-              AVW = GET(BANK(AVW),POS(L))
F-              WRITE (6,100) AMW
F-              WRITE (6,101) ATB
F-              WRTIE (6,102) ATC
F-              WRITE (6,103) APC
F-              WRITE (6,104) AVC
F-              WRITE (6,105) ADEL
F-              WRITE (6,106) AOMA
F-              WRITE (6,107) AVW
F-  100  FORMAT ( 'MOLECULAR WT',G15.7)
F-  101  FORMAT ('TEMP BOILING',G15.7)
F-  102  FORMAT (' TEMP CRIT ', G15.7)
F-  103  FORMAT (' PRESS CRIT ',G15.7)
F-  104  FORMAT (' VOLUME CRIT ',G15.7)
```

```
F-  105  FORMAT (' SOLUBILITY ' , G15.7)
F-  106  FORMAT (' ACENTRIC FAC', G15.7)
F-  107  FORMAT ('CHAR. VOLUME', G15.7, /)
F-  120  CONTINUE
F-  40    FORMAT (3X,4A4)
GENERAL
       COMPONENT = 2,3,4,5,62
       FLOW 1 = 5*100
       TP 1 = 104,200
END
```

Chapter 56: Fuel Cell Samples - c:\designii\samples\fuelcell

comcycle.psd – Combined Cycle Gas & Steam Turbine Power



Sample simulation showing combustion of natural gas with air and the combined cycle gas & steam power generation with a 70 MW Gas Turbine, Heat Recovery Steam Generator and a 30 MW Steam Turbine.

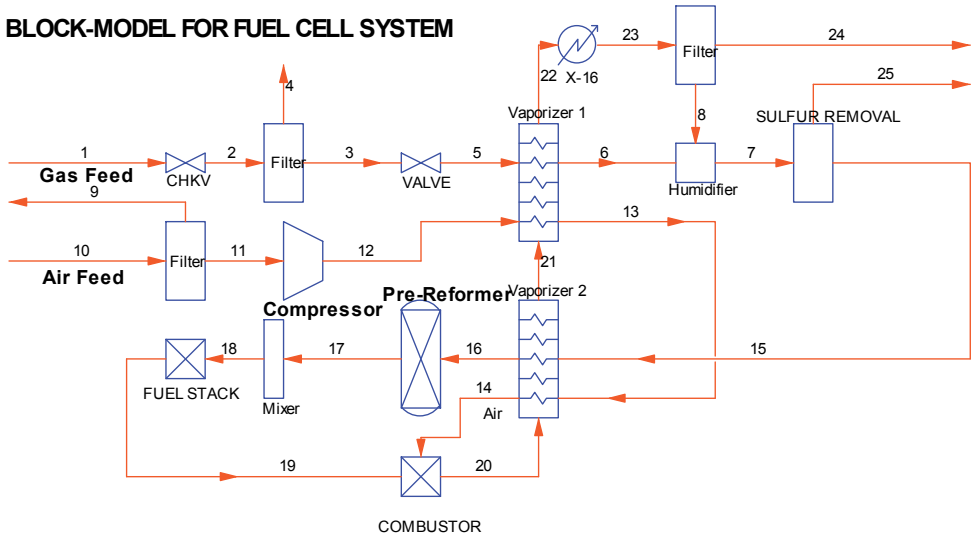
```
AB123.
*Combined Cycle Gas-Steam Power Generation

C- EQUIPMENT MODULES

COMPRE 1 = GT Compressor, 1,-3,
  PRE OUT(BAR) = 11.
  EFF = 0.92
  CAL POL
  ELE DRI
  STA = 1
  TEM INT(C) = 35.

MIX 2 = M-2, 2,3,-4,

EQU REA 3 = Combustor, 4,-5,
   ADIabatic

EXP 4 = GT Expander, 5,-6,
  PRE OUT(BAR) = 1.035
  EFF = 0.85

HEA EXC 5 = Superheater, 6,18,-7,-11,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 480.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 1.E-002, 1.8,

HEA EXC 6 = Evaporator, 7,12,-8,-18,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 248.6
```

```
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 1.E-002, 0.,

HEA EXC 7 = Economizer, 8,26,-9,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 248.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 2.E-003, 11.2,

VAL 8 = Deaerator, 22,25,-13,-10,
  DEL(BAR) = 0.

PUM 9 = Boiler Feed Pump, 10,-26,
  PRE OUT(BAR) = 50.
  EFF = 0.9
  ELE DRI

HEA EXC 10 = Steam Condenser, 14,-15,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(C) = 25.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.
  TEM WAT IN(F) = 60.
  TEM WAT OUT(F) = 80.

EXP 13 = HP Steam Turbine, 11,-21,
  PRE OUT(BAR) = 2.9
  EFF = 0.8

DIV 14 = Deaerting Steam, 21,-22,-24,
  FLO(KG/SEC) = 5.4

EXP 15 = LP Steam Turbine, 24,-14,
  PRE OUT(BAR) = 4.E-002
  EFF = 0.8

PUM 19 = Hotwell Pump, 15,-25,
  PRE OUT(BAR) = 10.
  EFF = 0.9
  ELE DRI

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,46,47,48,49,62,
  RKK
  STDH
  STDD
  SUR TEN = STD
  STE TAB
  QUA UNI OUT = KGMOL
  TEM UNI OUT = C
  PRE UNI OUT = BAR
  ENTH UNI OUT = KCAL
  TIM UNI OUT = SEC
  MAX = 50
  CON TOL = 1.E-005
  REC STR = 18,
  MET UNI
  VOL STA = NTP
  FLO (FRA) 1 = 2*0.,0.79,0.21,3*0.,
  TOT FLO (KG/SEC) 1 = 283.
  TP (C,BAR) 1 = 15.,1.013
```

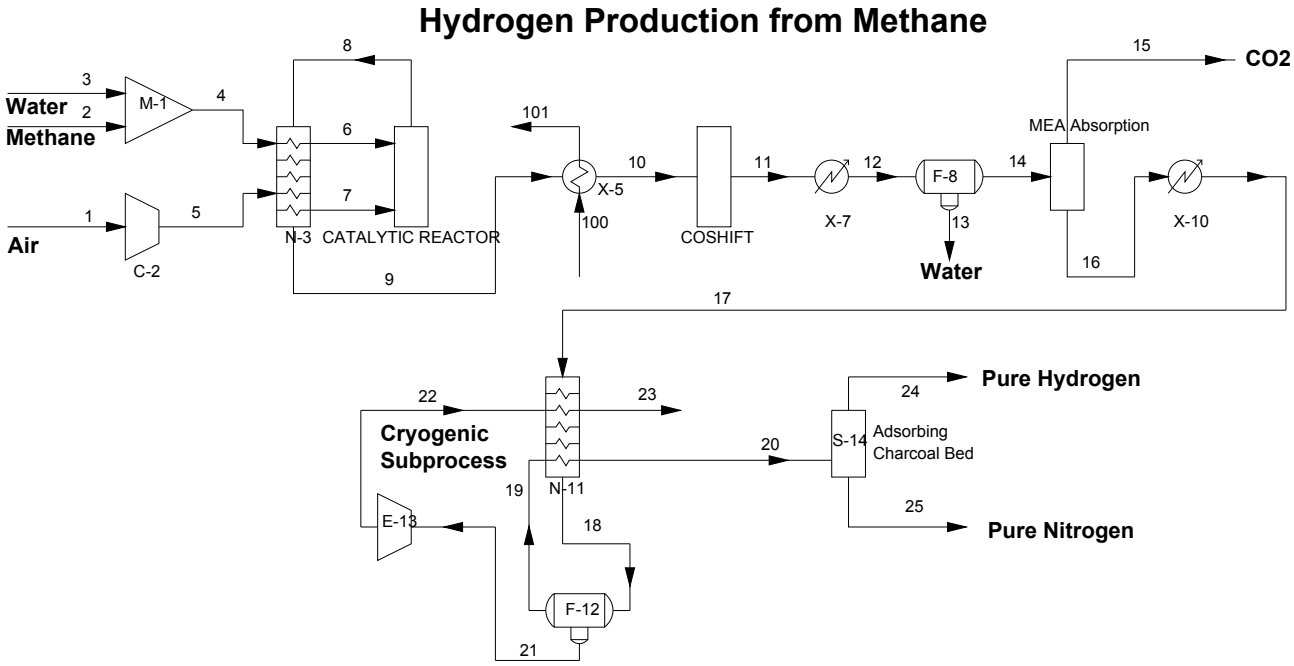
```
  NAM STR 1 = Air
  FLO(KG/SEC)2 = 0.,4.7,0.4,2*0.,0.4,0.,
  TP (C,BAR) 2 = 25.,11.
  NAM STR 2 = Natural Gas
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  FLO(KG/SEC)12 = 6*0.,34.9,
  TP (C,BAR) 12 = 200.,38.8
  VAP FRA 12 = 1.E-002, 1
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  FLO(KG/SEC)18 = 6*0.,34.9,
  TP (C,BAR) 18 = 250.,38.8
  VAP FRA 18 = 0.99, 1
  NAM STR 18 = Strm 18
  NAM STR 21 = Strm 21
  NAM STR 22 = Strm 22
  NAM STR 24 = Strm 24
  NAM STR 25 = Strm 25
  NAM STR 26 = Strm 26
END
```

fuelcell.psd - Simplified block diagram of a methane and air powered fuel cell

Illustrates the use of an equilibrium reactor to account for the production and translation of electrical energy to heat. The reactor redistributes all atomic species among the specified molecular species to achieve the most thermodynamically stable molecular compositions (via Gibbs Free Energy of minimization). All the necessary heat and reaction change is calculated from the pure component database properties and the product temperature rise will be very close to that seen in a purely electrical environment. The loss or gain of heat can be accounted via heat exchangers (as duty -BTUs lost or gained) and placing them in right locations within the flowsheet scheme that would balance the overall heat transfer.



H2plant.psd - Hydrogen Production from Methane



The above process is an illustration of hydrogen manufacture via a catalytic reactor. Methane, steam and air are each pre-heated to 600-700 C, and fed into a catalytic reactor containing a bed of refractory nickel catalyst (secondary reformer is used for simulation purposes). Initial combustion results in a temperature up to 1000 C. The product gas passes through heat exchange and a heat recovery boiler before entering the multi-stage CO shift converter. The gas then passes through a CO2 wash tower (can use MEA if needed in a separate flowsheet simulation...please see General: Gas Sweetening with MEA Example). Following CO2 absorption, the gas passes through a cryogenic separator where the hydrogen is separated from the CO, nitrogen, and methane.

The simulation described here produces 50 MMSCF/DAY of 99.93% pure hydrogen from methane and steam.

```
AB123.
*Hydrogen Production from Methane/Steam via Catalytic Reactor

C- EQUIPMENT MODULES

MIX 1 = M-1, 3,2,-4,

COMPRE 2 = C-2, 1,-5,
  PRE OUT(ATM) = 30.
  EFF = 0.72
  CAL POL
  ELE DRI
  STA = 3
  PRE STA(ATM) = 1.,10.,30.
  TEM INT(C) = 35.

LNG 3 = N-3, 4,5,8,-6,-7,-9,
 TEMperature OUT (F) 1 =1292
 TEMperature OUT (F) 2 =1292

REA 4 = CATALYTIC REACTOR, 6,7,-8,
  ADIabatic
  SECondary REFormer
  LIM=2
  CON=0.9

HEA EXC 5 = X-5, 9,100,-10,-101,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 327.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
```

```
  HOR
  DEL(PSI) = 0., 0.,

REA 6 = COSHIFT, 10,-11,
  COSHIFt reaction
  TEMperature APProach COSHIFt (F) = 5

HEA EXC 7 = X-7, 11,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 95.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.

FLA 8 = F-8, 12,-14,-13,
  ADI, HEA(BTU/HR) = 0.
  DEL(PSI) = 0.

COM SPL 9 = MEA Absorption, 14,-15,-16,
  REC TOP(FRA) = 5*0.,1.,0.,

HEA EXC 10 = X-10, 16,-17,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 95.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.

LNG 11 = N-11, 22,19,17,-23,-20,-18,
  TEMperature OUT of SHEll (F) =-323.73

FLA 12 = F-12, 18,-19,-21,
  ADI, HEA(BTU/HR) = 0.
  DEL(PSI) = 0.

EXP 13 = E-13, 21,-22,
  PRE OUT(ATM) = 1.
  EFF = 0.72

COM SPL 14 = S-14, 20,-24,-25,
  REC TOP(FRA) = 2*1.,0.,4*1.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,46,47,48,49,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  STE TAB
  MAX = 50
  CON TOL = 1.E-003
  FLO(LBMOL/DAY)1 = 2*0.,108509.586,28840.46,3*0.,
  TP (C,ATM) 1 = 35.,1.
  NAM STR 1 = Air
  FLO(LBMOL/DAY)2 = 0.,49043.39,5*0.,
  TP (C,ATM) 2 = 35.,30.
  NAM STR 2 = Methane
  FLO(LBMOL/DAY)3 = 6*0.,98086.78,
  TP (C,ATM) 3 = 235.,30.
  NAM STR 3 = Water
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  FLO(LBMOL/HR)11 = 5650.5,13.572,4521.2,0.,65.7,1964.,2496.2,
```

```
  TP (F,ATM) 11 = 512.82,30.
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 13 = Water
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
  NAM STR 17 = Strm 17
  NAM STR 18 = Strm 18
  FLO(LBMOL/HR)19 = 5501.5,4.92E-002,293.69,0.,3.8322,2*0.,
  TP (F,ATM) 19 = -323.,30.
  NAM STR 19 = Strm 19
  NAM STR 20 = Strm 20
  NAM STR 21 = Strm 21
  FLO(LBMOL/HR)22 = 140.,13.,4223.,62.,4.5E-002,26.7,0.12057,
  TP (F,ATM) 22 = -329.81,1.
  NAM STR 22 = Strm 22
  NAM STR 23 = Strm 23
  NAM STR 24 = Strm 24
  NAM STR 25 = Strm 25
  FLO(LBMOL/HR)100 = 6*0.,3874.,
  TP (F,ATM) 100 = 90.,1.
  NAM STR 100 = Strm 100
  NAM STR 101 = Strm 101
END
```

Chapter 57: Gas Processing Samples - c:\designii\samples\gasproc

Btextegd.dat – ChemTran Data File for TEG/BTEX Dehydration

It is a ChemTran created binary data file created by the ChemTran input file TEGBTEXCT.IN for use by DESIGN II at runtime of TEGBTEX.PSD flowsheet file below.

Deglwatr.dat – ChemTran Data File for Glycol Dehydration

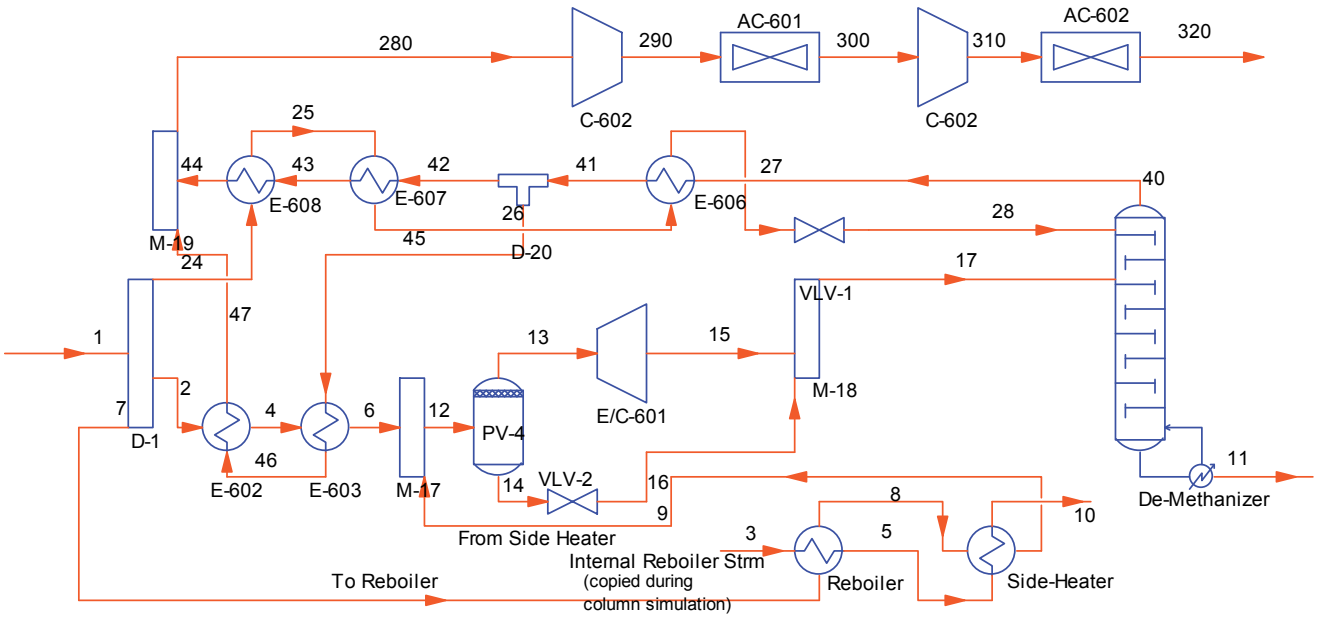
It is a ChemTran created binary data file created by the ChemTran input file DGMPNCT.IN for use by DESIGN II at runtime of GLYCOLDH.PSD flowsheet file below.

Dgmpnct.in – VLE Data Regression for Glycol Dehydration

It is a ChemTran VLE data regression input file for DEG and water mixture for GLYCOLDH.PSD file below. This is already built into the flowsheet file GLYCOLDH.PSD.

Expander Plant Model.psd – Typical Expander Plant Model

The simulation is a model of a typical expander process with a demethanizer.

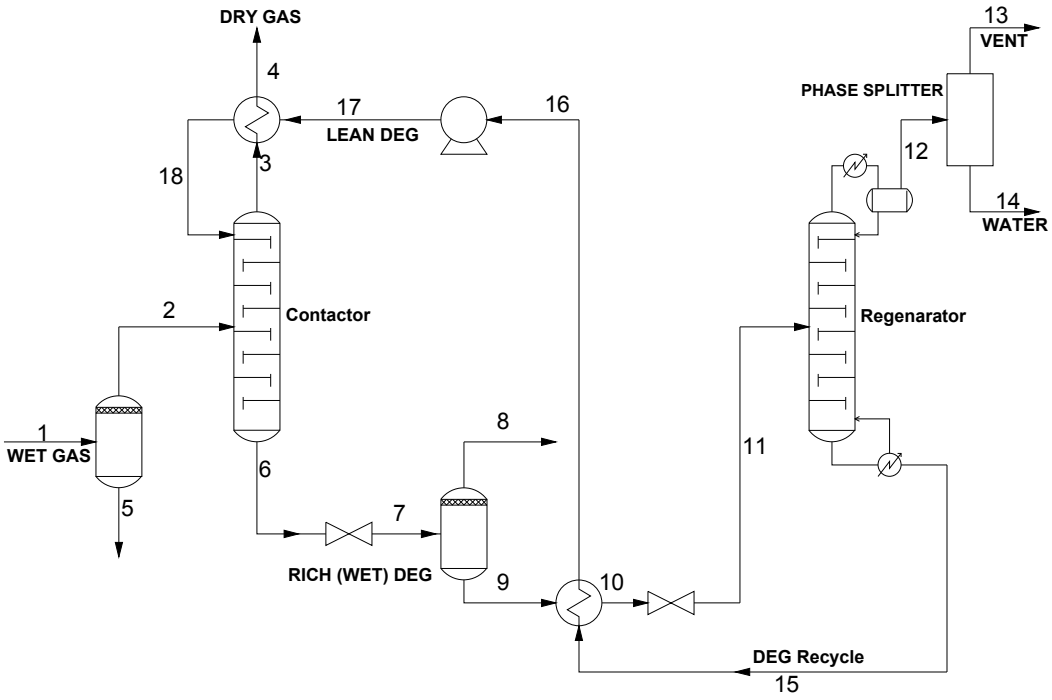


Glycwate.dat – ChemTran Data File for TEG / CO2 / Water Dehydration

It is a ChemTran created binary data file created by the ChemTran input file TGWCO2CT.IN for use by DESIGN II at runtime of TGCO2WT.PSD flowsheet file.

Glycoldh.psd - Glycol Dehydration

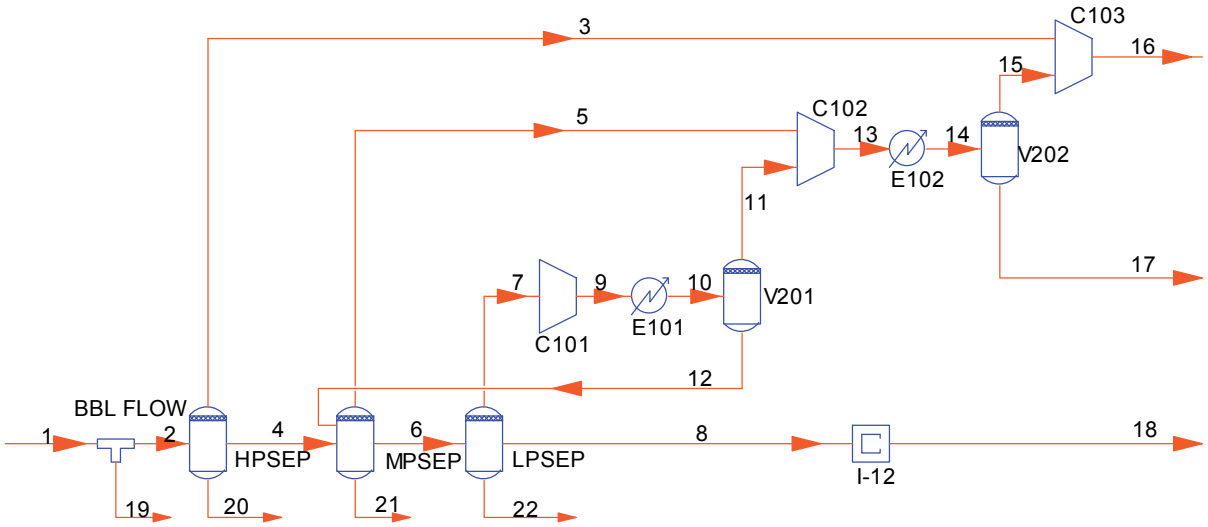
The sample problem is a simulation scheme for glycol/water dehydration system. The objective is to evaluate the amount of glycol recycled in order to dehydrate natural gas to sufficient purity often encountered in gas processing industry.



About 32450 lbmol/hr of wet natural gas is contacted counter-currently by diethylene glycol (stream 17) in the absorber module operating at 1250 psia. The wet diethylene glycol (rich in water) is passed through a valve and after flashing off trace hydrocarbons is fed to a glycol regenerator at about 14.7psia. The regenerator bottoms is the pure glycol recycled back to contact the wet gas again. A ChemTran regressed phase equilibrium data file is used for this flowsheet.

Gosp.psd - Gas-Oil Separation Plant

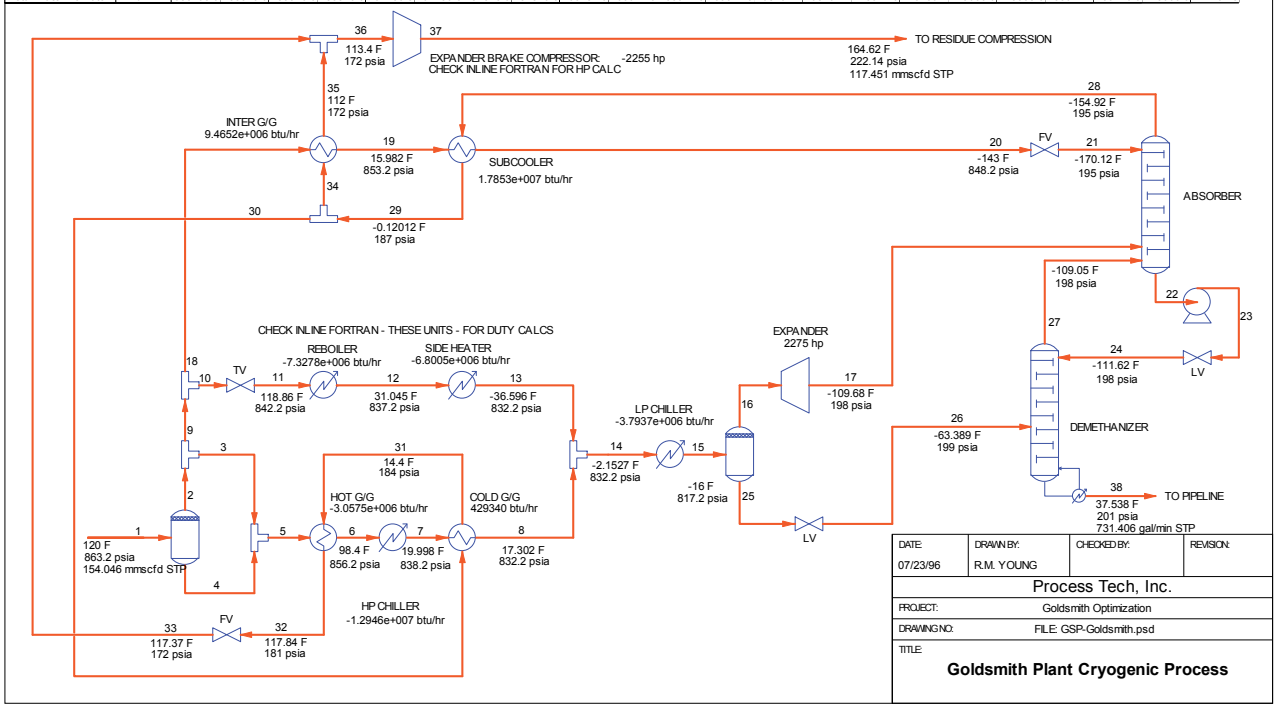
Illustration of a gas-oil separation plant with three phase flashes and Reid vapor pressure for the product. Petroleum fractions are also characterized in this problem.



GSP-Goldsmith.psd – GSP Process

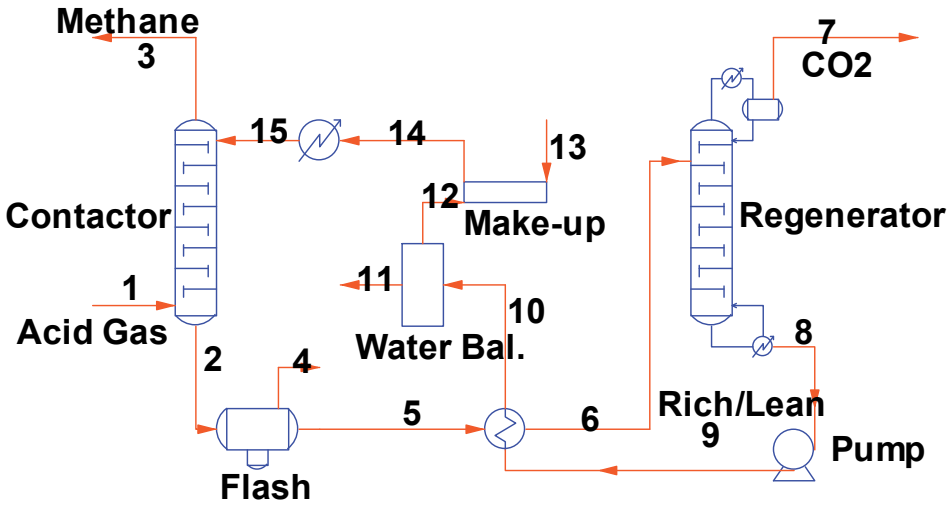
The simulation is a model of the Ortloff GSP Process used at the Goldsmith Plant.

Stream Number		1	6	7	8	12	13	14	15	17	19	20	21	22	24	26	27	28	29	31	35	37	38
Vapor Fraction		0.9981	0.9774	0.8282	0.8218	0.8639	0.6185	0.7633	0.7176	0.9196	0.8197	0	0.1776	0	0.0013	0.3855	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.002 9059E-006
Temperature	F	120.0	98.4	20.0	17.3	31.0	36.6	2.2	-16.0	-109.7	16.0	-143.0	-170.1	-112.0	-111.6	-63.4	-109.1	-154.9	-1.2E-001	14.4	112.0	164.6	37.5
Pressure	psia	863.2	856.2	838.2	832.2	837.2	832.2	832.2	817.2	198.0	853.2	848.2	195.0	198.0	198.0	199.0	198.0	195.0	187.0	184.0	172.0	222.1	201.0
Molecular Weight		22.82	22.89	22.89	22.89	22.76	22.76	22.84	22.84	18.73	22.76	22.76	22.76	32.30	32.30	33.29	17.35	16.68	16.68	16.68	16.68	16.68	42.40
Enthalpy	btu/hr	2433000	-1993900	-1494000	-15369000	-6665900	-1346600	-2883600	-3262900	-17257000	-8757600	-2661000	-2661000	-2448600	-2445500	-21162000	-355270	-22935000	-5081700	-892720	57024001	360200	-27936000
Vapor Density	lb/ft3	3.89	4.00	4.39	4.37	4.32	5.07	4.56	4.60	1.05	4.54	0	1.33	0	1.06	0.96	1.05	1.19	0.66	0.63	0.48	0.56	1.42
Liquid Density	lb/ft3	33.41	33.06	31.62	31.63	31.69	29.91	31.19	31.01	36.50	31.25	30.50	32.66	36.35	36.35	38.36	0	0	0	0	0	0	33.52
Vapor Flowrate (STP)	mmscf/day	154.05	72.16	61.15	60.67	33.61	24.06	86.06	80.91	74.40	34.10	0	7.39	0	3.49E-002	12.27	22.80	117.45	117.45	30.54	86.91	117.45	1.07E-004
Liquid Flowrate (STP)	gal/min	6.27	34.32	224.19	231.86	95.12	234.82	453.69	529.61	111.55	130.72	565.51	481.70	479.69	479.27	375.54	0	0	0	0	0	0	731.41
Water Flowrate (STP)		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Molar Flowrate	lbmol/hr	16948.4	8107.9	8107.9	8107.9	4272.6	4272.6	12380.5	12380.5	8884.8	4567.9	4567.9	4567.9	3058.6	3058.6	3495.7	2503.6	12897.8	12897.8	3353.4	9544.4	12897.8	4050.7
Total Mass Flowrate	lb/hr	386798.0	185576.0	185576.0	185576.0	97250.3	97250.3	282826.0	282826.0	166452.0	103971.0	103971.0	103971.0	98797.1	98797.1	116374.0	43439.4	215066.0	215066.0	55917.1	159149.0	215066.0	171732.0



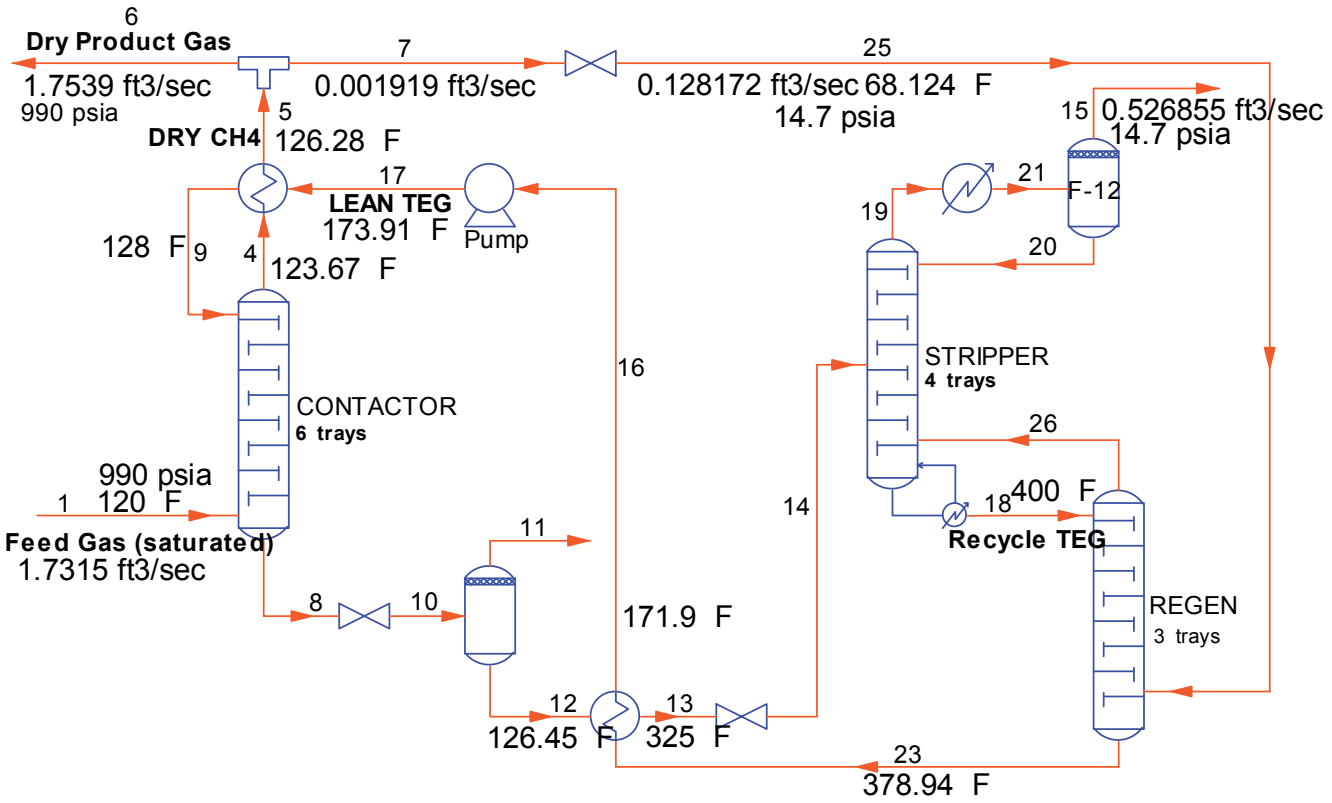
Meacdgs.psd - Gas Sweetening with MEA Example

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a monoethanolamine solution. Residual H2S in the sweet gas stream is to be no greater than 1/4 grains of H2S per 100 standard cubic feet (SCF) of gas. The reboiler and condenser duties will also be determined for the amine regenerator.



Teg.psd – Dehydration using TEG

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses a data file created via the regression of the water dewpoint of TEG / Water / Methane mixtures (reference: 5th edition of Gas Purification by Arthur Kohl) using ChemTran.

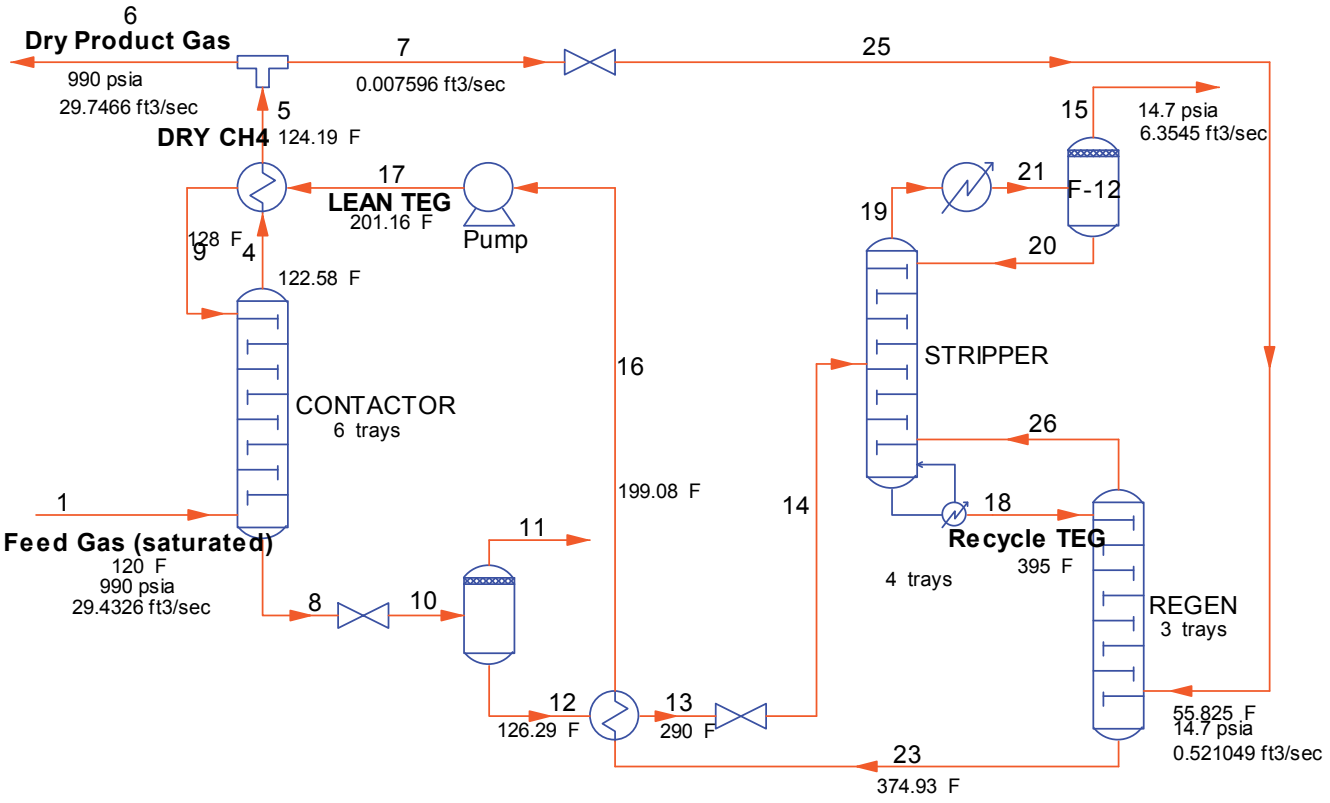


Teg_with ChemTran BIP.psd – VLE Data Regression of TEG/Water/Methane

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses the ChemTran BIPs via VLE data regression of TEG / Water / Methane mixture for Modified Peng-Robinson thermodynamics in DESIGN II.

Teg1.psd – Dehydration using TEG

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses the built in TEG / Water / Methane three phase Peng-Robinson mixture thermodynamics in DESIGN II.



Teg1watr.dat – ChemTran Data File for Dehydration using TEG

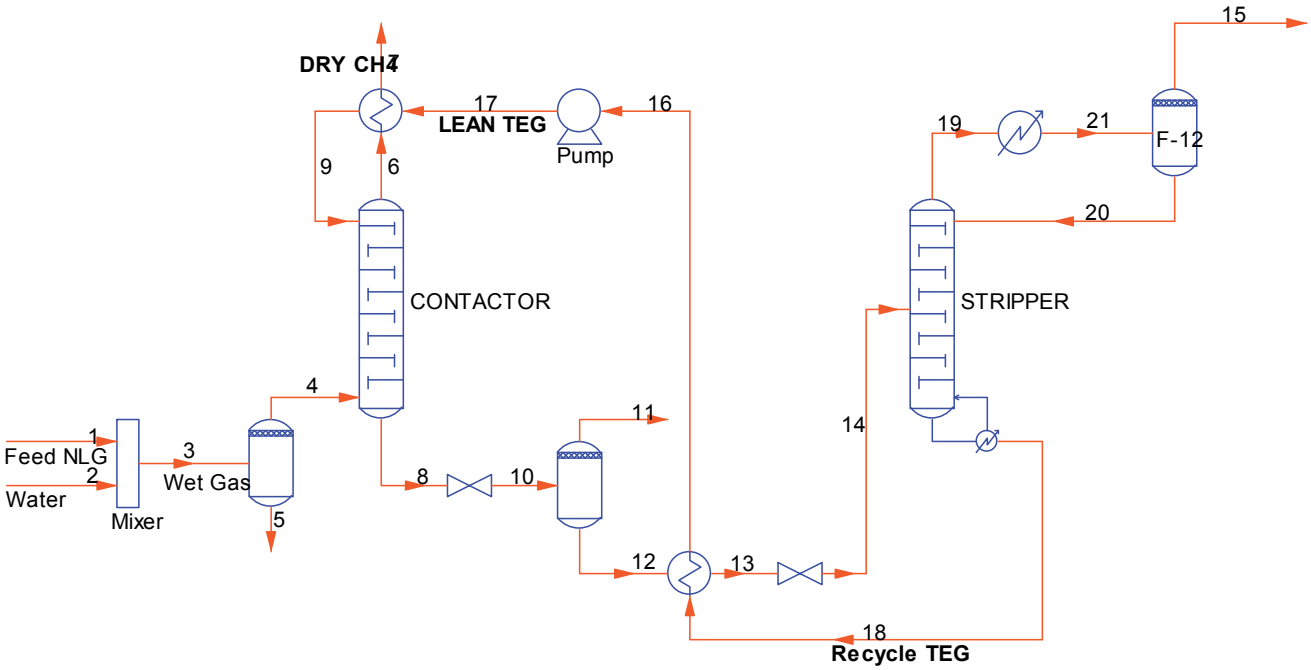
It is a ChemTran created binary data file created by the ChemTran input file TEGCHM.IN for use by DESIGN II at runtime of TEG.PSD flowsheet file.

Teg1_with inbuilt BIP.psd – Dehydration using TEG

Sample simulation [See Teg1.psd above.] illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses the built in TEG / Water / Methane three phase Modified Peng-Robinson mixture thermodynamics in DESIGN II.

Tegbtex.psd - TEG/BTEX Dehydration

The sample problem is a simulation scheme for tri-ethylene glycol/water dehydration for BTEX system. The objective is to evaluate the amount of TEG recycled in order to dehydrate a mixture of Benzene, Toluene, Ethyl benzene, o-Xylene, (BTEX) and methane gas to sufficient purity often encountered in gas processing industry.



Tegbtxct.in – VLE Data Regression for TEG/BTEX Dehydration

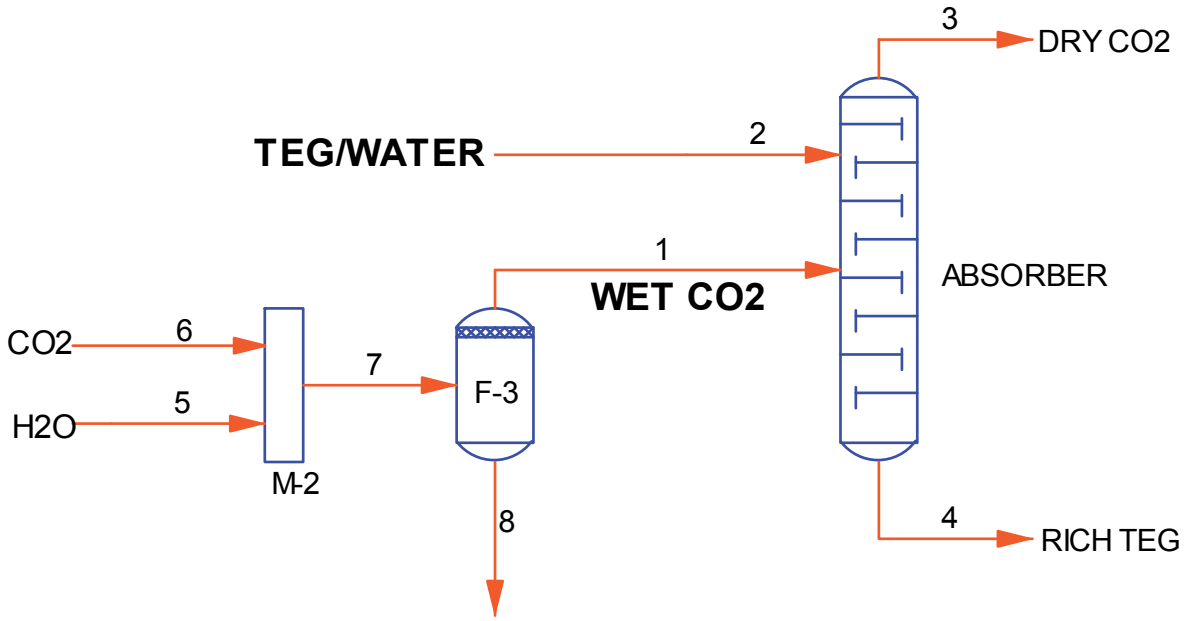
It is a ChemTran VLE data regression input file for TEG, BTEX and water mixture for TEGBTEX.PSD file. This is already built into the flowsheet file TEGBTEX.PSD.

Tegchm.in – VLE Data Regression for TEG and Water Mixture

It is a ChemTran VLE data regression input file for TEG and water mixture for TEG.PSD file. This is already built into the flowsheet file TEG.PSD.

Tgco2wt.psd – TEG / CO2 / Water Dehydration

The simulation is a simple contactor to dehydrate the wet CO2 / CH4 gas. TEG is used as a dehydrating solvent counter current to the flow of gas. A ChemTran regressed data file is used for the binary interactions of water/TEG.

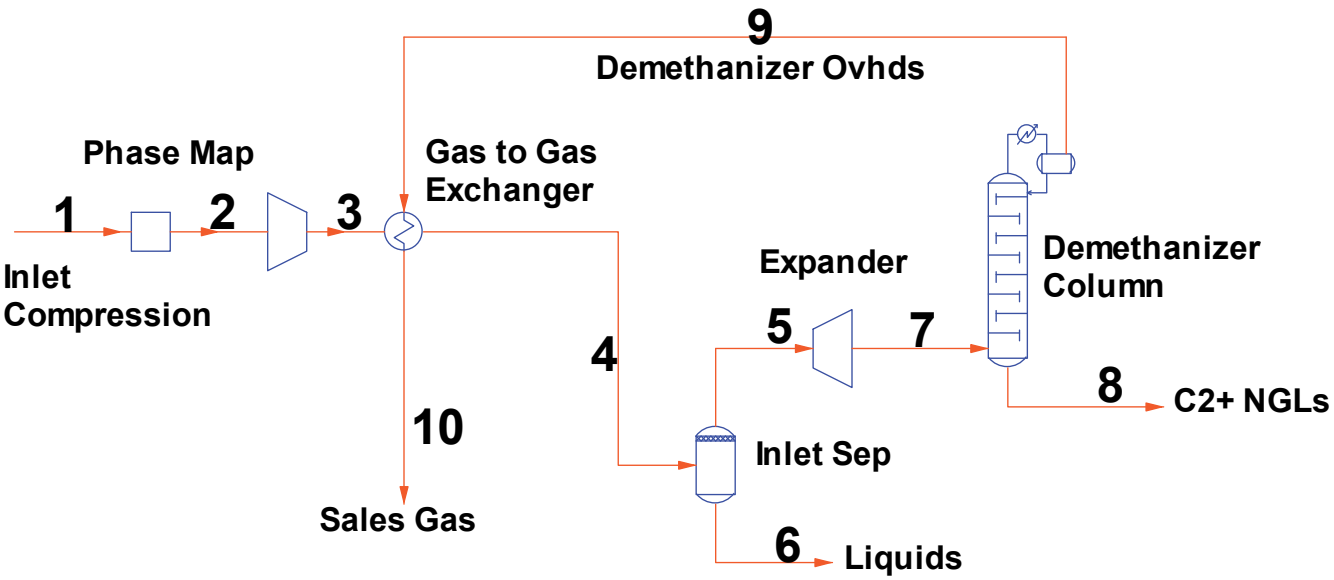


Tgwco2ct.in – VLE Data Regression for TEG

It is a ChemTran VLE data regression input file for TEG, and water mixture for TGCO2WT.PSD file. This is already built into the flowsheet file TGCO2WT.PSD.

Turbo.psd - Demethanizer For Sales Gas

A natural gas mixture (from C1-C5 & CO2, H2S) is used in a demethanizer expander plant process to separate methane from other NGL gas mixture. Pure methane is cross heat exchanged with the feed mixture as the product sales gas. The bottoms from the demethanizer are free of methane with C2 through C5.

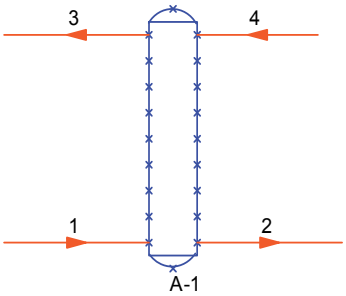


Chapter 57

Chapter 58: Gas Processing: Mixed Amine Samples - C:\designii\samples\gasproc\mixamine

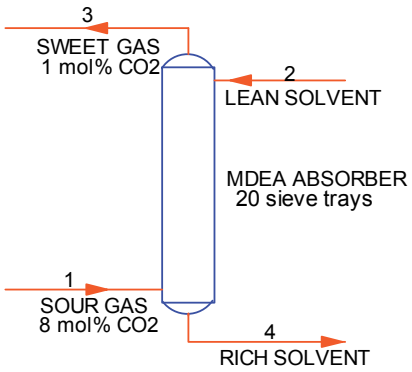
Co2_absorber.psd – Pure CO2 Absorber

Mixed Amine absorber for pure CO2 absorption. Please review other amine flowsheets for better examples.



Mxabs.psd - MEA/MDEA Absorber

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.



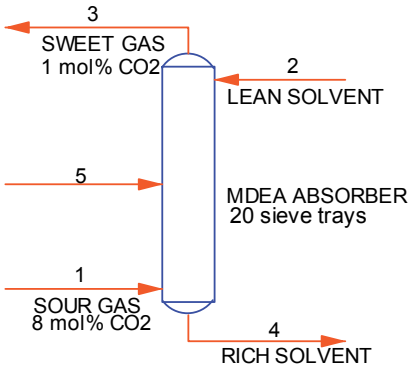
Assumptions
Use 39 weight % solution of MDEA.Amine (MEA 0.1 wt%).
Objectives
1 mol % of CO2 in the sweet gas, with top temperature of 130 F
Specification:
Tower diameter 70 inches. Active area 75%.
Tower internals - sieve trays, with 15% open area, single pass, 2 ft tray spacing.

```
AB123.
*CO2 BULK REMOVAL

C- EQUIPMENT MODULES
AMCL 1 = MDEA ABSORBER, 2,1,-3,-4,
  ABS1
  TRA = 20
  TRA TYP = 20*SIE,
  TRA PAS = 20*1,
  TOW DIA(IN) = 20*70.,
  TRA WEI(IN) = 20*2.5,
  TRA SPA(FT) = 20*2.,
  TRA ARE PER TOW = 20*75.,
  LOC FEE = 1,20
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
```

```
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004
  TRAy PERcent OPEn = 20*15

C- GENERAL SECTION

GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO(FRA)1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
  TOT FLO(MMSCF/DAY)1 = 60.
  TP (F,PSIA) 1 = 90.,900.
  NAM STR 1 = SOUR GAS
  NAM STR 3 = SWEET GAS
  FLO(FRA)2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO(GAL/MIN)2 = 500.
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 4 = RICH SOLVENT
END
```

Mxabsf.psd - Amine Absorber with Side Feed

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). A side feed stream is also processed in this absorber. Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.

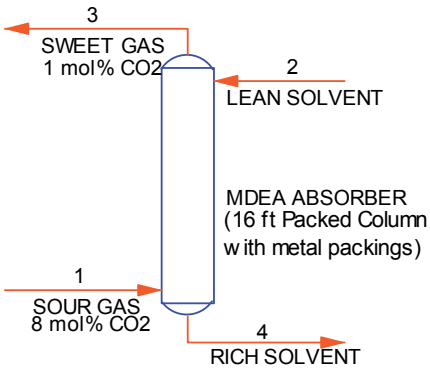



```
AB123.
*CO2 BULK REMOVAL -Side Feed

C- EQUIPMENT MODULES

AMCL 1 = MDEA ABSORBER, 2,5,1,-3,-4,
  ABS2
  TRA = 20
  TRA TYP = 20*SIE,
  TRA PAS = 20*1,
  TOW DIA(IN) = 20*70.,
  TRA WEI(IN) = 20*2.5,
  TRA SPA(FT) = 20*2.,
  TRA ARE PER TOW = 20*75.,
  LOC FEE = 1,10,20
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004
```

```
  TRAy PERcent OPEn = 20*15

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO(FRA)1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
  TOT FLO(MMSCF/DAY)1 = 60.
  TP (F,PSIA) 1 = 90.,900.
  NAM STR 1 = SOUR GAS
  NAM STR 3 = SWEET GAS
  FLO(FRA)2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO(GAL/MIN)2 = 500.
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 4 = RICH SOLVENT
  FLO(LBMOL/HR)5 = 3*1.,0.5,1.E-003,3*0.,
  TP (F,PSIG) 5 = 100.,880.
  NAM STR 5 = Strm 5
END
```

Mxpkabsf.psd - Amine Packed Column Absorber with Side Feed

The objective of this simulation is to remove CO2 (no H2S) from the natural gas stream by contacting the gas stream with a mixture of two amines (MEA and MDEA). A side feed stream is also processed in this packed column (2 ft ID, 16 ft in depth, 5/8 " metal packings, dry area 109 ft2/ft3, 75 % packing factor) absorber. Residual CO2 in the sweet gas stream is to be no greater than 1 mol %.

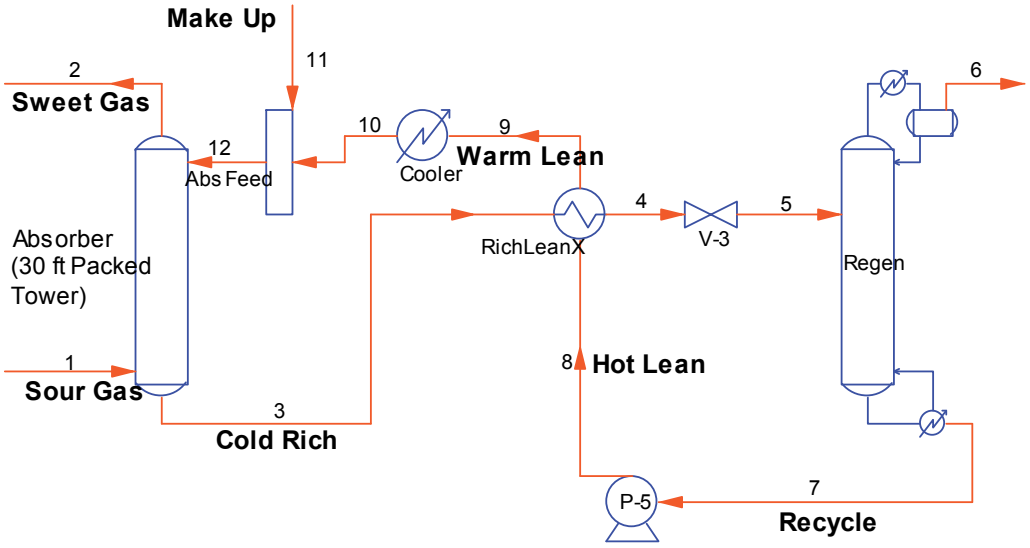



```
AB123.
*CO2 BULK REMOVAL -Packed Column with Side Feed

C- EQUIPMENT MODULES

AMCL 1 = MDEA ABSORBER, 2,5,1,-3,-4,
  ABS2
  PACK DEPTH PER TYPE(FT) = 2*8.,
  PACK MAT = 2*METAL,
  PACK TYP = 2*DUMPED,
  PACK SIZE(IN) = 2*0.625,
  PACK FAC = 2*75.,
  PACK DRY AREA(1/FT) = 2*109.,
  TOW DIA(IN) = 2*24.,
  REDIST DEPTH(FT) = 8.,
  LOC FEE (FT) = 1,10.,16.
  PRE TOP(PSIA) = 900.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 5500.,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004
```

```
C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,62,50,4155,7051,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  MAX = 10
  CON TOL = 1.E-003
  FLO (FRA) 1 = 0.9015,6.8E-003,4.65E-003,8.E-002,1.5E-004,3*0.,
  TOT FLO (MMSCF/DAY) 1 = 60.
  TP (F,PSIA) 1 = 98.,900.
  NAM STR 1 = SOUR GAS
  FLO (FRA) 2 = 3*0.,1.E-003,0.6,0.,1.E-003,0.4,
  TOT FLO (GAL/MIN) 2 = 400.
  TP (F,PSIA) 2 = 120.,900.
  NAM STR 2 = LEAN SOLVENT
  NAM STR 3 = SWEET GAS
  NAM STR 4 = RICH SOLVENT
  FLO(LBMOL/HR)5 = 3*1.,0.5,1.E-003,3*0.,
  TP (F,PSIA) 5 = 100.,900.
  NAM STR 5 = Strm 5
END
```

Mxpkplnt.psd - Amine Packed Column Absorber with Regenerator

The objective of this simulation is to contact CO2 (7 mol%) and H2S (2 mol %) in the natural gas stream with an aqueous stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .015 mol % of H2S and 0.50 mole % of CO2 per 25 MMSCFD of gas processed.



Use a packed column absorber with 30 ft of packing, 75 % tower flooding, 2/3 " metal packings, dry area 105 ft2/ft3, 70 % packing factor and a redistributor at mid point of the two packing sections

Assumptions

Use 10 weight % solution of DEA blended with 39 wt % of MDEA to make about 240 gal/min of aqueous recirculating amine. Make-up water and the amine losses are estimated and made-up automatically by the Mass Balance unit module.

Objectives

To achieve 0.50 mol % of CO2 in the sweet gas, with top temperature of about 115 F

Coding

```
AB123.
*Mixed Amine Example MEA/MDEA using Packed Absorber

C- EQUIPMENT MODULES
```

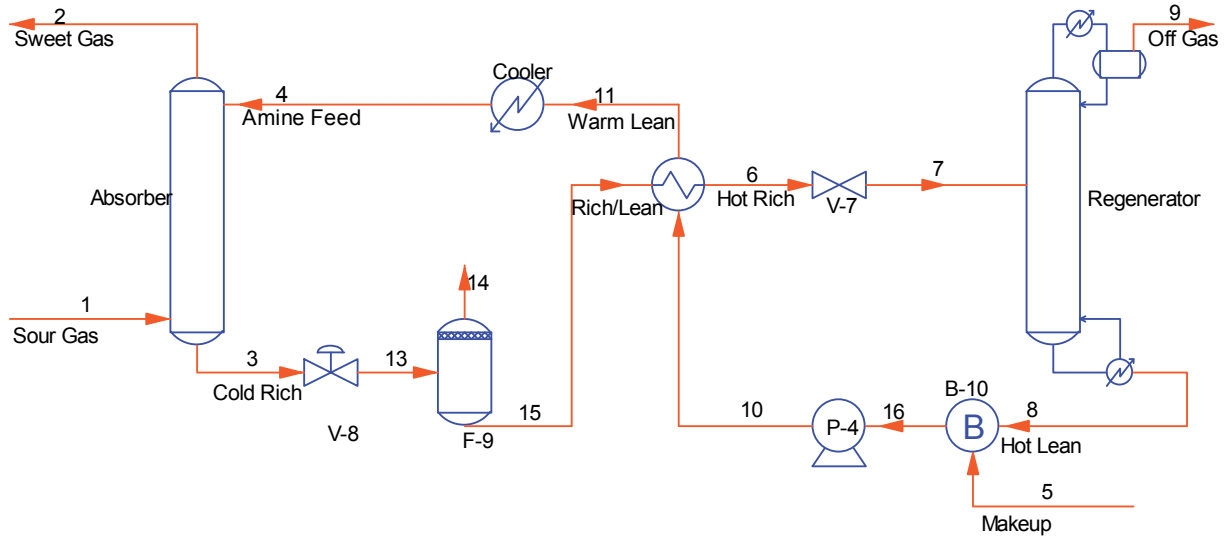
```
MIX 1 = M-1, 11,10,-12,

HEA EXC 2 = Cooler, 9,-10,
  U(BTU/HR/FT2/F) = 50.
  SHE = 1
  TEM OUT(F) = 120.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 0.

VAL 3 = V-3, 4,-5,
  PRE OUT(PSIG) = 35.3

HEA EXC 4 = RichLeanX, 8,3,-9,-4,
  U(BTU/HR/FT2/F) = 50.
  TEM APP(F) = 25.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,

PUM 5 = P-5, 7,-8,
  PRE OUT(PSIG) = 1000.
  EFF = 0.9
  ELE DRI

AMCL 6 = Absorber, 12,1,-2,-3,
  ABS1
  PACK DEPTH PER TYPE(FT) = 2*15.,
  PACK MAT = 2*METAL,
  PACK TYP = 2*DUMPED,
  PACK SIZE(IN) = 2*0.675,
  PACK FAC = 2*70.,
  PACK DRY AREA(1/FT) = 2*105.,
  TOW DIA FLOOD = 2*75.,
  REDIST DEPTH(FT) = 15.,
  LOC FEE (FT) = 1,30.
  PRE TOP(PSIG) = 1000.
  PRE DEL = DPC
  PRO (LBMOL/HR) = 0,4000.
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 5.E-004

AMCL 7 = Regen, 5,-6,-7,
  REG1
  TRA = 10
  TRA TYP = 10*SIE,
  TRA PAS = 2*2,8*4,
  TOW DIA(FT) = 10*3.,
  TRA WEI(IN) = 10*2.,
  TRA SPA(IN) = 10*30.,
  TRA ARE PER TOW = 10*80.,
  TRA PER OPE = 10*15.,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIG) = 8.
  PRE REB(PSIG) = 15.
  PRO (LBMOL/HR) = 242.07,0
  TEM CON(F) = 110.
  REB DUT(BTU/HR) = 12500000.
  SYS FAC = 0.85
  MAX ITER = 150
  TOL = 1.E-003
F-      POST
F-      F11 = GET (STR(11), FLO COM(1))
F-      F1 = GET (STR(1), FLO COM(1))
F-      F6 = GET (STR(6), FLO COM(1))
F-      F2 = GET (STR(2), FLO COM(1))
F-      FIN = F1 + F11
F-      FOUT = F2 + F6
F-      DELFLO = (FOUT-FIN) * 0.8
F-      FLO11 = F11 + DELFLO
```

```
F-      WRITE (1, 111) FIN, F2, F6
F-      WRITE (6, 111) FIN, F2, F6
F-  111 FORMAT (' WATER IN = ', F8.2, ' F2= ', F8.2, ' F6 =', F8.2)
F-      WRITE (1, 121) F11, FLO11
F-      WRITE (6, 121) F11, FLO11
F-  121 FORMAT (' WATER DELTA = ', F8.2, '   NEW WATER MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (1)) = FLO11
F-      FM11 = GET (STR(11), FLO COM(2))
F-      FM6 = GET (STR(6), FLO COM(2))
F-      FM2 = GET (STR(2), FLO COM(2))
F-      FMOUT = FM6 + FM10
F-      DELFM = (FMOUT-FM11) * 0.8
F-      FLOM11 = FM11 + DELFM
F-      WRITE (1, 112) FM11, FM2, FM6
F-      WRITE (6, 112) FM11, FM2, FM6
F-  112 FORMAT (' DEA IN = ', F8.2, ' FM2= ', F8.2, ' FM6 =', F8.2)
F-      WRITE (1, 122) FM11, FLOM11
F-      WRITE (6, 122) FM11, FLOM11
F-  122 FORMAT (' DEA DELTA = ', F8.2, '   NEW DEA MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (2)) = FLOM11
F-      FMA11 = GET (STR(11), FLO COM(3))
F-      FMA6 = GET (STR(6), FLO COM(3))
F-      FMA2 = GET (STR(2), FLO COM(3))
F-      FMAOUT = FMA6 + FMA2
F-      DELFMA = (FMAOUT-FMA11) * 0.8
F-      FLOA11 = FMA11 + DELFMA
F-      WRITE (1, 113) FMA11, FMA2, FMA6
F-      WRITE (6, 113) FMA11, FMA2, FMA6
F-  113 FORMAT (' MDEA IN = ', F8.2, ' FMA2= ', F8.2, ' FMA6 =', F8.2)
F-      WRITE (1, 123) FMA11, FLOA11
F-      WRITE (6, 123) FMA11, FLOA11
F-  123 FORMAT (' MDEA DELTA = ', F8.2, '   NEW MDEA MAKEUP = ', F8.2)
F-      SET (STR(11), FLO COM (3)) = FLOA11
F-      CALL INISTR (11)

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 20
  CON TOL = 1.E-003
  REC STR = 7,12,
  SAT FEE WAT = 1,
  FLO(LBMOL/HR)1 = 2.565,2*0.,54.851,191.978,2495.709,
  TP (F,PSIG) 1 = 100.,1000.
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO (FRA) 3 = 0.79984,2.74842E-002,9.69907E-002,2.35E-002,5.138E-002,
  8.0298E-004,
  TOT FLO (LBMOL/HR) 3 = 4156.6623
  TP (F,PSIA) 3 = 179.19,1014.696
  NAM STR 3 = Cold Rich
  NAM STR 4 = Strm 8
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 12
  FLO(LBMOL/HR)7 = 3363.,114.,403.,4.,6.,0.,
  TP (F,PSIG) 7 = 256.9,15.4
  NAM STR 7 = Recycle
  NAM STR 8 = Hot Lean
  NAM STR 9 = Warm Lean
  NAM STR 10 = Strm 5
  FLO (FRA) 11 = 1.,5*0.,
  TOT FLO (LBMOL/HR) 11 = 1.E-002
  TP (F,PSIG) 11 = 120.,1000.
  NAM STR 11 = Make Up
  FLO(LBMOL/HR)12 = 3386.,114.,403.,5.,6.,0.,
  TP (F,PSIA) 12 = 125.,1000.
  NAM STR 12 = Abs Feed
```

```
REC SEQ = 5, 6, 4, 2, 1, 3, 7, 5, 4, 2, 1
END
```

MxpInt1.psd - Mixed Amine Plant for CO2 / H2S Removal

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



Assumptions
Use 16 weight % solution of DEA.blended with 28 wt % of MDEA to make about 100 gal/min of aqueous recirculating amine. Make-up water is estimated automatically by the Mass Balance unit module.

Objectives
To achieve about 1.7 mol % of CO2 in the sweet gas, with top temperature of 150 F

Specifications
Feed Gas
Temperature = 90 F
Pressure = 880 psig

Component	Mole Percent (Dry Basis)
Water	0.000
Nitrogen	0.170
CO2	5.600
H2S	0.150
Methane	75.119
Ethane	11.895
Propane	4.799
i-Butane	0.716
n-Butane	1.551

Equipment
Absorber:
Diameter = 4 ft
Number of Trays=10
Tray Parameters = Valve Trays, 2 ft spacing, 2 inch weirs, 63.6 % Active Area
Rich/Lean Exchanger: Temperature Approach = 10 F
Valve: Pressure Out = 25.2 psia
Lean Cooler: Temperature Out = 127 F
Pump: Pressure Out = 880 psig
Mixer: None
Regenerator:
Diameter = 3 ft
Number of Trays=10
Tray Parameters = 1 pass Valve Trays, 2 ft spacing, 2 inch weirs,
64 % Active Area
Overhead Condenser Temperature = 165 F
Overhead Condenser Pressure = 11.5 psig

Reboiler Pressure = 15 psig

Process Parameters:

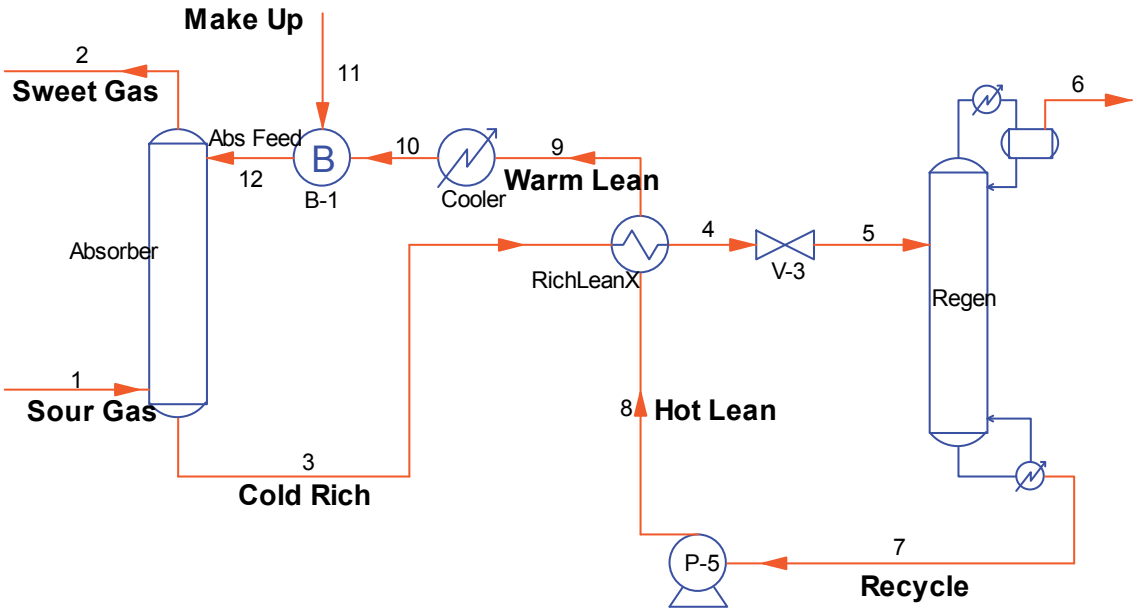
Lean Solution Rate =100 USgpm
DEA Concentration = 16 wt percent
MDEA Concentration = 28 wt percent

```
AB123.
*Sour Gas treatment with Mixed Amine

C- EQUIPMENT MODULES

AMCL 1 = Absorber, 4,1,-2,-3,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*4,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*63.6,
  LOC FEE = 1,10
  PRE TOP(PSIG) = 880
  PRE DEL = DPC
  PRO (LBMOL/HR) = 2000,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

AMCL 3 = Regenerator, 7,-9,-8,
  REG1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*64,
  LOC FEE = 2,
  PRE TOP(PSIG) = 11.5
  PRE DEL = DPC
  PRE CON(PSIG) = 11.5
  PRE REB(PSIG) = 15
  PRO (LBMOL/HR) = 100,0
  TEM CON(F) = 165
  REB DUT(BTU/HR) = 6700000
  SYS FAC = 0.85
  MAX ITER = 200
  TOL = 0.001

PUM 4 = P-4, 16,-10,
  PRE OUT(PSIG) = 880
  EFF = 0.9
  ELE DRI

HEA EXC 5 = Rich/Lean, 10,15,-11,-6,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

HEA EXC 6 = Cooler, 11,-4,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
```

```
  DEL(PSI) = 0

VAL 7 = V-7, 6,-7,
  PRE OUT(PSIA) = 25.2

VAL 8 = V-8, 3,-13,
  PRE OUT(PSIG) = 75

FLA 9 = F-9, 13,-14,-15,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

MAS BAL 10 = B-10, 5,8,-16,
  REC(GAL/MIN) = 105
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 5
  INP = -1
  OUT = -1

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,49,50,2,3,4,5,6,4051,7051,
  NAM 62 = 'WATER'
  NAM 46 = 'NITROGEN'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 5 = 'ISOBUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  AME UNI OUT
  MAX = 15
  CON TOL = 0.001
  REC STR = 8,4,
  ONE LOOP
  FLO (FRA) 1 = 0,0.0017,0.056,0.0015,0.75119,0.11895,0.04799,0.00716,0.01551,
  TOT FLO (MMSCF/DAY) 1 = 24.8
  TP (F,PSIG) 1 = 97,880
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  NAM STR 3 = Cold Rich
  FLO (FRA MAS) 4 = 0.51,8*0,0.16,0.33,
  TOT FLO (GAL/MIN) 4 = 105
  TP (F,PSIG) 4 = 127,880
  NAM STR 4 = Amine Feed
  FLO (FRA) 5 = 1,
  TOT FLO (GAL/MIN) 5 = 1
  TP (F,PSIA) 5 = 85,30
  NAM STR 5 = Makeup
  NAM STR 6 = Hot Rich
  NAM STR 7 = Strm 7
  FLO (FRA MAS) 8 = 0.51,8*0,0.16,0.33,
  TOT FLO (GAL/MIN) 8 = 105
  TP (F,PSIG) 8 = 248,15
  NAM STR 8 = Hot Lean
  NAM STR 9 = Off Gas
  NAM STR 10 = Strm
  NAM STR 11 = Warm Lean
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
END
```

Mxplnt2.psd - Mixed Amine Plant 2

The objective of this simulation is to contact CO2 (7 mol %) and H2S (2 mol %) in the natural gas stream with a aqueous stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .005 mol % of H2S and 1.25 mole % of CO2 per 25 MMSCFD of gas processed.



Assumptions

Use 10 weight % solution of DEA.blended with 39 wt % of MDEA to make about 250 gal/min of aqueous recirculating amine. Make-up water and the amine losses are estimated and made-up automatically by the Mass Balance unit module.

Objectives

To achieve 1.25 mol % of CO2 in the sweet gas, with top temperature of 137 F

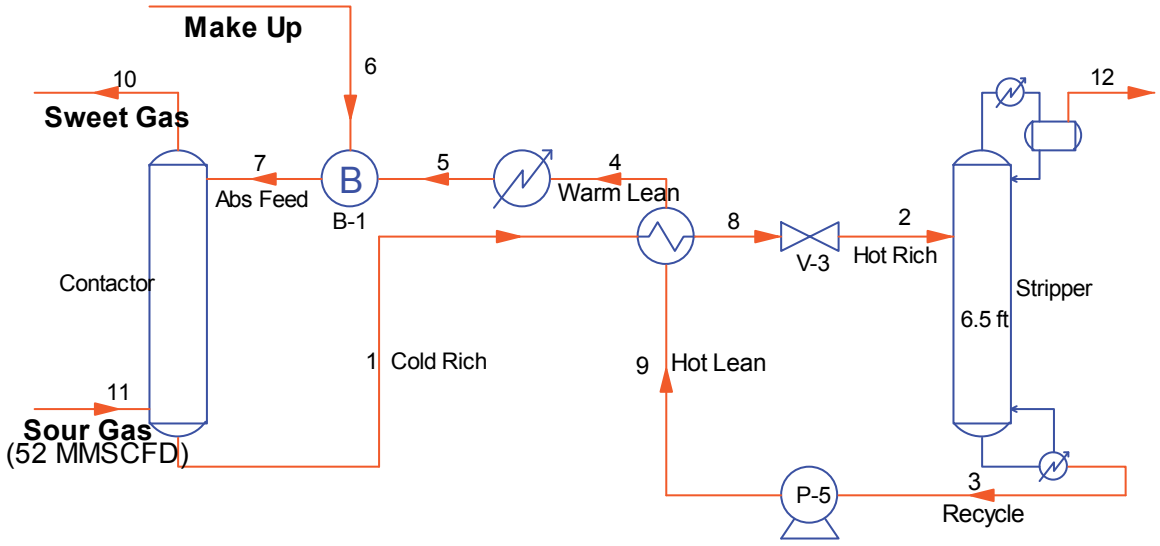
```
AB123.
*Mixed Amine Example MEA/MDEA

C- EQUIPMENT MODULES

MAS BAL 1 = B-1, 11,10,-12,
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 11
  INP = -1
  OUT = -1

HEA EXC 2 = Cooler, 9,-10,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 125
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 3 = V-3, 4,-5,
  PRE OUT(PSIG) = 35.3

HEA EXC 4 = RichLeanX, 8,3,-9,-4,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 25
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
```

```
  HOR
  DEL(PSI) = 0, 0,

PUM 5 = P-5, 7,-8,
  PRE OUT(PSIG) = 1000
  EFF = 0.9
  ELE DRI

AMCL 6 = Absorber, 12,1,-2,-3,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(IN) = 10*54,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 1,10
  PRE TOP(PSIG) = 1000
  PRE DEL = DPC
  PRO (LBMOL/HR) = 0,4000
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

AMCL 7 = Regen, 5,-6,-7,
  REG1
  TRA = 10
  TRA TYP = 10*SIE,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3.5,
  TRA WEI(IN) = 10*1.5,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*80,
  TRA PER OPE = 10*15,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28
  PRE DEL = DPC
  PRE CON(PSIG) = 12.6
  PRE REB(PSIG) = 15.4
  PRO (LBMOL/HR) = 253,0
  REF RAT = 0.87
  SYS FAC = 0.85
  MAX ITER = 300
  TOL = 0.007

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  NAM 62 = 'WATER'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 2 = 'METHANE'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  AME UNI OUT
  MAX = 20
  CON TOL = 0.003
  REC STR = 7,12,
  REC SEQ = 6,4,3,7,5,4,2,1,
  SAT FEE WAT = 1,
  FLO(LBMOL/HR)1 = 2.565,2*0,54.851,191.978,2495.709,
  TP (F,PSIG) 1 = 100,1000
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO (FRA) 3 = 0.79984,0.0274842,0.0969907,0.0235,0.05138,0.00080298,
```

```
  TOT FLO (LBMOL/HR) 3 = 4156.6623
  TP (F,PSIA) 3 = 179.19,1014.696
  NAM STR 3 = Cold Rich
  NAM STR 4 = Strm 8
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 12
  FLO(LBMOL/HR)7 = 3363,114,403,4,6,0,
  TP (F,PSIG) 7 = 256.9,15.4
  NAM STR 7 = Recycle
  NAM STR 8 = Hot Lean
  NAM STR 9 = Warm Lean
  NAM STR 10 = Strm 5
  FLO (FRA) 11 = 1,
  TOT FLO (LBMOL/HR) 11 = 0.01
  TP (F,PSIG) 11 = 125,1000
  NAM STR 11 = Make Up
  FLO(LBMOL/HR)12 = 3386,114,403,5,6,0,
  TP (F,PSIA) 12 = 125,1000
  NAM STR 12 = Abs Feed
END
```

MxpInt3.psd - CO2 / H2S Removal with only MEA

The objective of this simulation is to contact CO2 (2 mol%) and H2S (.5 mol %) in the natural gas stream with an aqueous stream with a one amine, MEA. Note that for using pure MEA, the other amine (DEA) can be a negligible amount but should be present in the flowsheet so that Mixed Amine thermo can be used. In the sweet gas stream the acid gases should be no greater than 500 ppm of H2s and 0.025 mole % of CO2 per 52 MMSCFD of gas processed.



Assumptions

Use 15 weight % solution of MEA.blended with 0 wt % of DEA to make about 370 gal/min of aqueous recirculating amine.

Make-up water and the amine losses are estimated and made-up automatically by the Mass Balance unit module.

Objectives

To achieve less than 500 ppm of H2S in the sweet gas, with top temperature of 107 F

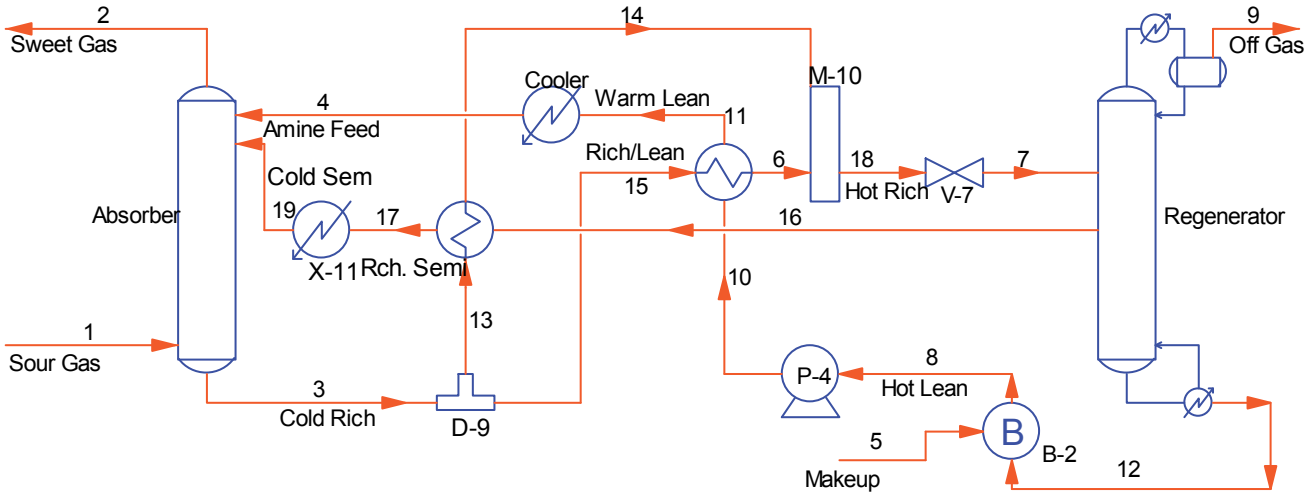
```
AB123.
*Mixed Amine Example MEA 15 % /MDEA- 0 %

C- EQUIPMENT MODULES

MAS BAL 1 = B-1, 6,5,-7,
  REC(GAL/MIN) = 350
  ACC = 0.8
  IDS = 62,
  VAR = 6
  INP = -1
  OUT = -1

HEA EXC 2 = X-2, 4,-5,
  U(BTU/HR/FT2/F) = 50
```

```
  TEM OUT(F) = 110
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 3 = V-3, 8,-2,
  PRE OUT(PSIA) = 26

HEA EXC 4 = X-4, 9,1,-4,-8,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 25
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

PUM 5 = P-5, 3,-9,
  PRE OUT(PSIA) = 890
  EFF = 0.9
  ELE DRI

AMCL 6 = Contactor, 7,11,-10,-1,
  ABS1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(IN) = 10*55,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 1,10
  PRE TOP(PSIA) = 890
  PRE BOT(PSIA) = 900
  PRO (LBMOL/HR) = 5000,0
  SYS FAC = 0.75
  MAX ITER = 200
  TOL = 0.0005

AMCL 7 = Stripper, 2,-12,-3,
  REG1
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*6.5,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*70,
  LOC FEE = 2,
  PRE TOP(PSIG) = 12
  PRE DEL = DPC
  PRE CON(PSIG) = 12
  PRE REB(PSIG) = 14.5
  PRO (LBMOL/HR) = 145,0
  TEM CON(F) = 119
  REB DUT(BTU/HR) = 14500000
  SYS FAC = 0.75
  MAX ITER = 200
  TOL = 0.001

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,49,50,62,4155,4051,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 62 = 'WATER'
```

```
  NAM 4155 = 'MEA'
  NAM 4051 = 'DEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  AME UNI OUT
  MAX = 15
  CON TOL = 0.005
  REC STR = 7,3,
  NAM STR 1 = Cold Rich
  FLO (FRA) 2 = 0.00080298,2*0,0.05138,0.0235,0.79984,0,0.0274842,
  TOT FLO (LBMOL/HR) 2 = 4156.6623
  TP (F,PSIA) 2 = 179.19,50
  NAM STR 2 = Hot Rich
  FLO(LBMOL/HR)3 = 3*0,60.567,1.052,8575.602,446.398,0,
  TP (F,PSIA) 3 = 230,890
  NAM STR 3 = Recycle
  NAM STR 4 = Warm Lean
  FLO(LBMOL/HR)5 = 1,2*0,86.9,5.3,9431.1,491,0,
  TP (F,PSIA) 5 = 110,914.7
  NAM STR 5 = Strm 5
  FLO (FRA) 6 = 5*0,1,
  TOT FLO (LBMOL/HR) 6 = 0.01
  TP (F,PSIA) 6 = 110,890
  NAM STR 6 = Make Up
  FLO(LBMOL/HR)7 = 1,2*0,60.567,1.052,8575.602,446.398,0,
  TP (F,PSIA) 7 = 110,890
  NAM STR 7 = Abs Feed
  NAM STR 8 = Strm 8
  NAM STR 9 = Hot Lean
  NAM STR 10 = Sweet Gas
  FLO(LBMOL/HR)11 = 5014,557.1,0,111.4,27.86,3*0,
  TP (F,PSIA) 11 = 90,914.7
  NAM STR 11 = Sour Gas
  NAM STR 12 = Strm 12
     CONverge on TEMperature = 3, 7
END
```

MxpIntsd.psd - Mixed Amine Plant with Side Draw and Side Feed

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a mixture of two amines (DEA and MDEA). In the sweet gas stream the acid gases should be no greater than .05 mol % of H2S and 1.7 mole % of CO2 per 24.8 MMSCFD of gas processed.



Assumptions

Use 16 weight % solution of DEA.blended with 28 wt % of MDEA to make about 100 gal/min of aqueous recirculating amine. Make-up water is estimated automatically by the Mass Balance unit module.

Objectives
To achieve 1.7 mol % of CO2 in the sweet gas, with top temperature of 150 F
Sidedraw from tray#5 is 2 lbmol/hr back to absorber as side feed.

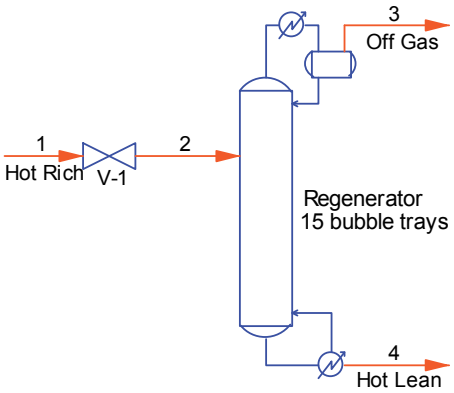
```
AB123.
*Sour Gas treatment with Mixed Amine-Side Feed & Side Draw

C- EQUIPMENT MODULES

AMCL 1 = Absorber, 4,19,1,-2,-3,
  ABS2
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*4,
  TRA WEI(IN) = 10*3,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*63.6,
  LOC FEE = 1,2,10
  PRE TOP(PSIG) = 880
  PRE DEL = DPC
  PRO (LBMOL/HR) = 2000,0
  SYS FAC = 0.8
  MAX ITER = 200
  TOL = 0.0005

MAS BAL 2 = B-2, 5,12,-8,
  REC(GAL/MIN) = 105
  ACC = 1
  IDS = 62,4051,7051,
  VAR = 5
  INP = -1
  OUT = -1

AMCL 3 = Regenerator, 7,-9,-16,-12,
  REG2
  TRA = 10
  TRA TYP = 10*VAL,
  TRA PAS = 10*1,
  TOW DIA(FT) = 10*3,
  TRA WEI(IN) = 10*2,
  TRA SPA(FT) = 10*2,
  TRA ARE PER TOW = 10*64,
  LOC FEE = 2,
  LOC SID = 5
  PRE TOP(PSIG) = 11.5
  PRE DEL = DPC
  PRE CON(PSIG) = 11.5
  PRE REB(PSIG) = 15
  PRO (LBMOL/HR) = 100,2,0
  TEM CON(F) = 165
  REB DUT(BTU/HR) = 6700000
  SYS FAC = 0.85
  MAX ITER = 200
  TOL = 0.001

PUM 4 = P-4, 8,-10,
  PRE OUT(PSIG) = 880
  EFF = 0.9
  ELE DRI

HEA EXC 5 = Rich/Lean, 10,15,-11,-6,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

HEA EXC 6 = Cooler, 11,-4,
```

```
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

VAL 7 = V-7, 18,-7,
  PRE OUT(PSIA) = 25.2

HEA EXC 8 = Rch. Semi, 16,13,-17,-14,
  U(BTU/HR/FT2/F) = 50
  TEM APP(F) = 10
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0, 0,

DIV 9 = D-9, 3,-13,-15,
  FLO(FRA) = 0.05

MIX 10 = M-10, 14,6,-18,

HEA EXC 11 = X-11, 17,-19,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = 127
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,46,49,50,2,3,4,5,6,4051,7051,
  NAM 62 = 'WATER'
  NAM 46 = 'NITROGEN'
  NAM 49 = 'CARBON DIOXIDE'
  NAM 50 = 'HYDROGEN SULFIDE'
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 5 = 'ISOBUTANE'
  NAM 6 = 'N-BUTANE'
  NAM 4051 = 'DEA'
  NAM 7051 = 'MDEA'
  AMINEK
  AMINEH
  AMINED
  DEN LIQ = AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  AME UNI OUT
  MAX = 100
  CON TOL = 0.001
  REC STR = 8,4,3,16,
  FLO (FRA) 1 = 0,0.0017,0.056,0.0015,0.75119,0.11895,0.04799,0.00716,0.01551,
  TOT FLO (MMSCF/DAY) 1 = 24.8
  TP (F,PSIG) 1 = 97,880
  NAM STR 1 = Sour Gas
  NAM STR 2 = Sweet Gas
  FLO(LBMOL/HR)3 = 1556.8,0,114.74,4.11,1.28,0.17,0.04,0.001,0.007,77.57,120,
  TP (F,PSIG) 3 = 142.4,880
```

```
  NAM STR 3 = Cold Rich
  FLO(LBMOL/HR)4 = 1572,0,3.4,0.001,5*0,77.4,119.6,
  TP (F,PSIG) 4 = 127,880
  NAM STR 4 = Amine Feed
  FLO (FRA) 5 = 1,
  TOT FLO (LBMOL/HR) 5 = 1
  TP (F,PSIA) 5 = 85,30
  NAM STR 5 = Makeup
  NAM STR 6 = Hot Rich
  NAM STR 7 = Strm 7
  FLO (FRA) 8 = 0.6,0,0.01,0.001,0,3*0.005,0.015,0.16,0.33,
  TOT FLO (GAL/MIN) 8 = 105
  TP (F,PSIG) 8 = 248,15
  NAM STR 8 = Hot Lean
  NAM STR 9 = Off Gas
  NAM STR 10 = Strm
  NAM STR 11 = Warm Lean
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  FLO(LBMOL/HR)16 = 2,
  TP (F,PSIA) 16 = 150,14.7
  NAM STR 16 = Strm 16
  NAM STR 17 = Strm 17
  NAM STR 18 = Strm 18
  NAM STR 19 = Cold Sem
C- REC SEQ=4,1,5,6,2,7,3,4,5,6,2
END
```

Mxregn.psd - Mixed Amine Stripper Example

The objective of this simulation is to strip the enriched mixed amine free of water and residual light ends by contacting with a mixture of two amines (DEA and MDEA). The rich hot lean amine exits at the bottom of stripper with no loss of amine at the top of the regenerator in the off gas.



Assumptions
Use 36 wt % MDEA and 9 wt% DEA

Objectives
1E-05 mol % of Amines in the off gas, with top temperature of 125 F

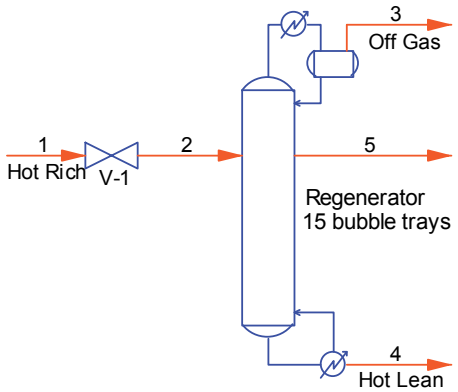
Specification:
Tower diameter 3 ft inches. Active area 80%.
Tower internals - 15 bubble trays, two pass, 2 ft tray spacing and 3 inches weir height.

```
AB123.
*Mixed Amine Stipper Example

C- EQUIPMENT MODULES

VAL 1 = V-1, 1,-2,
  PRE OUT(PSIA) = 50.

AMCL 2 = Regenerator, 2,-3,-4,
  REG1
  TRA = 15
  TRA TYP = 15*BUB,
```

```
  TRA PAS = 15*2,
  TOW DIA(FT) = 15*3.,
  TRA WEI(IN) = 15*3.,
  TRA SPA(FT) = 15*2.,
  TRA ARE PER TOW = 15*80.,
  LOC FEE = 1,
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIA) = 27.3
  PRE REB(PSIA) = 29.66
  PRO (LBMOL/HR) = 250.,0
  TEM CON(F) = 125.
  REB DUT(BTU/HR) = 12005700.
  SYS FAC = 0.75
  MAX ITER = 150
  TOL = 1.E-003

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 10
  CON TOL = 1.E-003
  NAM STR 3 = Off Gas
  FLO(FRA)1 = 0.7998,2.748E-002,9.699E-002,2.35E-002,5.138E-002,8.029E-004,
  TOT FLO(LBMOL/HR)1 = 4156.662
  TP (F,PSIA) 1 = 230.,1000.
  NAM STR 1 = Hot Rich
  NAM STR 2 = Strm 2
  NAM STR 4 = Hot Lean
END
```

Mxregnsd.psd - Amine Stripper with Side Draw

The objective of this simulation is to strip the enriched mixed amine free of water and residual light ends by contacting with a mixture of two amines (DEA and MDEA). A side draw at tray #5 is also used. The rich hot lean amine exits at the bottom of stripper with no loss of amine at the top of the regenerator in the off gas.



Assumptions
Use 36 wt % MDEA and 9 wt% DEA

Objectives
1E-05 mol % of Amines in the off gas, with top temperature of 125 F.
Specification:
Tower diameter 3 ft inches. Active area 80%.
Tower internals - 15 bubble trays, two pass, 2 ft tray spacing and 3 inches weir height.
Side draw 2 lbmol/hr at tray 5.

```
AB123.
*Mixed Amine Stipper Example -Side Draw

C- EQUIPMENT MODULES

VAL 1 = V-1, 1,-2,
  PRE OUT(PSIA) = 50.

AMCL 2 = Regenerator, 2,-3,-5,-4,
  REG2
  TRA = 15
  TRA TYP = 15*BUB,
  TRA PAS = 15*2,
  TOW DIA(FT) = 15*3.,
  TRA WEI(IN) = 15*3.,
  TRA SPA(FT) = 15*2.,
  TRA ARE PER TOW = 15*80.,
  LOC FEE = 1,
  LOC SID = 5
  PRE TOP(PSIA) = 28.
  PRE DEL = DPC
  PRE CON(PSIA) = 27.3
  PRE REB(PSIA) = 29.66
  PRO (LBMOL/HR) = 250.,2.,0
  TEM CON(F) = 125.
  REB DUT(BTU/HR) = 12005700.
  SYS FAC = 0.75
  MAX ITER = 150
  TOL = 1.E-003

C- GENERAL SECTION
GENERAL
  COMPONENTS = 62,4051,7051,50,49,2,
  AMINEK
  AMINEH
  AMINED
  VIS VAP = AMINE
  VIS LIQ = AMINE
  THE CON VAP = AMINE
  THE CON LIQ = AMINE
  SUR TEN = AMINE
  STE TAB
  NO IMM
  MAX = 10
  CON TOL = 1.E-003
  NAM STR 3 = Off Gas
  FLO(FRA)1 = 0.7998,2.748E-002,9.699E-002,2.35E-002,5.138E-002,8.029E-004,
  TOT FLO(LBMOL/HR)1 = 4156.662
  TP (F,PSIA) 1 = 230.,1000.
  NAM STR 1 = Hot Rich
  NAM STR 2 = Strm 2
  NAM STR 4 = Hot Lean
  NAM STR 5 = Strm 5
END
```

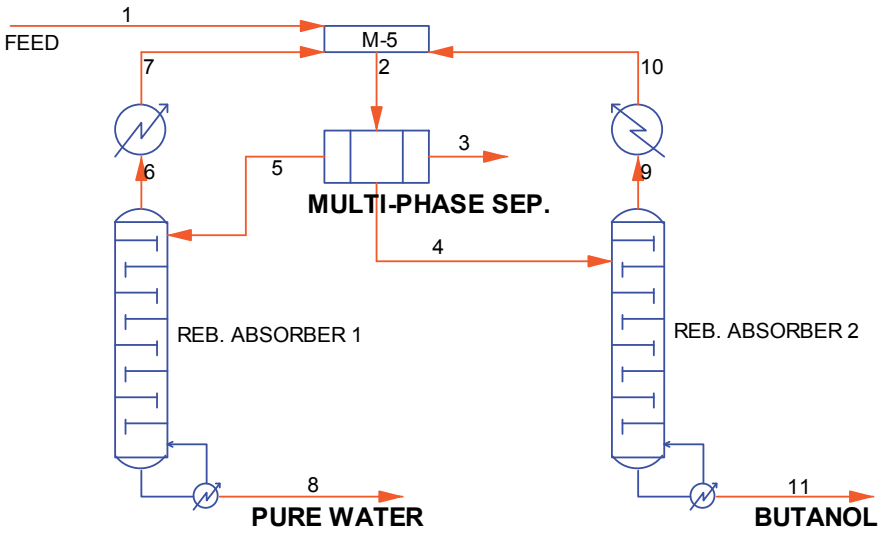
Chapter 59: General Samples - c:\designii\samples\general

Btextegd.dat – ChemTran Data File for TEG/BTEX Dehydration

It is a ChemTran created binary data file created by the ChemTran input file TEGBTEXCT.IN for use by DESIGN II at runtime of TEGBTEX.PSD flowsheet file.

Btnlh2o.psd - Azeotropic Separation of Butanol and Water

The sample problem is to illustrate the simulation of an azeotropic separation of Butanol and Water. The first distill column bottoms is pure water and the second column's bottom is pure butanol. The cooled top product from both the columns are mixed with the feed and taken through a multi-phase separator where two liquid phases separate and each stream is recycled back to the top of each column.



Butnwatr.dat – ChemTran Data File for Azeotropic Separation of Butanol and Water

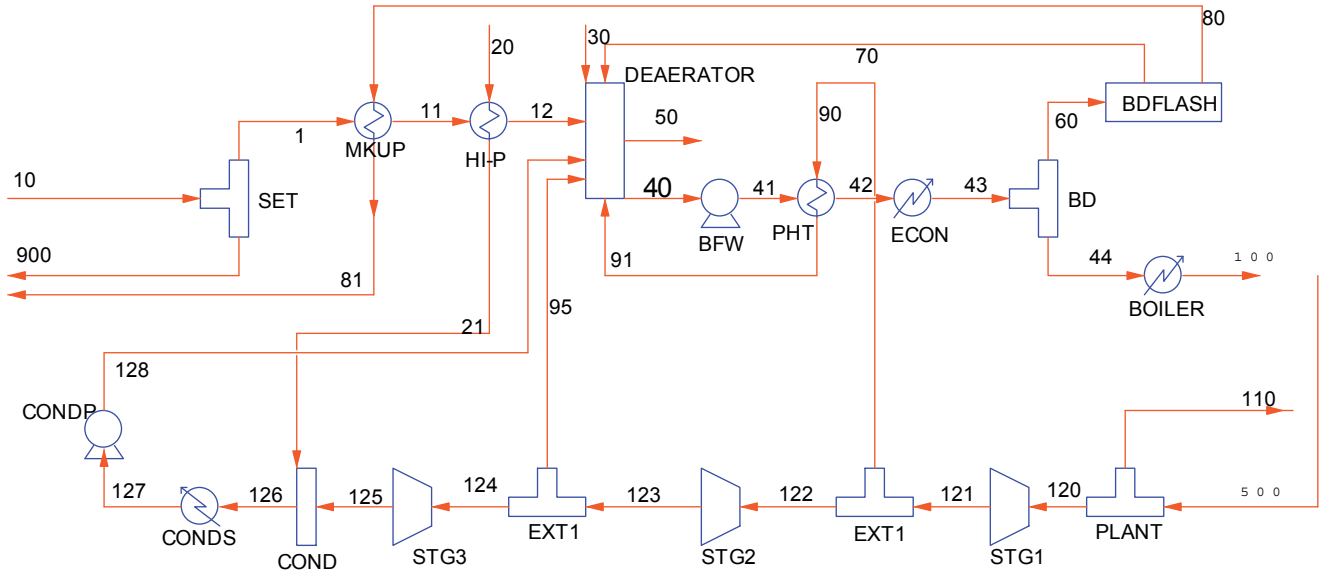
It is a ChemTran created binary data file created by the ChemTran input file BUTWTRCT.IN for use by DESIGN II at runtime of BTNLH2O.PSD flowsheet file.

Butwtrct.in – VLE Data Regression for Azeotropic Separation of Butanol and Water

It is a ChemTran VLE data regression input file for butanol, and water mixture for BTNLH2O.PSD file. This is already built into the flowsheet file BTNLH2O.PSD.

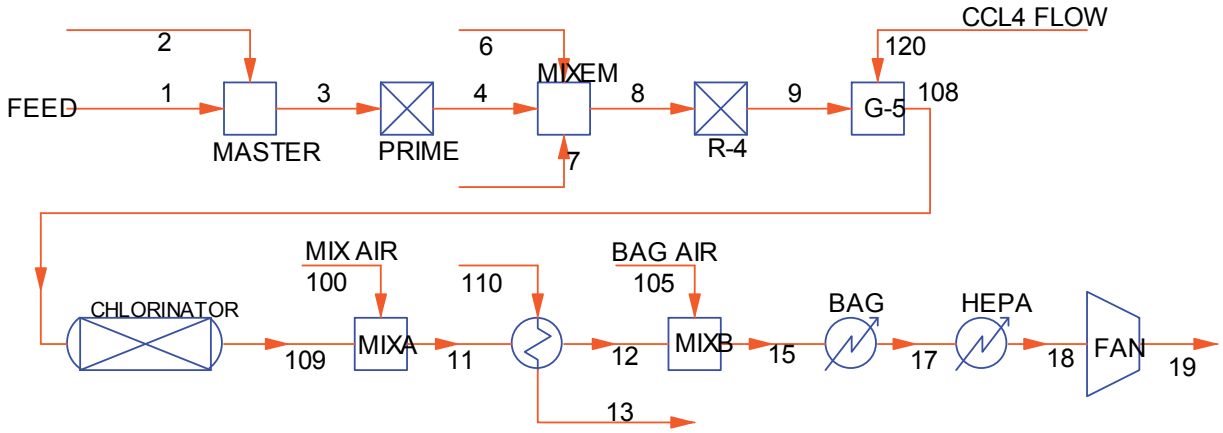
Cogents.psd - Cogeneration of Steam

A three stage expansion process is shown above to use the work provided by the generation of a 600 lbs of steam from a boiler. The expanded gases are taken through a deaerator (valve module) to 10 psig. The product is pumped to the boiler (heat exchanger module) to meet the required steam conditions (600psia & 800 F).



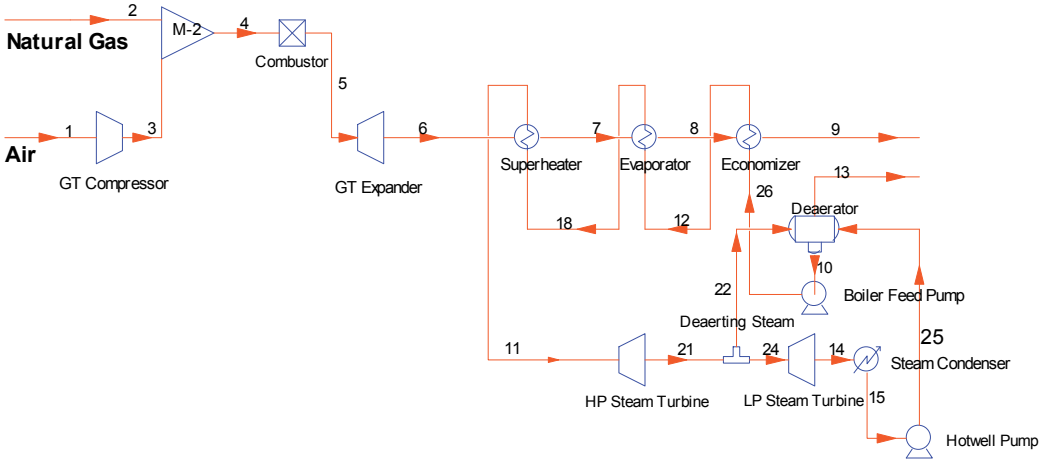
Combstn.psd - Combustion of Decanes, Natphalene & Phenanthrene

Sample simulation showing combustion of decanes, natphalene, & phenanthrene, with air and the production of HCL from CCL4 & water via a chlorinator.



Comcycle.psd – Combined Cycle Gas & Steam Turbine Power

Sample simulation showing combustion of natural gas with air and the combined cycle gas & steam power generation with a 70 MW Gas Turbine, Heat Recovery Steam Generator and a 30 MW Steam Turbine.

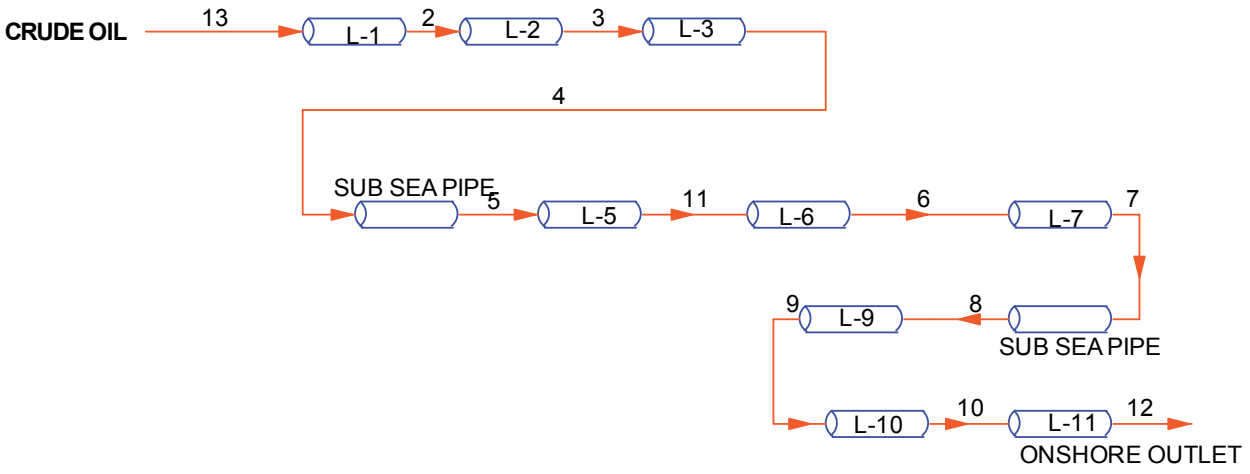


```
AB123.
*Combined Cycle Gas-Steam Power Generation

C- EQUIPMENT MODULES

COMPRE 1 = GT Compressor, 1,-3,
  PRE OUT(BAR) = 11.
  EFF = 0.92
  CAL POL
  ELE DRI
  STA = 1
  TEM INT(C) = 35.

MIX 2 = M-2, 2,3,-4,

EQU REA 3 = Combustor, 4,-5,
   ADIabatic

EXP 4 = GT Expander, 5,-6,
  PRE OUT(BAR) = 1.035
  EFF = 0.85

HEA EXC 5 = Superheater, 6,18,-7,-11,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 480.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 1.E-002, 1.8,

HEA EXC 6 = Evaporator, 7,12,-8,-18,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 248.6
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 1.E-002, 0.,

HEA EXC 7 = Economizer, 8,26,-9,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT TUB(C) = 248.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(BAR) = 2.E-003, 11.2,

VAL 8 = Deaerator, 22,25,-13,-10,
  DEL(BAR) = 0.

PUM 9 = Boiler Feed Pump, 10,-26,
  PRE OUT(BAR) = 50.
  EFF = 0.9
  ELE DRI

HEA EXC 10 = Steam Condenser, 14,-15,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(C) = 25.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.
  TEM WAT IN(F) = 60.
  TEM WAT OUT(F) = 80.

EXP 13 = HP Steam Turbine, 11,-21,
```

```
  PRE OUT(BAR) = 2.9
  EFF = 0.8

DIV 14 = Deaerting Steam, 21,-22,-24,
  FLO(KG/SEC) = 5.4

EXP 15 = LP Steam Turbine, 24,-14,
  PRE OUT(BAR) = 4.E-002
  EFF = 0.8

PUM 19 = Hotwell Pump, 15,-25,
  PRE OUT(BAR) = 10.
  EFF = 0.9
  ELE DRI

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,46,47,48,49,62,
  RKK
  STDH
  STDD
  SUR TEN = STD
  STE TAB
  QUA UNI OUT = KGMOL
  TEM UNI OUT = C
  PRE UNI OUT = BAR
  ENTH UNI OUT = KCAL
  TIM UNI OUT = SEC
  MAX = 50
  CON TOL = 1.E-005
  REC STR = 18,
  MET UNI
  VOL STA = NTP
  FLO (FRA) 1 = 2*0.,0.79,0.21,3*0.,
  TOT FLO (KG/SEC) 1 = 283.
  TP (C,BAR) 1 = 15.,1.013
  NAM STR 1 = Air
  FLO(KG/SEC)2 = 0.,4.7,0.4,2*0.,0.4,0.,
  TP (C,BAR) 2 = 25.,11.
  NAM STR 2 = Natural Gas
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  NAM STR 11 = Strm 11
  FLO(KG/SEC)12 = 6*0.,34.9,
  TP (C,BAR) 12 = 200.,38.8
  VAP FRA 12 = 1.E-002, 1
  NAM STR 12 = Strm 12
  NAM STR 13 = Strm 13
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  FLO(KG/SEC)18 = 6*0.,34.9,
  TP (C,BAR) 18 = 250.,38.8
  VAP FRA 18 = 0.99, 1
  NAM STR 18 = Strm 18
  NAM STR 21 = Strm 21
  NAM STR 22 = Strm 22
  NAM STR 24 = Strm 24
  NAM STR 25 = Strm 25
  NAM STR 26 = Strm 26
END
```

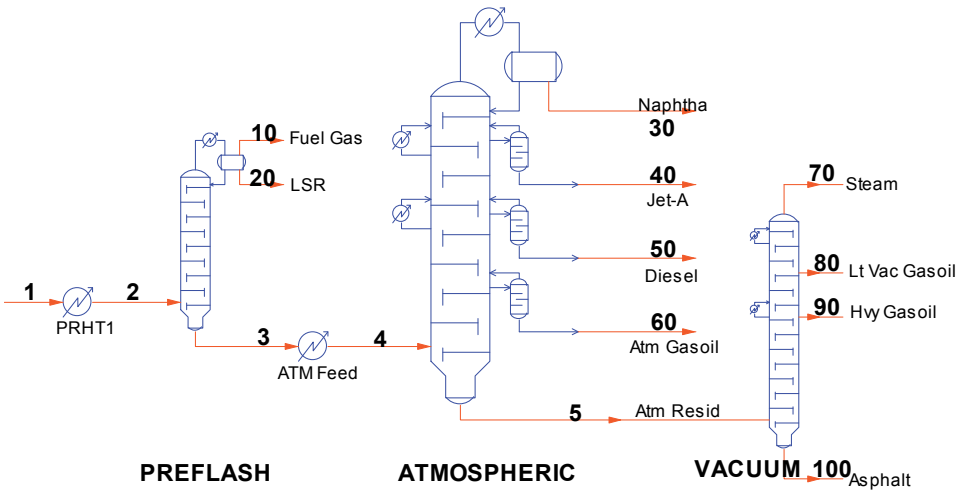
Crdpipln.psd - Transportation of Crude

The simulation above is an illustration of line modules transporting crude oil from an offshore production facility to onshore refining plants. Each line module has a specified diameter, elevation change, buried under soil or water. The results of the line module will show the flow regimes (turbulent, wavy, plug, annular, slug, etc...).



Crude.psd - Crude Unit Example

The purpose of this simulation is to determine the quality of products which can be made from a series of crude distillation columns for a feed of 100000 barrels per day. The amount of duty required for feed preheat (including feed furnaces for the atmospheric and vacuum columns) will also be calculated.



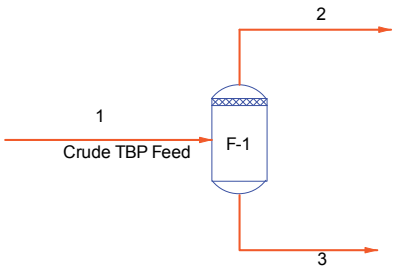

```
AB123.
*CRUDE UNITS -PREFLASH; ATMOSPHERIC AND VACUUM

C- HEAT EXCHANGER MODULE:    1   (PRHT)

HEAT EXCHANGER 1 = PRHT, 1,-2,
C- HEAT EXCHANGER SPECIFICATION
         TEMPERATURE OUT = 450,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 10,
C- REFINE CRUDE TOWER 2   (PRFL)

REFINE 2 = PRFL, 2,-10,-20,-3,

C-      PARTIAL CONDENSER
        PARTIAL CONDENSER
        STAGES = 8
        LOCATION OF FEEDS = 7,
        PRESSURE OF TOP = 41.7
        TEMPERATURE OF TOP TRAY = 340,
        TEMPERATURE OF BOTTOM = 450,
        PRODUCT RATES (BBL/DAY) = 0,0,80000,
        PRODUCT RATE GUESS = 280,1950,3900,
        TEMPERATURE OF CONDENSER = 160,

C-      PRESSURE DROPS FOR MAIN COLUMN AND
C-      SIDE STRIPPERS
```

```
        DELTA PRESSURE = 3,

C-      CONDENSER DESCRIPTION
        PRESSURE OF CONDENSER = 39.7,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 8,
        TEMPERATURE OF STEAM = 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 5000,

C-      COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = FUEL GAS,LSR,ATM.FEED,
C- HEAT EXCHANGER MODULE:    3  (PRHT 2)
HEAT EXCHANGER 3 = PRHT 2, 3,-4,
C- HEAT EXCHANGER SPECIFICATION
        TEMPERATURE OUT = 625,
C- OPTIONAL COMMANDS
        DELTA PRESSURE = 15,

C- REFINE CRUDE TOWER 4  (ATM)
REFINE4 = ATM, 4,-30,-40,-50,-60,-5,
C-      TOTAL CONDENSER
        TOTAL CONDENSER
        STAGES = 24,4,3,2,
        LOCATION OF FEEDS = 21,
        PRESSURE OF TOP = 20.7
        TEMPERATURE OF TOP TRAY = 360,
        TEMPERATURE OF BOTTOM = 620,
        PRODUCT RATES (BBL/DAY) = 0,35410,
        PRODUCT RATE GUESS = 730,1050,

C-      PRESSURE DROPS FOR MAIN COLUMN AND SIDE C-    STRIPPERS
        DELTA PRESSURE = 4,

C-      CONDENSER DESCRIPTION
        PRESSURE OF CONDENSER = 15.7,

C-      SIDE STRIPPER DESCRIPTION AND
C-      SPECIFICATION
        LOCATION OF DRAW TRAYS = 5,12,17
        LOCATION OF VAPOR RETURN = 5,12,17

C-      SIDE STRIPPER SPECIFICATIONS
        PRODUCT RATES FROM STRIPPER (BBL/DAY) = 8165,
        16967,10000,
        PRODUCT FROM STRIPPER GUESS = 615,1050,470,

C-      PUMPAROUND DESCRIPTIONS
        LOCATION OF PUMPAROUND = 7,5,13,12,
C-      PUMPAROUND SPECIFICATIONS
        PRODUCT OF PUMPAROUND (BBL/DAY) = 49000,11000,
        HEAT FROM PUMPAROUND = 40E6,15E6,
C-      PUMPAROUND MOLAR RATE ESTIMATES
        PRODUCT OF PUMPAROUND GUESS = 3600,700,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 25,29,32,34,
        TEMPERATURE OF STEAM = 4*400,
        PRESSURE OF STEAM = 4*60,
        STEAM RATES (LB/HR) = 12000,3300,1000,800,

C-    PRODUCT PROPERTY SPECIFICATIONS
        PRODUCT SPEC 1 = 625, ASTM95PT , PROD3, PROD4,

C- TRAY FIX COMMANDS
                FIX LIQUID (BBL/DAY) 1 = 1500, 20, H1,
                FIX GUESS = 50,

C-      COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = NAPHTHA,JET,DIESEL,AGO,RESID,
C- REFINE CRUDE TOWER 15  (VTWR)
```

```
REFINE 15 = VTWR, 5,-70,-80,-90,-100,

C-      PUMPAROUND INTERNAL REFLUX
        PUMREF
        STAGES = 6
        LOCATION OF FEEDS = 6,
        PRESSURE OF TOP (MMHG) = 75
        TEMPERATURE OF TOP TRAY = 450,
        TEMPERATURE OF BOTTOM = 700,

C-     SIDE DRAW DESCRIPTION
        LOCATION OF SIDE DRAWS = 2, 4,
        PRODUCT RATES (BBL/DAY) = 0,6200,18600,10600,
        PRODUCT RATE GUESS = 1.2,269.4,592.3,195.2,

C-      PRESSURE DROPS FOR MAIN COLUMN AND
C-      SIDE STRIPPERS
        DELTA PRESSURE (MMHG) = 30,

C-      PUMPAROUND DESCRIPTIONS
        LOCATION OF PUMPAROUND = 2,1,4,3,
C-      PUMPAROUND SPECIFICATIONS
        PRODUCT OF PUMPAROUND (BBL/DAY) = 20000,
        49000,
        HEAT FROM PUMPAROUND = 14E6,54E6,
C-      PUMPAROUND MOLAR RATE ESTIMATES
        PRODUCT OF PUMPAROUND GUESS = 870,1560,

C-      STEAM FEED SPECIFICATIONS
        LOCATION OF STEAM = 7,
        TEMPERATURE OF STEAM= 400,
        PRESSURE OF STEAM = 59,
        STEAM RATES (LB/HR) = 20000,

C- TRAY FIX COMMANDS
        FIX TEMP 1 = 150, 1, 1,
        FIX LIQ NET 2 = 35, 2, 3,
        FIX LIQ NET 3 = 2, 5, H1,

C-     COLUMN PRINT CONTROL
        REPORT = 1, 0, 0, 0, 0, 0,
        PLOT OFF,
        PRODUCT NAMES = STEAM,LVGO,HVGO,ASPHALT,

GENERAL
        COMPONENTS = 62,2,3,4,5,6,7,8,
C- THERMO PROPERTY OPTIONS
        ESSOTAB,  API,  STDD,
        VISCOSITY OPTION = STD,
        THERMAL CONDUCTIVITY OPTION = NBS81,
        PETROLEUM STREAMS =     10,20,30,40,
        50,60,70,80,90,100,

CRUDE PROPERTY SECTION
C- DATA FOR PSEUDO COMPONENTS WITHOUT C- ID NUMBERS (CUTS)
        TEMPERATURE INCREMENTS FOR TBP CURVES = 30,
        60,120,
        TEMPERATURE BREAKS FOR TBP CURVES (F) = 500,900,
CRUDE STREAM
C- BULK STREAM - PSEUDO COMPONENTS WITHOUT ID C- NUMBERS: 1 (STREAM 1)
        FEED NAME 1 = STREAM 1,
        TP 1 = 215,59.7,
        FEED DRY RATE (BBL/DAY) 1 = 1000000,
        FEED REAL PERCENTS 1 = .1,.2,.5,.5,1,1.5,2.5,
        FEED TBP 1 = -50,100,200,300,400,470,550,
                        650,750,850,1100,1300,1475,1670,
        FEED VOLUME PERCENTS 1 = 2,5,10,20,30,40,
                50,60,70,80,90,95,98,100,
        FEED GRAVITIES (API) 1 = 150,95,65,45,40,
                38,33,30,25,20,15,10,5,
        FEED GRAVITY VOLUME PERCENTS 1 = 2,5,10,20,
                30,40,50,60,70,80,90,95,98,

END
```

CrudeProperties.psd – Crude Properties Example

Sample simulation showing specification of a Crude TBP feed with Crude Lights Ends and Crude Property Curves (Molecular Weight, Viscosities, Sulfur and Pour Point).



Deglwatr.dat – ChemTran Data File for Glycol Dehydration

It is a ChemTran created binary data file created by the ChemTran input file DGMPNCT.IN for use by DESIGN II at runtime of GLYCOLDH.PSD flowsheet file below.

Dgmpnct.in – VLE Data Regression for Glycol Dehydration

It is a ChemTran VLE data regression input file for DEG and water mixture for GLYCOLDH.PSD file below. This is already built into the flowsheet file GLYCOLDH.PSD.

Dmfactda.dat – ChemTran Data File for Extractive Distillation of 1,3 Butadiene

It is a ChemTran created binary data file created by the ChemTran input file EXTRCT.IN for use by DESIGN II at runtime of EXTRDIS1.PSD flowsheet file.

Edward_K_without ChemTran.psd – Specifying Ionic Species and Reactions

The user must specify ionic species and reactions using the IONic COMponents and LIBrary REActions commands present in the database. No user defined ions and reactions can be present.

Edward_K_with ChemTran.psd – Specifying Ionic Species and Reactions

The user must specify ionic species and reactions using the IONic COMponents and LIBrary REActions commands. If user defined ions and reactions are present, then CHEmtran file such as shown in this flowsheet file has to be built in for use with Edwards K-value method.

Edwards_Sour.psd – K-values Based on the Edwards, et. al. Model

K-values for weak, aqueous electrolytic solutions based on the Edwards et al., model. This method is applicable for all ions, reactions, and components which are in the Design II ionic library. Program uses the list of components for which Henry's constants in water are available

Exdis1ct.in – VLE Data Regression for Extractive Distillation of 1,3 Butadiene

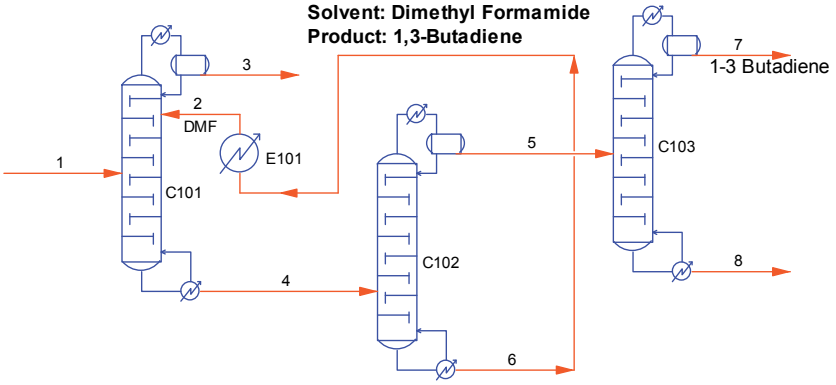
It is a ChemTran VLE data regression input file for N,N-dimethyl formamide and other species in the mixture for EXTRDIS1.PSD file. This is already built into the flowsheet file EXTRDIS1.PSD.

Exdis2ct.in – VLE Data Regression for Extractive Distillation with Phenol

It is a ChemTran VLE data regression input file for methyl cyclo hexane, toluene and phenol mixture for EXTRDIS2.PSD file. This is already built into the flowsheet file EXTRDIS2.PSD.

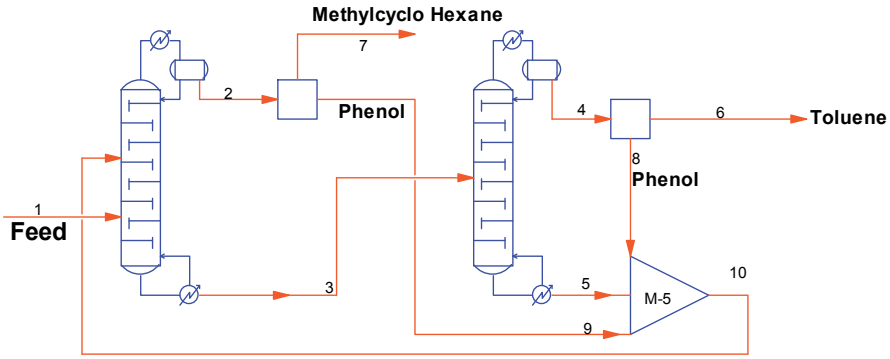
Extrdis1.psd - Extractive Distillation of 1,3 Butadiene

The sample problem is a simulation scheme for the separation of 1,3 butadiene using dimethyl formamide (DMF) as the solvent. The objective is to evaluate the purity level of 1,3 butadiene drawn off in the stream #7.



Extrdis2.psd - Extractive Distillation with Phenol

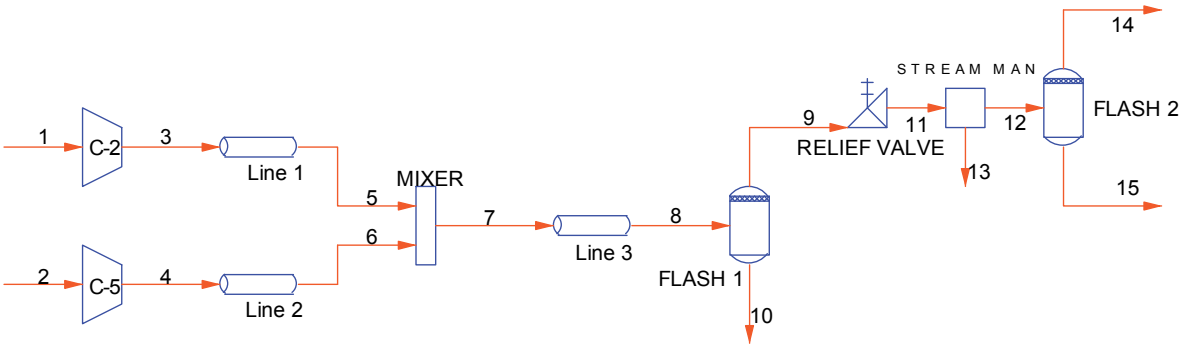
The sample problem is a simulation scheme for the separation of methylcyclohexane/toluene system using phenol as the solvent. The objective is to evaluate the amount of phenol used in order to draw off methylcyclohexane from one column and toluene in the other with sufficient purity.



About 1000 kg/hr of feed with a 50/50 mixture of toluene and methylcyclohexane is fed to the first column operating at 1 bar. The top product from the first column is a mixture of phenol and methylcyclohexane. The same amount of phenol escaping in the top product of both the columns are also accounted in the recycle to the first column via a stream manipulator and mixer module. The second column top product is pure toluene. A ChemTran regressed phase equilibrium data file for the methylcyclohexane/toluene /phenol is used for this flowsheet.

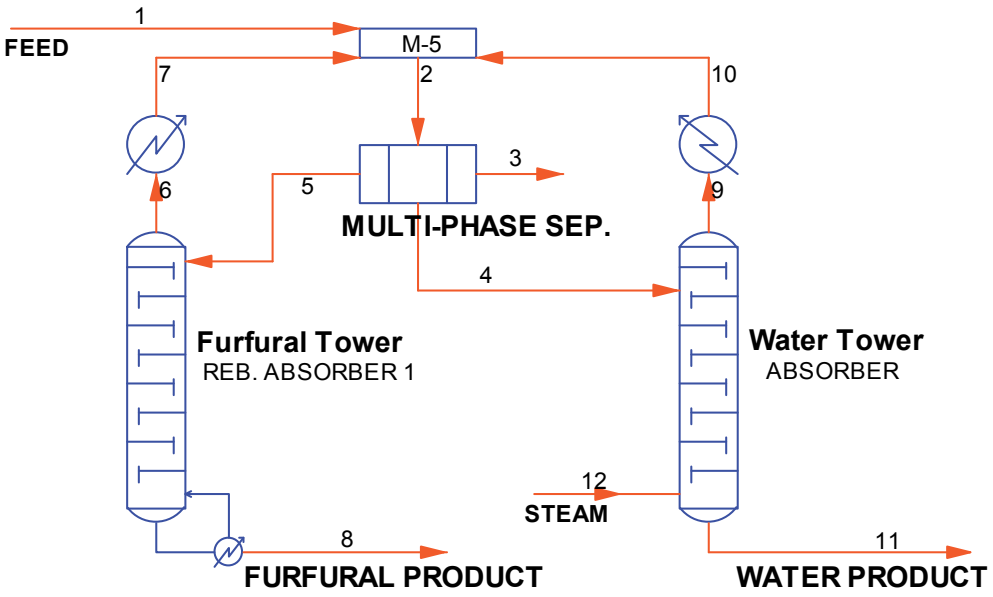
Flrehdr.psd - Flare Header Simulation

The sample simulation shows two different lines being compressed for a typical flare header relief valve sizing calculation. Methods such as API 520 and API521 relief valve sizing can be performed via the depressurizer module. The depressurizer module produces the vapor and liquid compositions relieved from the set pressure to a pre-defined reduced pressure. Ambient pressure is the default for the backpressure.



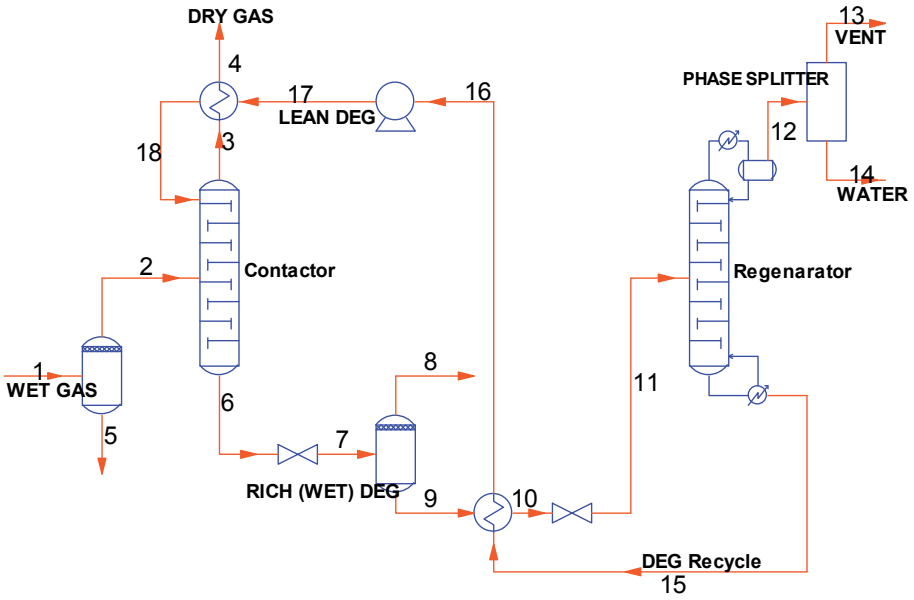
Furfural_water.psd – Fufural and Water Azeoptropic Separation

The sample problem is to illustrate the simulation of an azeotropic separation of Furfural and Water. The first distill column bottoms is pure water and the second column's bottom is pure furfural. The cooled top product from both the columns are mixed with the feed and taken through a multi-phase separator where two liquid phases separate and each stream is recycled back to the top of each column.



Glycoldh.psd - Glycol Dehydration

The sample problem is a simulation scheme for glycol/water dehydration system. The objective is to evaluate the amount of glycol recycled in order to dehydrate natural gas to sufficient purity often encountered in gas processing industry.



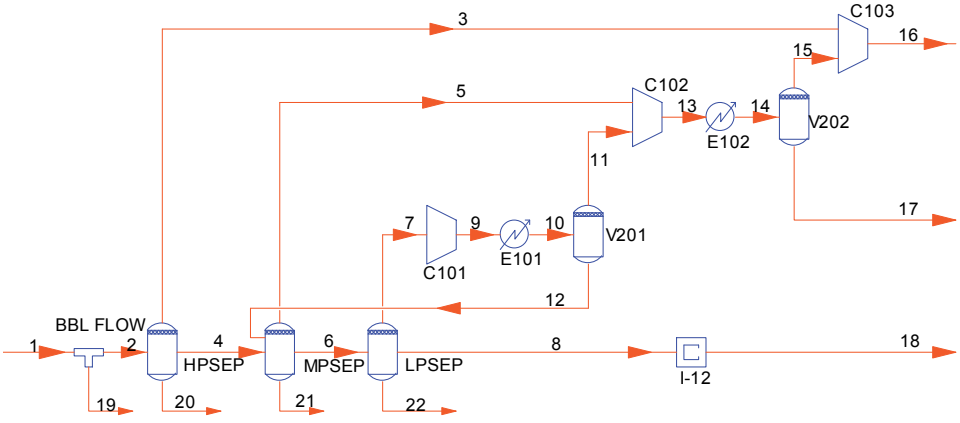
About 32450 lbmol/hr of wet natural gas is contacted counter-currently by diethylene glycol (stream 17) in the absorber module operating at 1250 psia. The wet diethylene glycol (rich in water) is passed through a valve and after flashing off trace hydrocarbons is fed to a glycol regenerator at about 14.7psia. The regenerator bottoms is the pure glycol recycled back to contact the wet gas again. A ChemTran regressed phase equilibrium data file is used for this flowsheet.

Glycwate.dat – ChemTran Data File for TEG / CO2 / Water Dehydration

It is a ChemTran created binary data file created by the ChemTran input file TGWCO2CT.IN for use by DESIGN II at runtime of TGCO2WT.PSD flowsheet file.

Gosp.psd - Gas-Oil Separation Plant

Illustration of a gas-oil separation plant with three phase flashes and Reid vapor pressure for the product. Petroleum fractions are also characterized in this problem.

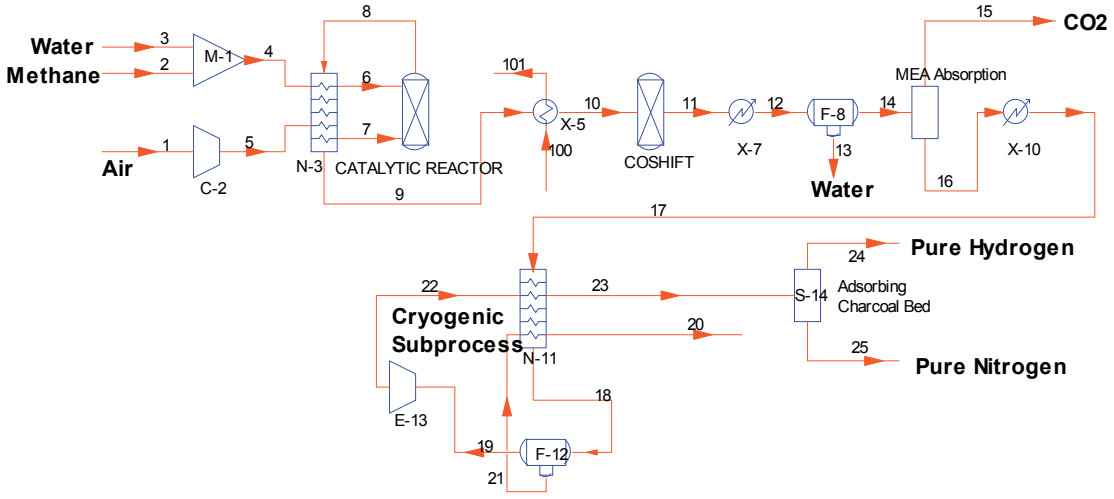


H2plant.psd - Hydrogen Production from Methane

This process is an illustration of hydrogen manufacture via a catalytic reactor. Methane, steam and air are each pre-heated to 600-700 C, and fed into a catalytic reactor containing a bed of refractory nickel catalyst (secondary reformer is used for simulation purposes). Initial combustion results in a temperature up to 1000 C. The product gas passes through heat exchange and a heat recovery boiler before entering the multi-stage CO shift converter. The gas then passes through a CO2 wash tower (can use MEA if needed in a separate flowsheet simulation...please see General: Gas Sweetening with MEA Example). Following CO2 absorption, the gas passes through a cryogenic separator where the hydrogen is separated from the CO, nitrogen, and methane.

Chapter 59

The simulation described here produces 50 MMSCF/DAY of 99.93% pure hydrogen from methane and steam.



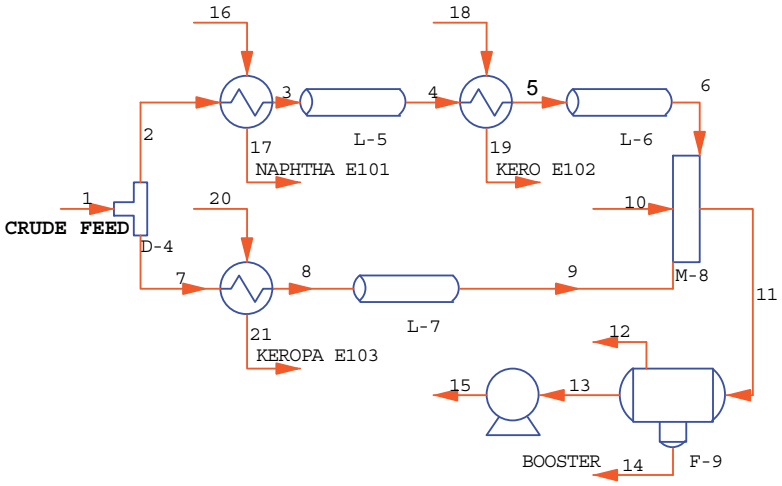

```
AB123.
*Hydrogen Production from Methane/Steam via Catalytic Reactor

C- EQUIPMENT MODULES

MIX 1 = M-1, 3,2,-4,

COMPRE 2 = C-2, 1,-5,
  PRE OUT(ATM) = 30.
  EFF = 0.72
  CAL POL
  ELE DRI
  STA = 3
  PRE STA(ATM) = 1.,10.,30.
  TEM INT(C) = 35.

LNG 3 = N-3, 4,5,8,-6,-7,-9,
 TEMperature OUT (F) 1 =1292
 TEMperature OUT (F) 2 =1292

REA 4 = CATALYTIC REACTOR, 6,7,-8,
  ADIabatic
  SECondary REFormer
  LIM=2
  CON=0.9

HEA EXC 5 = X-5, 9,100,-10,-101,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 327.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0., 0.,

REA 6 = COSHIFT, 10,-11,
  COSHIFt reaction
  TEMperature APProach COSHIFt (F) = 5

HEA EXC 7 = X-7, 11,-12,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 95.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.

FLA 8 = F-8, 12,-14,-13,
  ADI, HEA(BTU/HR) = 0.
```

```
  DEL(PSI) = 0.

COM SPL 9 = MEA Absorption, 14,-15,-16,
  REC TOP(FRA) = 5*0.,1.,0.,

HEA EXC 10 = X-10, 16,-17,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 95.
  SHE PAS = 1
  TUB PAS = 1
  SHE = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0.

LNG 11 = N-11, 22,19,17,-23,-20,-18,
  TEMperature OUT of SHEll (F) =-323.73

FLA 12 = F-12, 18,-19,-21,
  ADI, HEA(BTU/HR) = 0.
  DEL(PSI) = 0.

EXP 13 = E-13, 21,-22,
  PRE OUT(ATM) = 1.
  EFF = 0.72

COM SPL 14 = S-14, 20,-24,-25,
  REC TOP(FRA) = 2*1.,0.,4*1.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,2,46,47,48,49,62,
  APISOAVEK
  APISOAVEH
  STDD
  SUR TEN = STD
  STE TAB
  MAX = 50
  CON TOL = 1.E-003
  FLO(LBMOL/DAY)1 = 2*0.,108509.586,28840.46,3*0.,
  TP (C,ATM) 1 = 35.,1.
  NAM STR 1 = Air
  FLO(LBMOL/DAY)2 = 0.,49043.39,5*0.,
  TP (C,ATM) 2 = 35.,30.
  NAM STR 2 = Methane
  FLO(LBMOL/DAY)3 = 6*0.,98086.78,
  TP (C,ATM) 3 = 235.,30.
  NAM STR 3 = Water
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
  NAM STR 10 = Strm 10
  FLO(LBMOL/HR)11 = 5650.5,13.572,4521.2,0.,65.7,1964.,2496.2,
  TP (F,ATM) 11 = 512.82,30.
  NAM STR 11 = Strm 11
  NAM STR 12 = Strm 12
  NAM STR 13 = Water
  NAM STR 14 = Strm 14
  NAM STR 15 = Strm 15
  NAM STR 16 = Strm 16
  NAM STR 17 = Strm 17
  NAM STR 18 = Strm 18
  FLO(LBMOL/HR)19 = 5501.5,4.92E-002,293.69,0.,3.8322,2*0.,
  TP (F,ATM) 19 = -323.,30.
  NAM STR 19 = Strm 19
  NAM STR 20 = Strm 20
  NAM STR 21 = Strm 21
  FLO(LBMOL/HR)22 = 140.,13.,4223.,62.,4.5E-002,26.7,0.12057,
  TP (F,ATM) 22 = -329.81,1.
  NAM STR 22 = Strm 22
  NAM STR 23 = Strm 23
  NAM STR 24 = Strm 24
  NAM STR 25 = Strm 25
```

```
  FLO(LBMOL/HR)100 = 6*0.,3874.,
  TP (F,ATM) 100 = 90.,1.
  NAM STR 100 = Strm 100
  NAM STR 101 = Strm 101
END
```

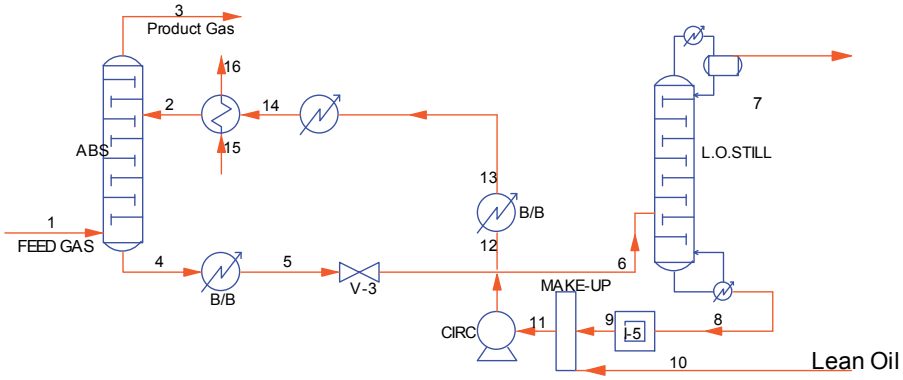
Httrain.psd - Crude Pre-Heat Train

The simulation illustrates the crude pre-heat train problem. Feed crude in stream #1 is divided and pre-heated before it reaches the booster pump for further refining. All three exchangers are being rated by the BELL method.



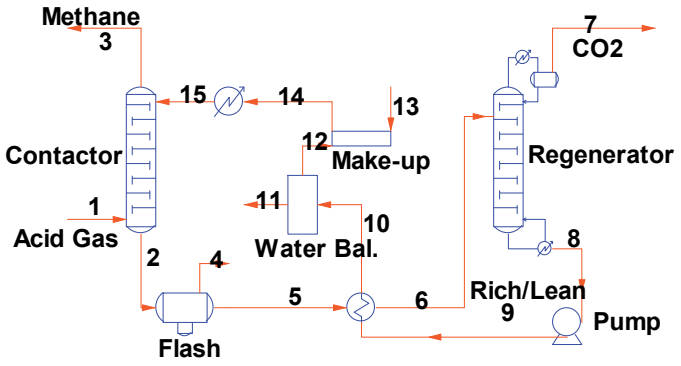
Lnoil.psd - Separation Using Lean Oil

The objective of this simulation is to use Lean Oil to absorb the heavies and strip the light ends. The lean oil is defined under the Specify...Component Data...General Properties. The feed gas is contacted by the rich lean oil at the absorber and subsequently is stripped at the Lean Oil Still for further contacting. The product gas rich in light ends (C1, through C4) leave the top of the absorber.



Meacdgs.psd – Acid Gas CO2 Stripper with MEA

The objective of this simulation is to remove CO2 and H2S from the natural gas stream by contacting the gas stream with a monoethanolamine solution. Residual H2S in the sweet gas stream is to be no greater than 1/4 grain of H2S per 100 standard cubic feet (SCF) of gas. The reboiler and condenser duties will also be determined for the amine regenerator.

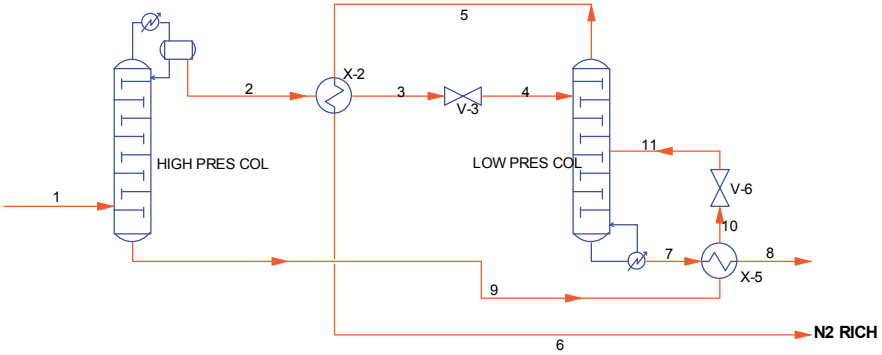


Mechtolu.dat – ChemTran Data File for Extractive Distillation with Phenol

It is a ChemTran created binary data file created by the ChemTran input file EXDIS2CT.IN for use by DESIGN II at runtime of EXTRDIS2.PSD flowsheet file.

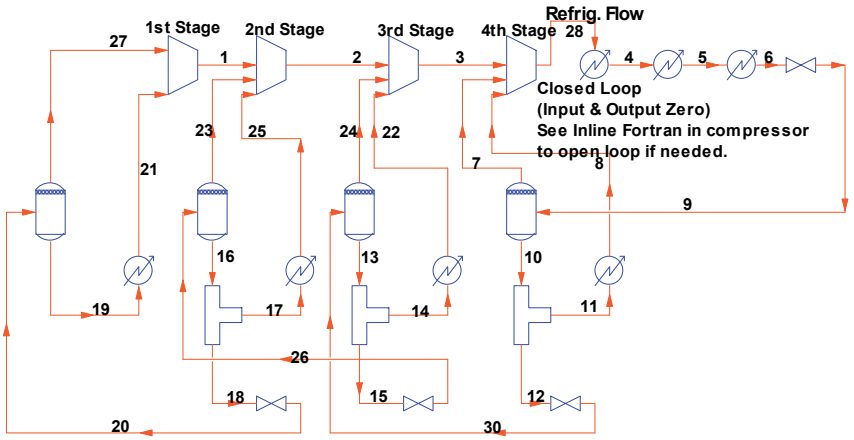
Nru.psd - Nitrogen Rejection Unit

This sample simulation illustrates a scheme for rejecting pure Nitrogen from a natural gas mixture. The flowsheet uses a high pressure (up to 377 psia) column to process a natural gas mixture feed and the top & the bottom products are fed to the low pressure (about 37 psia) column. Pure nitrogen is rejected to the top of the low pressure column.



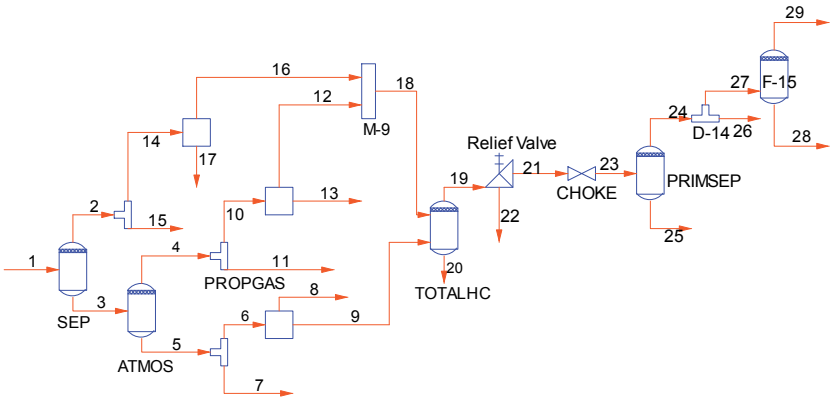
Refrign.psd - Four Stage Refrigeration

This sample problem is a "closed loop" simulation of a refrigeration system. The objective is to determine the compression achieved after four stages of an ethylene refrigeration system. The refrigerant flow of 12000 lb/hr in stream 28 is specified at known temperature and pressure and all equipment modules need to be specified. The compressors are specified at 53 psia, 70 psia, 99 psia and 290 psia in increasing order. The refrigerant is maintained at -60 F.



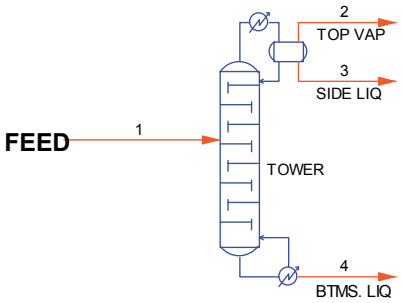
Resrvr.psd - Model for Reservoir Light Ends Depletion

In this simulation all streams represent quantities of fluid not flows. In order to obtain quantity multiply stream flow by one hour. The basic scheme is to set-the flash and stream manipulator modules with Inline Fortran so that the stream #19 (total reservoir contents at T, P) depletion rate (from 2500 to 650 psia) can be ascertained via the depressurizer.



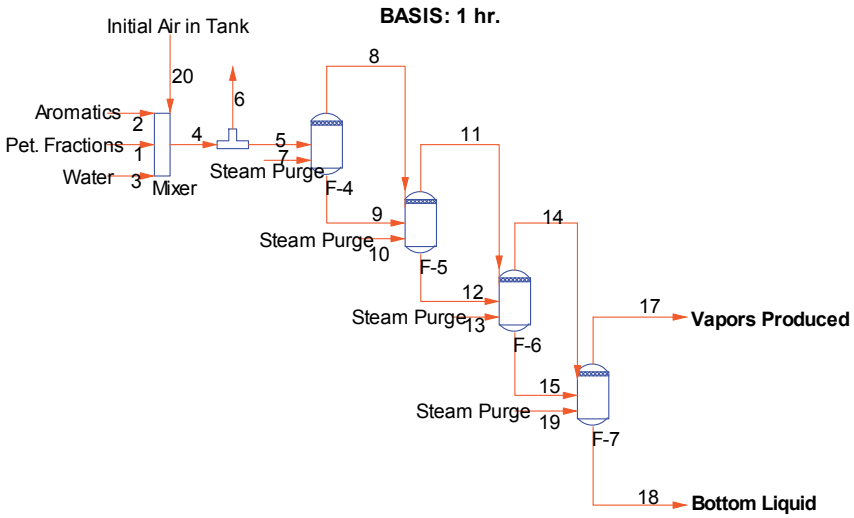
Sebsep.psd - Styrene/ Ethyl Benzene Separation

The above simulation is a 75 tray tower which separates ethyl benzene (top) and styrene (bottoms). The criteria will be to monitor whether the bottoms column temperature exceeds 221 F. Higher temperatures cause styrene to polymerize. Adequate and precise experimental vapor pressure data are required for both the components so that the prediction of the bottoms temperature is within bound.



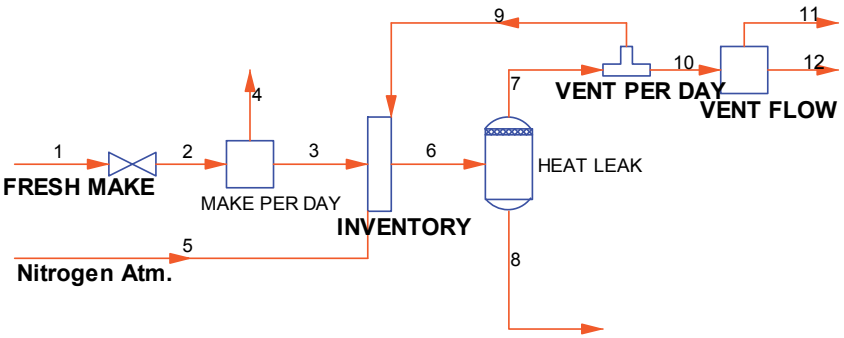
Tank1.psd - Simulation of Heating a Tank

The above scheme can compare with a time dependent heating of a tank problem via DESIGN II's steady state simulation. Each section within the tank is considered to be a steady state flash with constant heat duty supplied by the steam purge and then flashed adiabatically. The vapors and liquid produced from each flash stage represents the change in temperature that can occur during a certain interval of time.



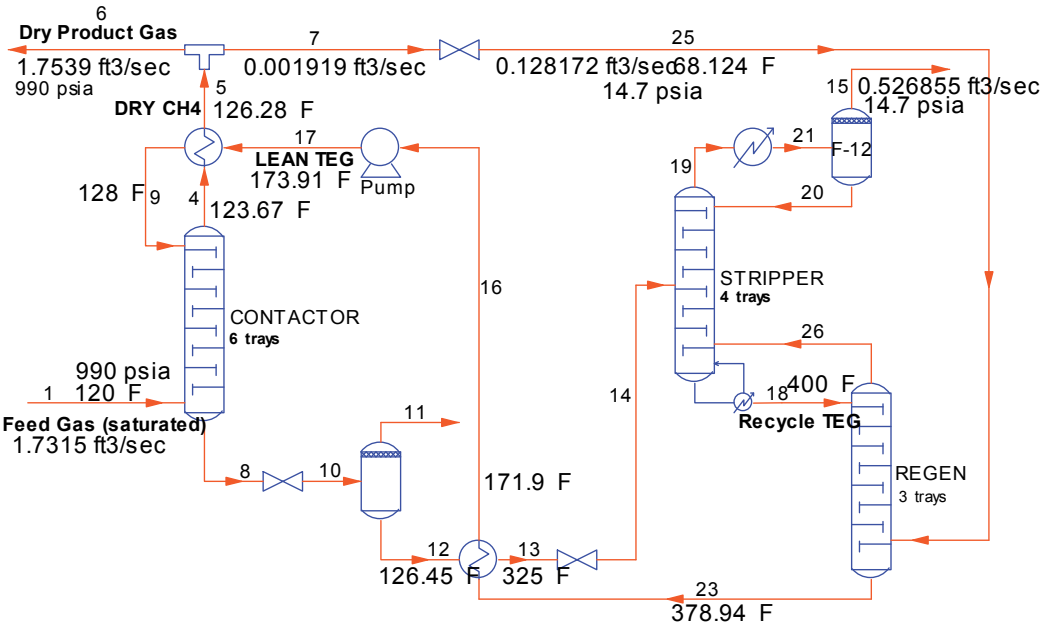
Tank2.psd - Static Tank Boiloff Model

This sample problem illustrates a static tank boiloff model at user-defined constant volume with N2 purge.



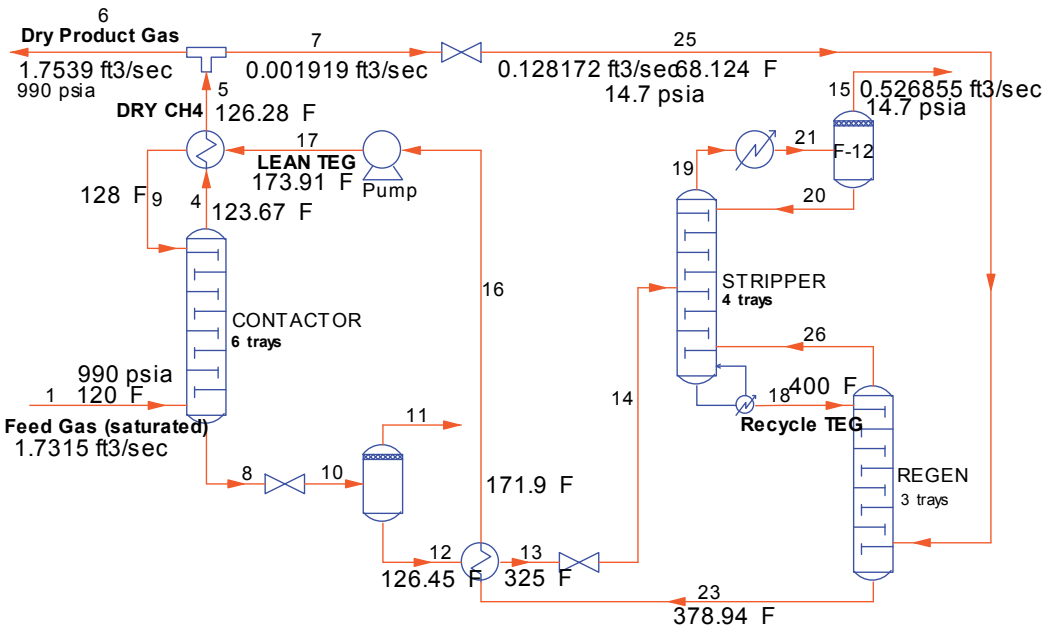
Teg.psd – Dehydration using TEG

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses a data file created via the regression of the water dewpoint of TEG / Water / Methane mixtures (reference: 5th edition of Gas Purification by Arthur Kohl) using ChemTran.



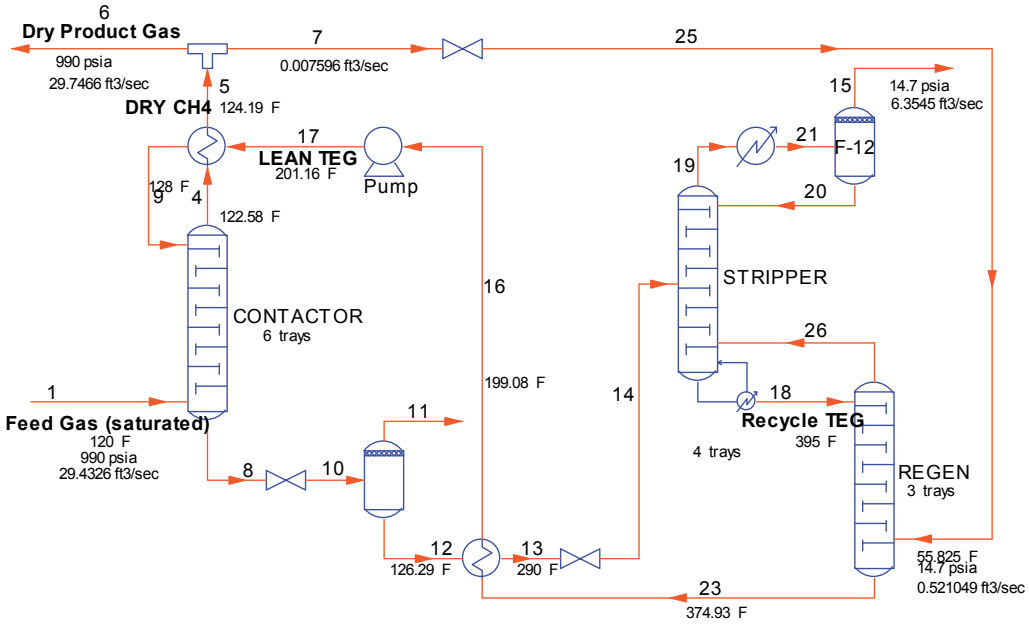
Teg_with ChemTran BIP.psd – Dehydration using TEG

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses a data file created via the regression of the water dewpoint of TEG / Water / Methane mixtures (reference: 5th edition of Gas Purification by Arthur Kohl) using ChemTran.



Teg1.psd – Dehydration using TEG

Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses the built in TEG / Water / Methane three phase Peng-Robinson mixture thermodynamics in DESIGN II.



Teg1_with inbuilt BIP.psd – Dehydration using TEG

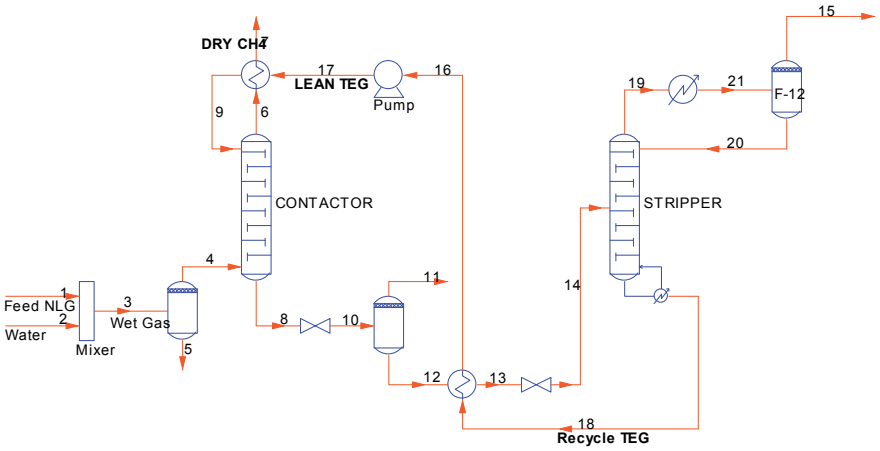
Sample simulation illustrating the use of high purity TEG for an ultra low dehydration of water saturated natural gas. This file uses the built in TEG / Water / Methane three phase Modified Peng-Robinson mixture thermodynamics in DESIGN II.

Teg1watr.dat – ChemTran Data File for Dehydration using TEG

It is a ChemTran created binary data file created by the ChemTran input file TEGCHM.IN for use by DESIGN II at runtime of TEG.PSD flowsheet file.

Tegbtex.psd - TEG/BTEX Dehydration

The sample problem is a simulation scheme for tri-ethylene glycol/water dehydration for BTEX system. The objective is to evaluate the amount of TEG recycled in order to dehydrate a mixture of Benzene, Toluene, Ethyl benzene, o-Xylene, (BTEX) and methane gas to sufficient purity often encountered in gas processing industry.



Tegbtxct.in – VLE Data Regression for TEG/BTEX Dehydration

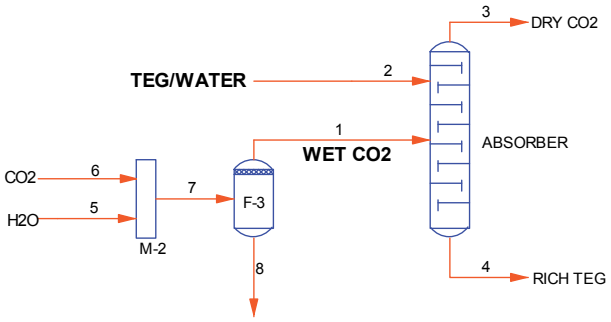
It is a ChemTran VLE data regression input file for TEG, BTEX and water mixture for TEGBTEX.PSD file. This is already built into the flowsheet file TEGBTEX.PSD.

Tegchm.in – VLE Data Regression for Dehydration using TEG

It is a ChemTran VLE data regression input file for TEG and water mixture for TEG.PSD file. This is already built into the flowsheet file TEG.PSD.

Tgco2wt.psd – TEG / CO2 / Water Dehydration

The simulation is a simple contactor to dehydrate the wet CO2 / CH4 gas. TEG is used as a dehydrating solvent counter current to the flow of gas. A ChemTran regressed data file is used for the binary interactions of water/TEG.

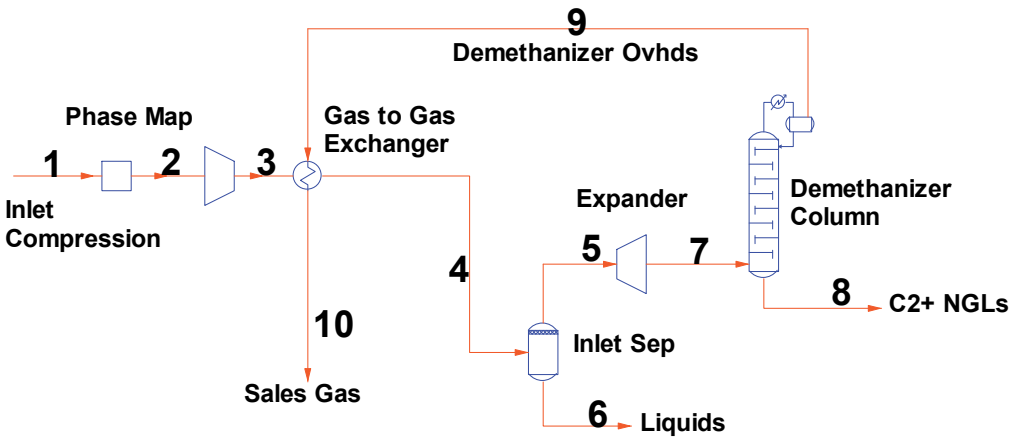


Tgwco2ct.in – VLE Data Regression for TEG / CO2 / Water Dehydration

It is a ChemTran VLE data regression input file for TEG, and water mixture for TGCO2WT.PSD file. This is already built into the flowsheet file TGCO2WT.PSD.

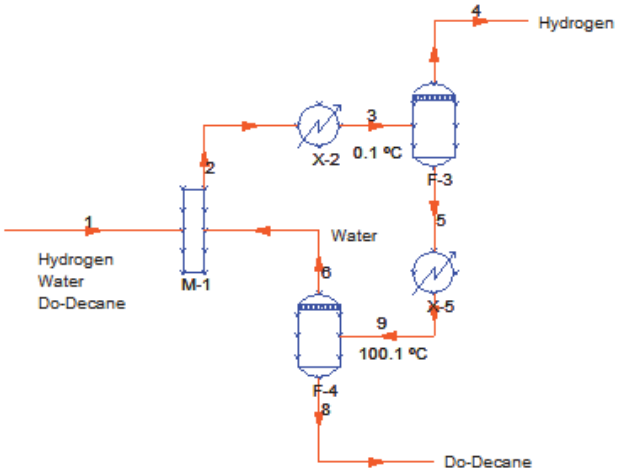
Turbo.psd – Demethanizer for Sales Gas

A natural gas mixture (from C1-C5 & CO2, H2S) is used in a demethanizer expander plant process to separate methane from other NGL gas mixture. Pure methane is cross heat exchanged with the feed mixture as the product sales gas. The bottoms from the demethanizer is free of methane with C2 through C5.



Water_trap.psd – Recycle Trapping Water

This sample illustrates how specifications on the exchanger(s) can trap a species (water) within the recycle. The recycle will never converge no matter how many iterations are given unless the exchanger specifications are modified to facilitate water to drop out.




```
AB123.
*

C- EQUIPMENT MODULES

MIX 1 = M-1, 1,6,-2,

HEA EXC 2 = X-2, 2,-3,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(C) = 0.1
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
   INL = OPP
  HOR
  DEL(PSI) = 0
  PRO 1 = SIN

FLA 3 = F-3, 3,-4,-5,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

FLA 4 = F-4, 9,-6,-8,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

HEA EXC 5 = X-5, 5,-9,
```

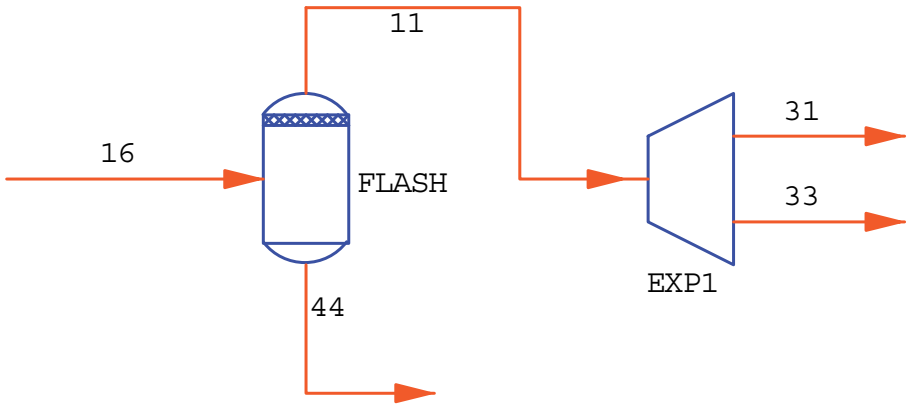
```
  U(BTU/HR/FT2/F) = 50
  TEM OUT(C) = 100.1
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 0
  PRO 1 = SIN


C- GENERAL SECTION
GENERAL
  COMPONENTS = 1,62,16
  NAM POS 1 = 'HYDROGEN'
  NAM POS 2 = 'WATER'
  NAM POS 3 = 'N-DODECANE'
C-  Thermo Comments
  PENK
  PENH
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AMB PRE(PSIA) = 14.696
  STA PRE(PSIA) = 14.696
  STA TEM(F) = 60
  NOR PRE(ATM) = 1
  BINPAR = PENG1
  EUR UNI OUT
  MAX = 200
  CON TOL = 0.001
  FLO (FRA) 1 = 0.5,0.3,0.2,
  TOT FLO (LBMOL/HR) 1 = 10
  TP (F,PSIA) 1 = 60,15
  NAM STR 1 = 'Strm 1'
  NAM STR 2 = 'Strm 2'
  NAM STR 3 = 'Strm 3'
  NAM STR 4 = 'Strm 4'
  NAM STR 5 = 'Strm 5'
  NAM STR 6 = 'Strm 6'
  NAM STR 8 = 'Strm 8'
  NAM STR 9 = 'Strm 9'
END
```

Chapter 59

Chapter 60: GERG 2008 Samples - c:\designii\samples\gerg2008

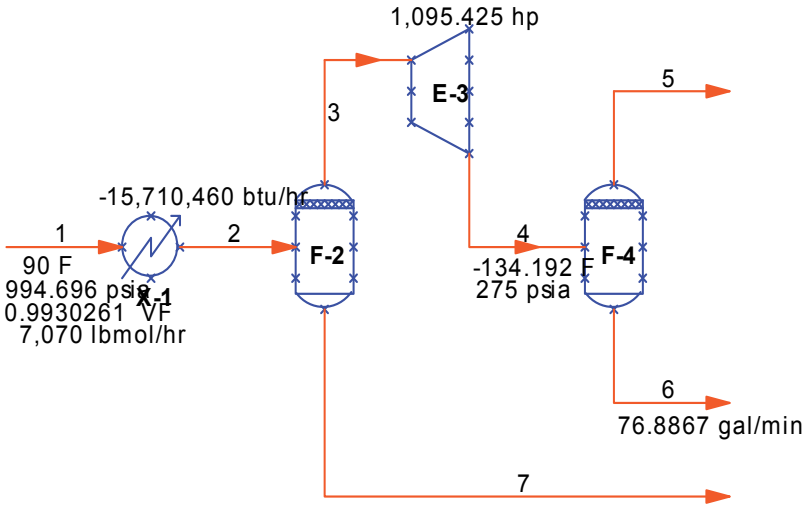
Expander.psd - Interactive Processing Example

One object of this problem is to remove enough heat from the flash drum inlet stream, 16, to recover 60% of the ethane (or 1.2 lb mol/hr) in the liquids. This is accomplished by interactively adjusting the heat removed from stream 16. The INTERACTIVE ON command is added to the FLAsh module. The initial guess for heat removed is 35500 Btu/hr.



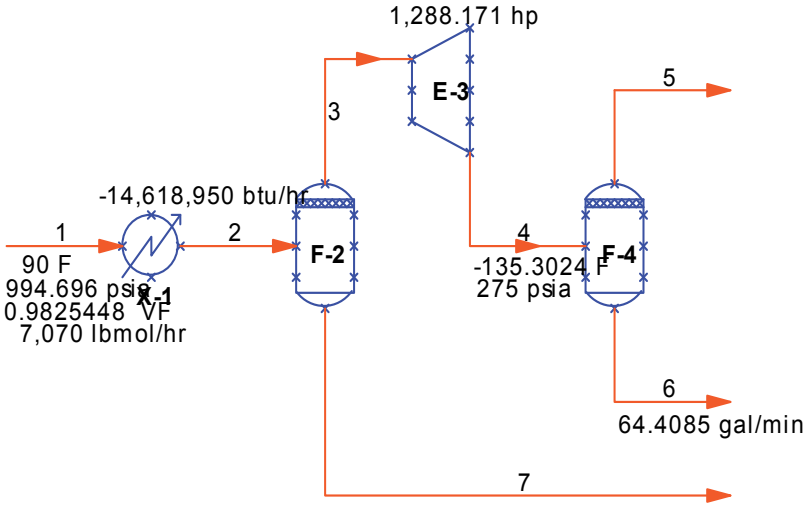
EXPANDERPeng.psd - Expander Case with Peng Robinson

The purpose of this simulation is to cool the feed stream, separate the resulting stream into two phases, and then expand the vapor phase to a lower pressure for further separations. This uses Peng-Robinson equation of state.



gerg2008.psd - Expander Case with GERG 2008

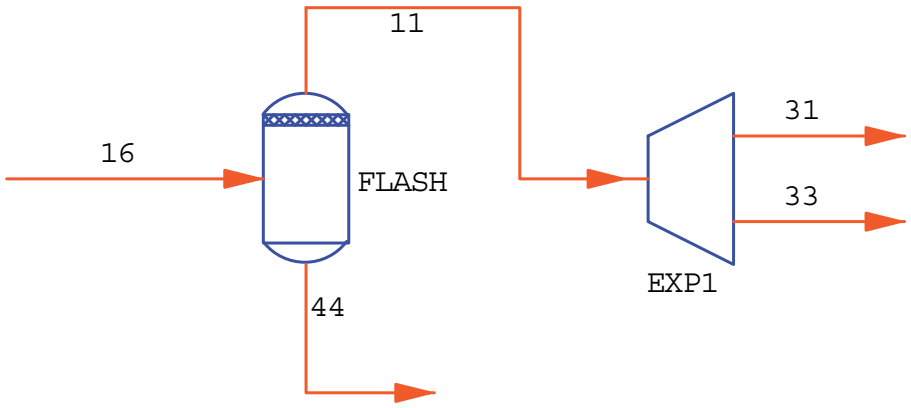
The purpose of this simulation is to cool the feed stream, separate the resulting stream into two phases, and then expand the vapor phase to a lower pressure for further separations. This uses GERG 2008 equation of state.



Chapter 61: Interactive Processing Samples - c:\designii\samples\interact

Interac.psd - Interactive Processing Example

One object of this problem is to remove enough heat from the flash drum inlet stream, 16, to recover 60% of the ethane (or 1.2 lb mol/hr) in the liquids. This is accomplished by interactively adjusting the heat removed from stream 16. The INTERACTIVE ON command is added to the FLAsh module. The initial guess for heat removed is 35500 Btu/hr.



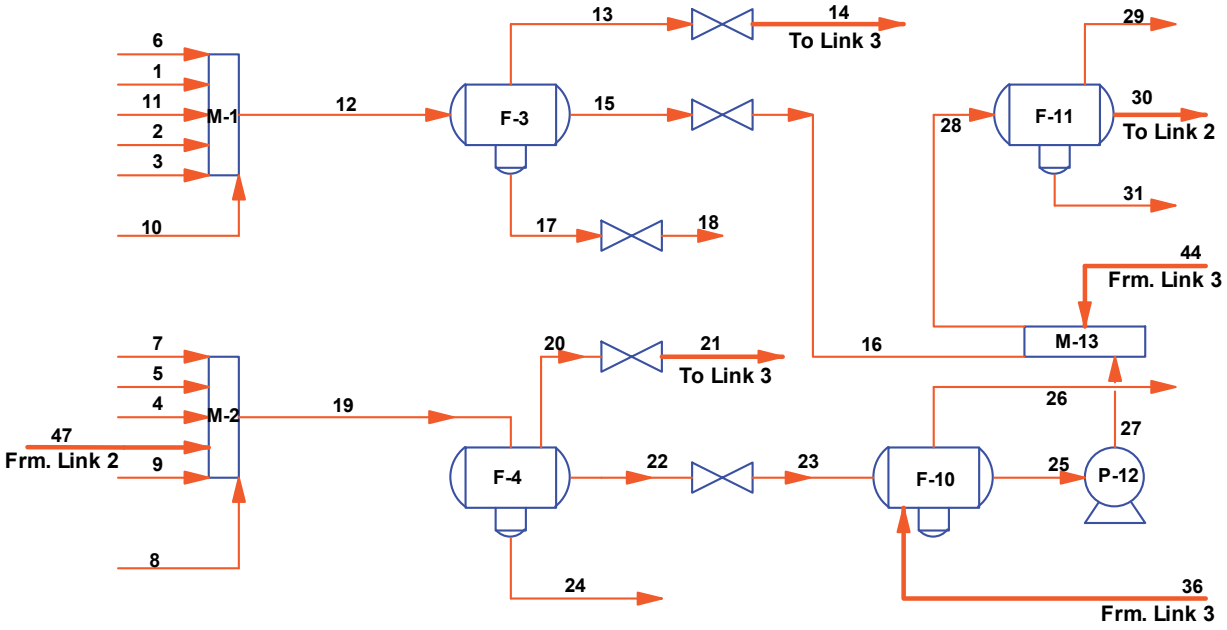

```
AB123.
*INTERACTIVE PROCESSING-EXPANDER PLANT

C- EQUIPMENT MODULES

FLA 1 = FLASH, 16,-11,-44,
  ADI, HEA(BTU/HR) = -35500
  DEL(PSI) = 0
INTERACTIVE ON

EXP 2 = EXP1, 11,-31,-33,
  PRE OUT(PSIA) = 300
  EFF = 0.72

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  KVAL
  COPE
  STDD
  DEN LIQ = STDD
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  NAM STR 11 = Strm 11
  FLO(LBMOL/HR)16 = 10,2,1,
  TP (F,PSIA) 16 = 100,800
  NAM STR 16 = Strm 16
  NAM STR 31 = Strm 31
  NAM STR 33 = Strm 33
  NAM STR 44 = Strm 44
END
```

Chapter 61

Chapter 62: Linked Flowsheet Samples - c:\designii\samples\linked

Link1.psd - Linked Flowsheet Example

All feed streams and equipment modules are specified for this assigned primary flowsheet. Streams extending from one flowsheet to the other are designated as continuation streams (via the Stream specification dialog). In this primary flowsheet, various feed streams are mixed and flashed through two sets of flash modules twice to separate the light and heavy ends from water.



Link1.unx – Linked Flowsheets

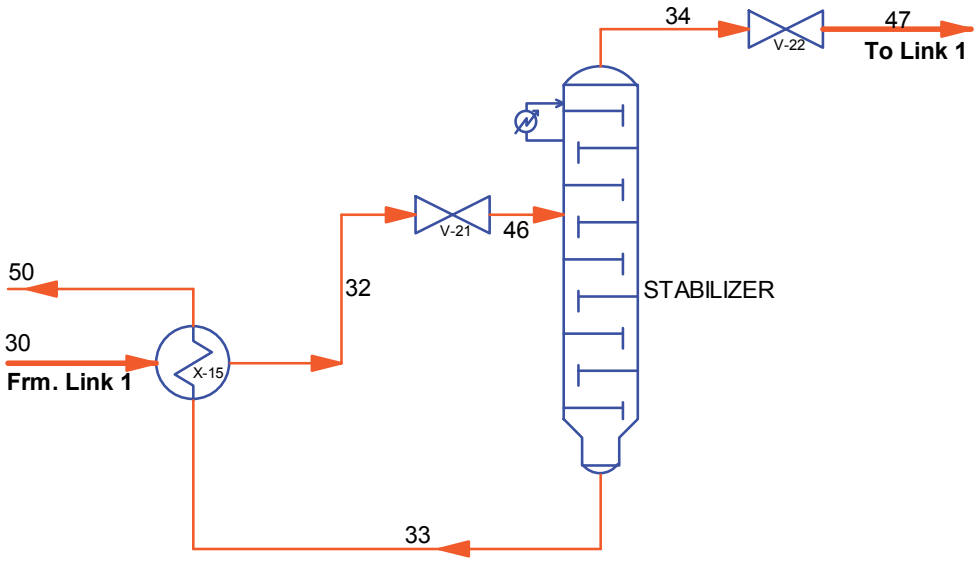
This is a links file that has information about the primary and dependent files stream links. This is an old method – please use LINKS.PSD which is an easier and newer method to linking sheets within a file.

Link2.in – Stream 30 from Link1.psd

This is an input file created from the dependent LINK2.PSD flowsheet file.

Link2.psd - Stream 30 from Link1.psd

The feed for this flowsheet is stream 30 from Link1.PSD. This stream is assigned as a continuation stream and is preheated via a heat exchanger before being processed in the Refine Column. The overhead stream 47 is connected to Mixer-2 in the Link1.PSD for continued processing. The bottoms from the Refine are the heavy end products.



Link3.in – Stream 21 from Link1.psd

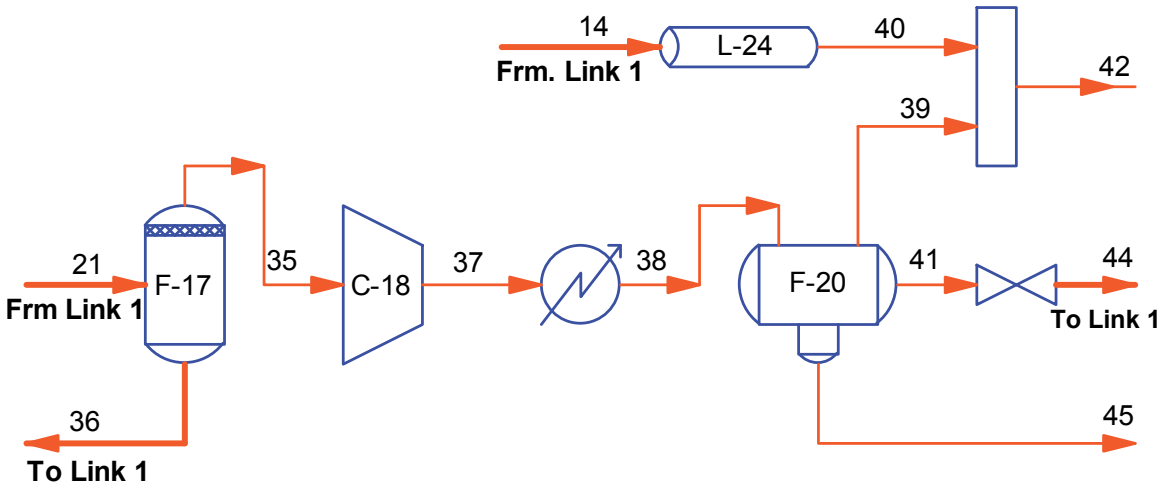
This is an input file created from the dependent LINK3.PSD flowsheet file.

Link3.psd - Stream 21 from Link1.psd

The stream 21 with light end vapors continuing from Link1.PSD, is further flashed, compressed to a higher pressure (725 psia) and cooled by a heat exchanger. It is subsequently flashed (F-20) and the liquids in stream 44 are taken for re-processing in Link1.PSD. The stream 14 from Link1.PSD is processed through a line module and mixed along with the vapors of the flash (F-20) to purge with flare. Liquids (stream 36) flashed out from stream 21 are also returned to Link1.PSD.

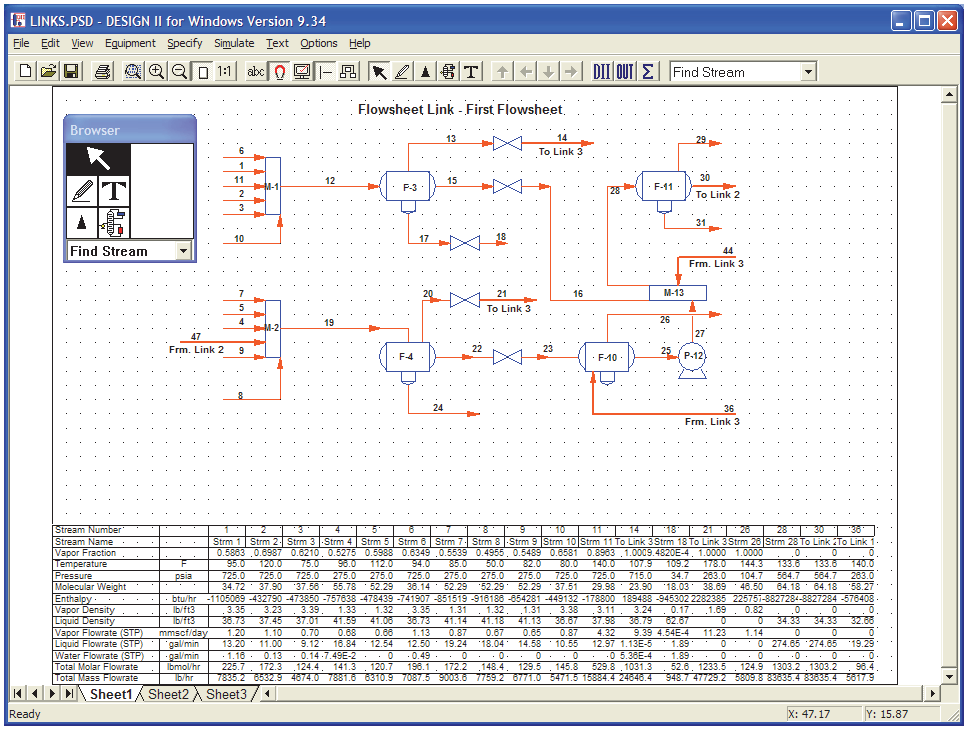
Notes:

1. If new equipments and streams are added to a dependent flowsheet after defining the flowsheet link, DESIGN II will automatically assign each stream or equipment a unique number in the set of linked flowsheets.
2. Only those streams which cross the drawing boundary to another linked flowsheet are allowed to have duplicate numbers. To maintain the continuity of a stream which extends to another flowsheet, change the current stream number on the flowsheet where the stream appears as a feed.



Links.psd - Linked Flowsheet Example

Links1.psd plus Links2.psd plus Links3.psd in one flowsheet on three different sheets. Each sheet links via the connected streams present in other sheets.

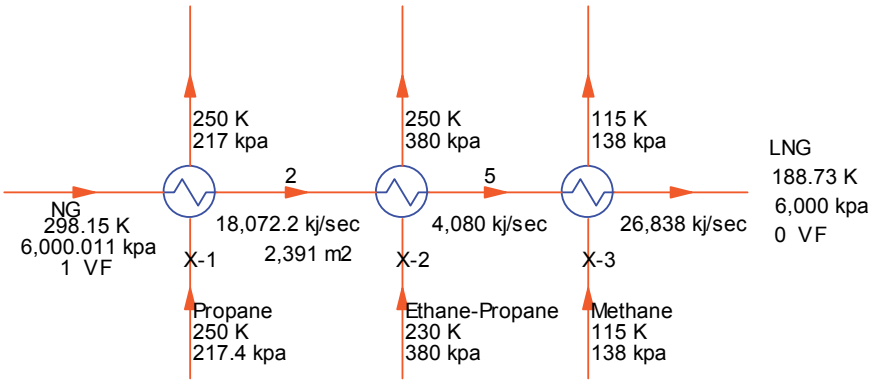


Stream Number		1	2	3	4	5	6	7	8	9	10	11	14	18	21	26	28	30	35
Stream Name		Strm 1	Strm 2	Strm 3	Strm 4	Strm 5	Strm 6	Strm 7	Strm 8	Strm 9	Strm 10	Strm 11	To Link 3	Strm 18	To Link 3	Strm 26	Strm 28	To Link 2	To Link 1
Vapor Fraction		0.5883	0.6967	0.6210	0.5275	0.5888	0.6349	0.5539	0.4955	0.5489	0.6591	0.8963	1.0009	4.820E-4	1.0000	1.0000	0	0	0
Temperature	F	95.0	120.0	75.0	95.0	112.0	94.0	85.0	50.0	82.0	80.0	140.0	107.5	109.2	178.0	144.3	133.6	133.6	140.0
Pressure	psia	725.0	725.0	725.0	275.0	275.0	725.0	275.0	275.0	275.0	725.0	725.0	715.0	34.7	263.0	104.7	564.7	564.7	263.0
Molecular Weight		34.72	37.90	37.56	55.78	52.29	36.14	52.29	52.29	52.29	37.51	29.98	23.90	18.03	38.69	46.50	64.18	64.18	58.27
Enthalpy	btu/hr	-1105089	-432790	-473850	-757638	-476439	-741907	-851519	-916186	-554281	-449132	-178800	189488	-945302	2282385	225751	-882728	-8827284	-576408
Vapor Density	lb/ft3	3.35	3.23	3.39	1.33	1.32	3.35	1.31	1.32	1.31	3.38	3.11	3.24	0.17	1.69	0.82	0	0	0
Liquid Density	lb/ft3	36.73	37.45	37.01	41.59	41.06	36.73	41.14	41.18	41.13	38.67	37.98	36.75	62.67	0	0	34.33	34.33	32.66
Vapor Flowrate (STP)	mmscf/day	1.20	1.10	0.70	0.68	0.66	1.13	0.87	0.67	0.65	0.87	4.32	9.39	4.54E-4	11.23	1.14	0	0	0
Liquid Flowrate (STP)	gal/min	13.20	11.00	9.12	16.84	12.54	12.50	19.24	18.04	14.58	10.55	12.97	1.13E-5	1.89	0	0	274.65	274.65	19.29
Water Flowrate (STP)	gal/min	1.16	0.13	0.14	7.49E-2	0	0.49	0	0	0	0	0	5.38E-4	1.89	0	0	0	0	0
Total Molar Flowrate	lbmol/hr	225.7	172.3	124.4	141.3	120.7	196.1	172.2	148.4	129.5	145.8	529.8	1031.3	52.6	1233.5	124.9	1303.2	1303.2	96.4
Total Mass Flowrate	lb/hr	7835.2	6532.9	4674.0	7861.6	6310.9	7087.5	9003.6	7759.2	6771.0	5471.5	15864.4	24646.4	948.7	47729.2	5809.8	83635.4	83635.4	5617.9

Chapter 63: LNG Samples - c:\designii\samples\lng

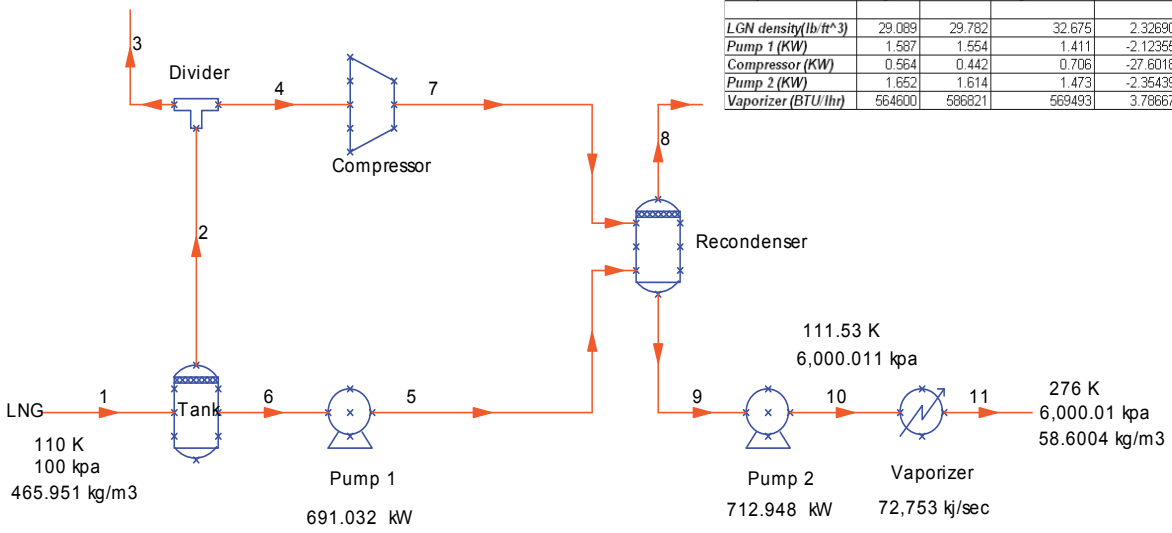
Lng_liquifaction_gerg2004.psd – conversion of natural gas to LNG using the GERG 2004 thermodynamic method

An example showing the liquefaction of natural gas to LNG using heat exchangers and GERG thermodynamic method.



Lng_vaporization_gerg2004.psd – conversion of LNG to natural gas using the GERG 2004 thermodynamic method

A more detailed scheme to re-convert LNG to natural gas.

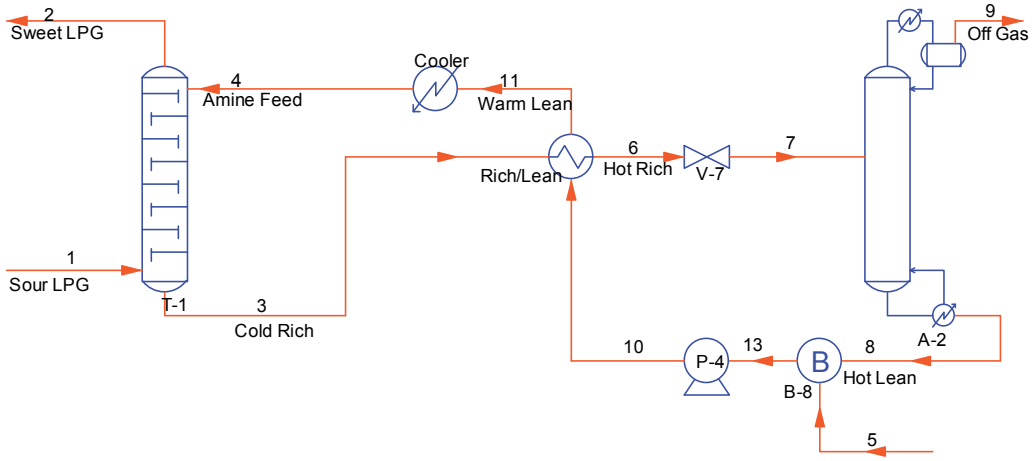


Components	Gerg2004	API-Soave	Peng-Robinson	%Deviation API	%Deviation PR
LGN density(lb/ft^3)	29.089	29.782	32.675	2.326908871	10.97475134
Pump 1 (KW)	1.587	1.554	1.411	-2.123552124	-12.4734231
Compressor (KW)	0.564	0.442	0.706	-27.60180995	20.11331445
Pump 2 (KW)	1.652	1.614	1.473	-2.354399009	-12.1520706
Vaporizer (BTU/hr)	564600	586821	569493	3.786674301	0.859185275

Chapter 64: LPG Treating Samples - c:\designii\samples\lpg

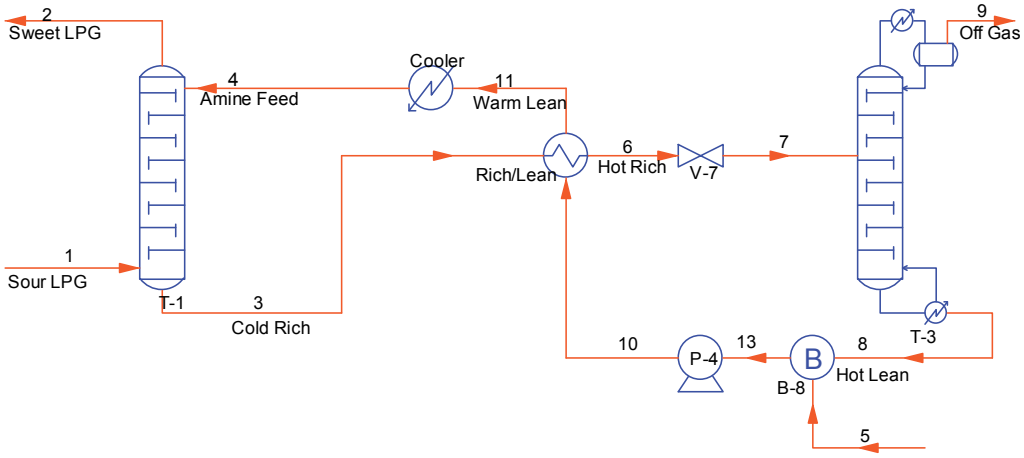
Sourlpg_dea_1.psd – removal of CO2 from sour LPG using DEA

Sour LPG is fed into liquid-liquid extractror column using the MEA/DEA thermodynamic method. The rich DEA is then recycled in a Mixed Amine Regenerator column using the Mixed Amine thermodynamic method. Water loss from the tops of the columns for the cycle is automatically calculated by the Mass Balance unit module. The mass balance unit module maintains 44 gpm flowrate in the recycle.



Sourlpg_dea_2.psd – removal of CO2 from sour LPG using DEA

Sour LPG is fed into liquid-liquid extractror column using the MEA/DEA thermodynamic method. The rich DEA is then recycled in a Regenerator column using the MEA / DEA thermodynamic method. Water loss from the tops of the columns for the cycle is automatically calculated by the Mass Balance unit module. The mass balance unit module maintains 44 gpm flowrate in the recycle.

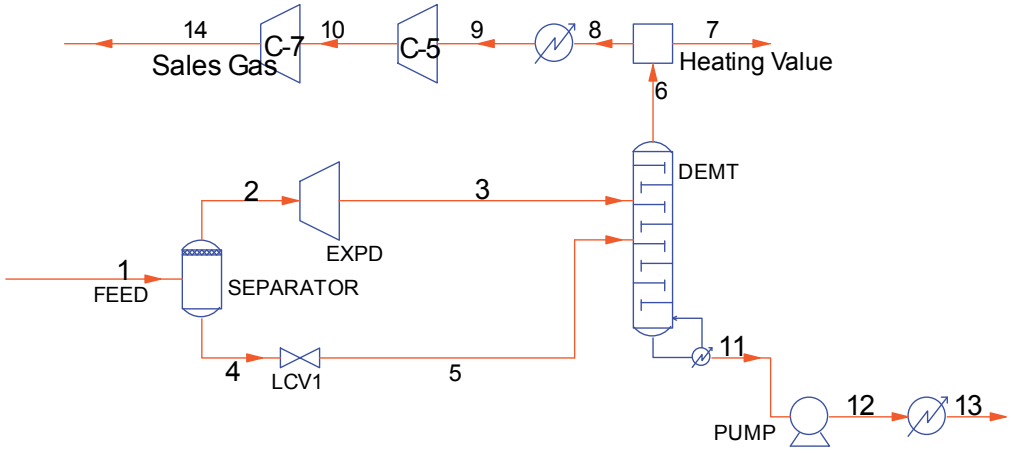


Chapter 64

Chapter 65: Optimization Samples - c:\designii\samples\optimize

Optimiz1.psd - Flowsheet Optimization of an Expander Plant

This optimization problem looks at one section of a gas plant as shown in the flowsheet below. The goal of this problem is to minimize the sales gas recompression work. The design variables are the FLAsh 2 temperature and the outlet pressure from EXPander 3. The problem constraint specifies that at least 87% of the ethane fed to the FLAsh 2 cold separator is to be recovered in the DIStillation 6 bottom product (stream 11). The heating value of the sales gas is also of importance, so it will be calculated in STReam MANipulator 11.




```
AB123.
*Assignment 1: Expander Plant - Optimization to minimize recompression HP

C- EQUIPMENT MODULES

HEA EXC 1 = IGEX, 1,-2,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = -45.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 10.

FLA 2 = SEPARATOR, 2,-3,-4,
  ISO, TEM OUT(F) = -45.
  PRE OUT(PSIA) = 985.

EXP 3 = EXPD, 3,-5,
  PRE OUT(PSIA) = 300.
  EFF = 0.8
C-
C-  USE INLINE FORTRAN TO SET PRESSURES FOR COLUMN AND VALVE
C-  PASS EXPANDER HORSEPOWER TO COMPRESSOR 7, SET TEM TOP
C-  FOR COLUMN EQUAL TO OUTLET TEMP. FOR EXPANDER
C-
F-      POST
F-      SET(VAL(4),PRE OUT) = GET(EXP(3),PRE OUT) + 2
F-      SET(COMPRE(5),WOR AVA) = GET(EXP(3),CAL WOR)*.98
F-

VAL 4 = LCV1, 4,-6,
  PRE OUT(PSIA) = 302.

HEA EXC 8 = GSEX, 7,-8,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 110.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 10.

PUM 10 = PUMP, 11,-12,
```

```
  WOR AVA(HP) = 500.
  PRE OUT(PSIA) = 500.
  ENT,   EFF = 0.9
  FUEL

DIS 6 = DEMT, 5,6,-7,-11,
  ABS REB
  TRA = 10
  TEM TOP(F) = -140.
  TEM BOT(F) = 25.
  PRE TOP(PSIA) = 300.
  DEL(PSI) = 6.
  LOC FEE = 1,3,
  PRO (LBMOL/HR) = 0.,1000.
  SUP
  RAT BOT (MOL, REL) 2,3 = 1.5E-002
  LOC HEA = 5,
  HEA(BTU/HR) = 4250000.,
  TEM PRO(F) = -120.,0,0,0,-90.,-60.,-32.,0,0,5.,
     CO2 FREeze

COMPRE 5 = C-5, 8,-9,
  WOR AVA(HP) = 1500.
  PRE OUT(PSIA) = 700.
  ENT, EFF = 0.78
  FUEL
  STA = 1

COMPRE 7 = C-7, 9,-10,
  PRE OUT(PSIA) = 950.
  ENT, EFF = 0.7
  FUEL
  STA = 1

HEA EXC 9 = X-9, 12,-13,
  U(BTU/HR/FT2/F) = 50.
  TEM OUT(F) = 70.
  SHE PAS = 1
  TUB PAS = 1
  DEL(PSI) = 10.

STR 11 = Z-11, 10,-15,-14,
  HEA VAL

C- FLOWSHEET OPTIMZATION
OPTIM
  VARY (-70 .LE. (HEA EXC(1),TEM OUT) .LE. -40,SCA=1, PER REL=0.02), AND
  (250.LE.(EXP(3),PREOUT).LE.500, SCA=5)
C-
C- C2 RECOVERY AS FRACTION OF FEED IS CONSTRAINT
C-
  SUBJECT TO CON(1) .GE. 0.87 (MUL=1000)
  UNTIL FUN IS MINIMIZED
C-
C- MINIMIZE RECOMPRESSION HORSEPOWER
C-
F-      CON(1)=GET(STR(11),FLO COM(4))/GET(STR(1),FLO COM(4))
F-      FUN = -1*GET(COMPRE(7),CAL WORK)
C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,
  APISOAVEK
  APISOAVEH
  STDD
  MAX = 1
  CON TOL = 1.E-003
  FLO(LBMOL/HR)1 = 110.4,137.69,6131.42,500.47,193.28,80.36,66.95,29.77,18.65,
  31.29,
  TP (F,PSIA) 1 = 90.,995.
END
```

Any FORTRAN syntax or compilation errors will be reported immediately after the "ECHO PRINT OF INPUT DATA". The optimization results will be reported in the "NOW CALLING" section of the output.

Selected portions of the results for this example are shown on the following pages. The intermediate results may vary from one computer system to another due to differences in precision.

```
+------------------------ OPTIMIZATION ITERATION   1 ------------------------+
```

```
+                                                                              +
+ OBJECTIVE FUNCTION VALUE:                    4470.54     IN INTERNAL UNITS +
+                                                                              +
+ VAR   1 -- (HEAEXC(1),TEMOUT)                414.670     IN INTERNAL UNITS +
+            LOWER BOUND:                      389.670     (INSIDE  BOUND)   +
+            UPPER BOUND:                      419.670     (INSIDE  BOUND)   +
+ VAR   2 -- (EXP(3),PREOUT)                   300.000     IN INTERNAL UNITS +
+            LOWER BOUND:                      250.000     (INSIDE  BOUND)   +
+            UPPER BOUND:                      500.000     (INSIDE  BOUND)   +
+                                                                              +
+                 +++ CONSTRAINT SUMMARY (LHS MINUS RHS) +++                   +
+                                                                              +
+ CONSTRAINT   1 (INEQUALITY):                -216.850     NOT SATISFIED     +
+------------------------------------------------------------------------------+
+------------------------ OPTIMIZATION ITERATION   2 -------------------------+
+                                                                              +
+ OBJECTIVE FUNCTION VALUE:                    4233.06     IN INTERNAL UNITS +
+                                                                              +
+ VAR   1 -- (HEAEXC(1),TEMOUT)                414.645     IN INTERNAL UNITS +
+            LOWER BOUND:                      389.670     (INSIDE  BOUND)   +
+            UPPER BOUND:                      419.670     (INSIDE  BOUND)   +
+ VAR   2 -- (EXP(3),PREOUT)                   250.000     IN INTERNAL UNITS +
+            LOWER BOUND:                      250.000     (****ON  BOUND)   +
+            UPPER BOUND:                      500.000     (INSIDE  BOUND)   +
+                                                                              +
+                 +++ CONSTRAINT SUMMARY (LHS MINUS RHS) +++                   +
+                                                                              +
+ CONSTRAINT   1 (INEQUALITY):                -72.5669     NOT SATISFIED     +
+------------------------------------------------------------------------------+

+------------------------ OPTIMIZATION ITERATION   3 -------------------------+
+                                                                              +
+ OBJECTIVE FUNCTION VALUE:                    4232.68     IN INTERNAL UNITS +
+                                                                              +
+ VAR   1 -- (HEAEXC(1),TEMOUT)                413.943     IN INTERNAL UNITS +
+            LOWER BOUND:                      389.670     (INSIDE  BOUND)   +
+            UPPER BOUND:                      419.670     (INSIDE  BOUND)   +
+ VAR   2 -- (EXP(3),PREOUT)                   250.000     IN INTERNAL UNITS +
+            LOWER BOUND:                      250.000     (****ON  BOUND)   +
+            UPPER BOUND:                      500.000     (INSIDE  BOUND)   +
+                                                                              +
+                 +++ CONSTRAINT SUMMARY (LHS MINUS RHS) +++                   +
+                                                                              +
+ CONSTRAINT   1 (INEQUALITY):                -72.5663     NOT SATISFIED     +
+------------------------------------------------------------------------------+

+------------------------ FINAL OPTIMIZATION RESULTS -------------------------+
+                                                                              +
+               +++ LINE SEARCH GIVES NO BETTER SOLUTION +++                   +
+                                                                              +
+ OBJECTIVE FUNCTION VALUE:                    4232.68     IN INTERNAL UNITS +
+                                                                              +
+ VAR   1 -- (HEAEXC(1),TEMOUT)                413.943     IN INTERNAL UNITS +
+            LOWER BOUND:                      389.670     (INSIDE  BOUND)   +
+            UPPER BOUND:                      419.670     (INSIDE  BOUND)   +
+ VAR   2 -- (EXP(3),PREOUT)                   250.000     IN INTERNAL UNITS +
+            LOWER BOUND:                      250.000     (****ON  BOUND)   +
+            UPPER BOUND:                      500.000     (INSIDE  BOUND)   +
+                                                                              +
+                 +++ CONSTRAINT SUMMARY (LHS MINUS RHS) +++                   +
+                                                                              +
+ CONSTRAINT   1 (INEQUALITY):                -72.5663     NOT SATISFIED     +
+------------------------------------------------------------------------------+

                   +++++ END OF OPTIMIZATION LOOP +++++

                              STREAM SUMMARY
                              --------------
STREAM NUMBER            1             2             3             4
STREAM NAME       FEED           Strm 2         Strm 3         Strm 4
EQUIP CONXION        FEED-IGEX      IGEX-SEPA      SEPA-EXPD      SEPA-LCV1
                    (   0)-(   1)  (   1)-(   2)  (   2)-(   3)  (   2)-(   4)
VAPOR FRACTION        1.0000         0.79924        1.0000         0.00000E+00
TEMPERATURE F         90.000         -45.727        -45.000        -45.000
PRESSURE PSIA         995.00         985.00         985.00         985.00
ENTHALPY BTU/HR    -0.10453E+07   -0.18712E+08   -0.10963E+08   -0.76050E+07
V Cp  BTU/LBMOL/R     12.731         21.568         21.343
```

V Cv	BTU/LBMOL/R	8.3070	7.1981	7.2044	
L Cp	BTU/LBMOL/R		21.100		21.118
V DEN	LB/FT3	4.0257	6.7983	6.7577	
L DEN	LB/FT3		25.736		25.823
L S.G.(60F)	STP		0.40825		0.40942
V VIS	CP	0.13634E-01	0.13041E-01	0.13021E-01	
L VIS	CP		0.73544E-01		0.74209E-01
V THC	BTU/FT/HR/F	0.23816E-01	0.23061E-01	0.23024E-01	
L THC	BTU/FT/HR/F		0.62593E-01		0.62733E-01
V MMSCF/DAY	STP	66.478	53.132	53.408	
L GAL/MIN	STP		197.96		194.41
TOTAL MOLE WEIGHT		19.977	19.977	18.078	27.739

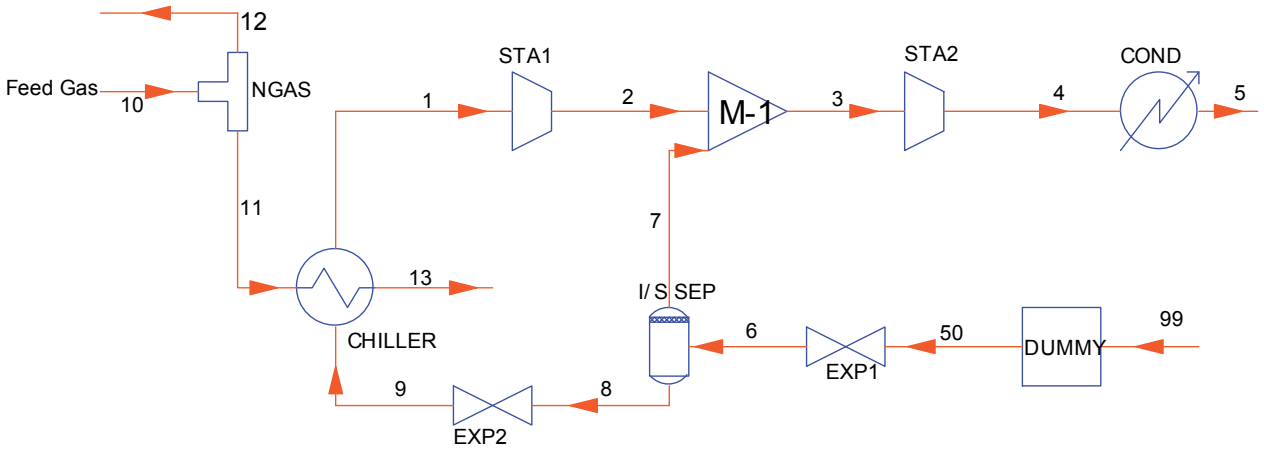
L-LIQUID PHASE (EXCLUDING IMMISCIBLE WATER), V-VAPOR PHASE
Cp,Cv-HEAT CAPACITY, DEN-DENSITY, VIS-VISCOSITY, THC-THERMAL CONDUCTIVITY

FLOW RATES LBMOL/HR

NITROGEN	110.40	110.40	103.36	7.0413
CO2	137.69	137.69	100.77	36.921
METHANE	6131.4	6131.4	5248.9	882.53
ETHANE	500.47	500.47	304.65	195.82
PROPANE	193.28	193.28	71.244	122.04
I-BUTANE	80.360	80.360	18.313	62.047
N-BUTANE	66.950	66.950	12.204	54.746
I-PENTANE	29.770	29.770	2.9667	26.803
N-PENTANE	18.650	18.650	1.4832	17.167
N-HEXANE	31.290	31.290	1.0520	30.238
TOTAL LBMOL/HR	7300.3	7300.3	5864.9	1435.4
TOTAL LB/HR	0.14584E+06	0.14584E+06	0.10602E+06	39816.

Optimiz2.psd - Two Stage Refrigeration Cycle Optimization

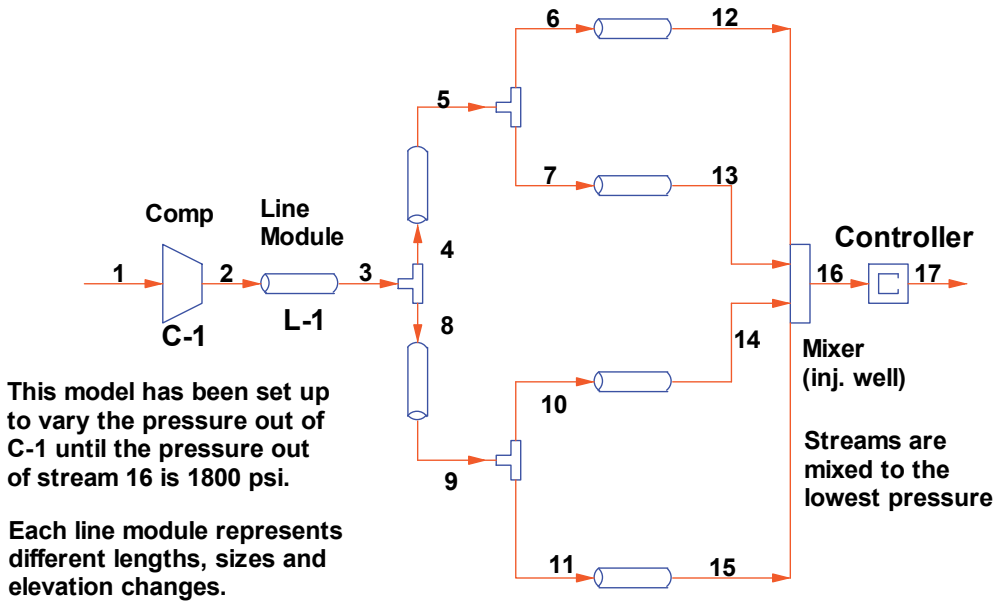
The objective of this simulation is to optimize the temperature achieved at the chiller by maximizing the tube side flow to the chiller. This is accomplished by varying the divider flowrate constrained on two conditions. Both the conditions will make sure a minimum temperature difference of 0.5 between stream # 13 and # 9.



Chapter 66: Pipeline Samples - c:\designii\samples\pipeline

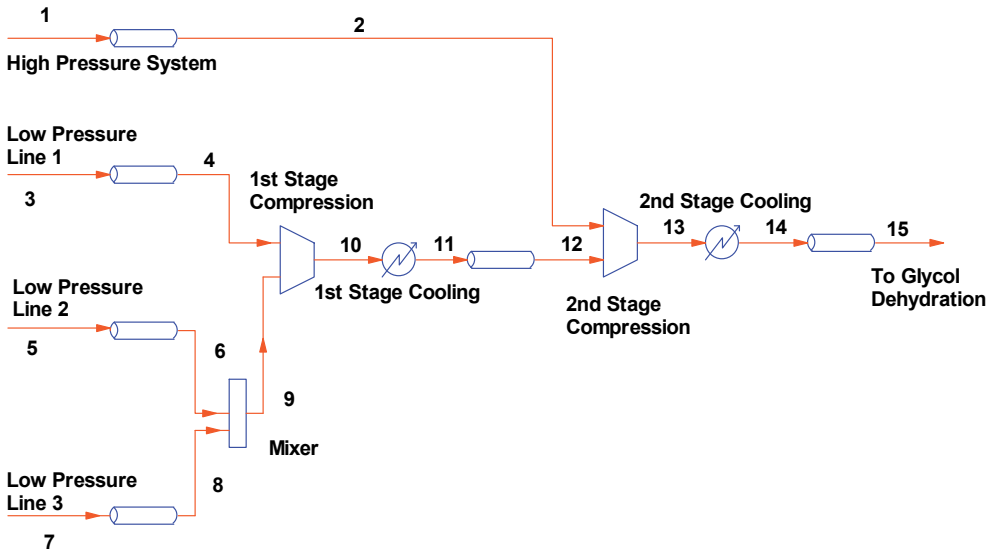
Co2injn.psd - Back Pressure Calculation For A CO2 Injection Field

This example simulation has been set up to vary the pressure out of Compressor 1 until the pressure in the stream 16 is 1800 psi. The variables to be varied and converged are set-up within the controller module. Each line module represents different lengths, sizes and elevation changes. Streams at the Mixer are mixed to the lowest pressure.



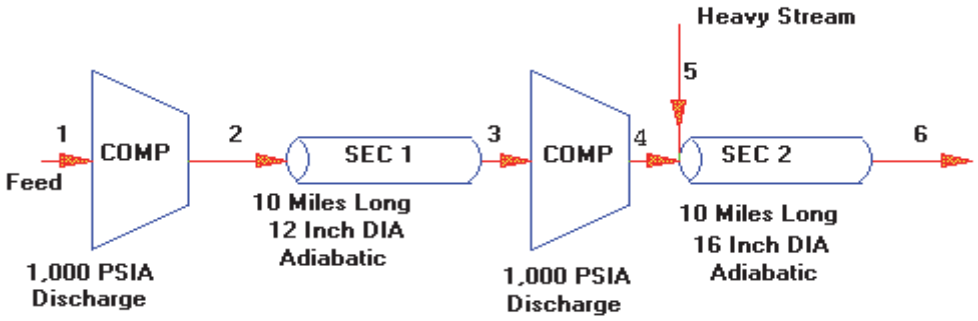
Gathrng.psd - Gas Gathering Network System

The objective of the simulation is to set-up a sequence of line modules to gather and transmit gas from various reservoirs with different pressures. A two stage compression and cooling is performed so that wet gas mixture can be further processed. Each line module has specified diameter, elevation change and heat transfer characteristics that are used for pressure drop calculations.



Line1.psd - Intermediate Compression and Junction

Gas is compressed to 1000 PSIA (COMPREssor 1) for transmission through a 12-inch diameter adiabatic line (LINE 10). At the next compressor station (COMPREssor 2) there is a junction with another stream of heavier composition (stream 5) entering the next line section (LINE 20) along with the compressor outlet stream (stream 4). LINE 20 is a 16-inch, two-phase adiabatic line. The plot of pressure and elevation profiles includes both LINE 10 and LINE 20.



20 Mile Pipeline with Intermediate Compression

```
COMPREssor 1 = COMP,1,-2, ENTropy, PREssure OUT = 1000,
LINE 10 = SEC 1,2,-3,
        DIAmeter (IN) = 12.0,
        ADIabatic,
        LENgth (MILES) = 10*1.0,
        ELEvation (FT) = 5,-2,3,4*1,3*2
COMPREssor 2 = COMP, 3,-4,
        PREssure OUT = 1000,
        ENTropy
LINE 20 = SEC 2,4,5,-6,
        DIAmeter (IN) = 16,
        ADIabatic
        LENgth (MILES) = 10*1.0,
        ELEvation (FT) = 10,5,2,5,-5,10,3,4,2,-8,
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,11,12,13,14,
        FLOw 1 = 5,3,5000,3000,1000,700,500,7*0
        TP 1 = 120,800
        SOAVEK
        TRANSport properties = NBS81
        FLOw 5 = 2*0, 50,100,3*50,7*200
        TP 5 = 120,1000
        PLOT PREssure AND ELEvation against LINEs = 10,20
END
```

A partial output file from Example 1 appears below.

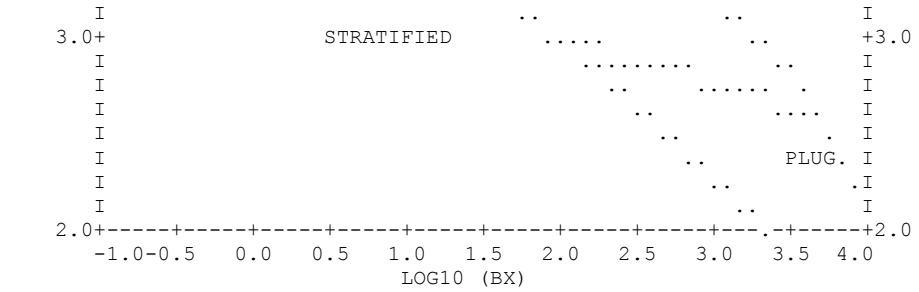
```
LIN        10 = SEC
             RESULTS OF LINE    10     -     GAS FLOW
              INLET STREAM(S)   2
             OUTLET STREAM(S)   3
             LINE CHARACTERISTICS: HORIZONTAL
                                   ADIABATIC
             ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       3.8732E+06  LIQUID FLOWRATE GAL/HR       0.0000E+00
PIPE ROUGHNESS    FT        1.5000E-04  PIPE EFFICIENCY FACTOR       1.000
INTERNAL DIAMETER INCHES   12.000       OUTSIDE DIAMETER INCHES     12.750
NOMINAL DIAMETER  INCHES   12.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              DARCY               MOODY               PHASE DENSITY

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.
```

LINE NODE	LENGTH FT	ELEV. FT	TEMP. DEG F	DELP-FRIC PSI	DELP-ELEV PSI	PRESS. PSIA	VELOC. FT/SEC	LIQUID HOLDUP	FLOW REGIME
0	0.0	0.0	150.5			1000.00	16.8	0.0000	TURB
1	5280.0	5.0	149.5	12.60	0.21	987.19	17.0	0.0000	TURB
2	10560.0	3.0	148.5	12.77	-0.08	974.51	17.2	0.0000	TURB
3	15840.0	6.0	147.4	12.94	0.12	961.44	17.5	0.0000	TURB
4	21120.0	7.0	146.4	13.12	0.04	948.28	17.7	0.0000	TURB
5	26400.0	8.0	145.3	13.31	0.04	934.93	18.0	0.0000	TURB
6	31680.0	9.0	144.1	13.51	0.04	921.38	18.3	0.0000	TURB
7	36960.0	10.0	143.0	13.72	0.04	907.63	18.6	0.0000	TURB
8	42240.0	12.0	141.8	13.94	0.08	893.62	18.9	0.0000	TURB
9	47520.0	14.0	140.5	14.16	0.07	879.38	19.2	0.0000	TURB
10	52800.0	16.0	139.3	14.41	0.07	864.90	19.5	0.0000	TURB

```
      ESTIMATED SONIC VELOCITY IN GAS PHASE
        =    940.3 FT/SEC   AT  INLET CONDITIONS.
        =    931.4 FT/SEC   AT OUTLET CONDITIONS.

LIN        20 = SEC
             RESULTS OF LINE    20     -     TWO PHASE FLOW
              INLET STREAM(S)   6
             OUTLET STREAM(S)   7
             LINE CHARACTERISTICS: HORIZONTAL
                                   ADIABATIC
             ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       2.8999E+06  LIQUID FLOWRATE GAL/HR       5.1535E+04
PIPE ROUGHNESS    FT        1.5000E-04  PIPE EFFICIENCY FACTOR       1.000
INTERNAL DIAMETER INCHES   15.250       OUTSIDE DIAMETER INCHES     16.000
NOMINAL DIAMETER  INCHES   16.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              LOCKHART-MARTINELLI MOODY               PHASE DENSITY
METHODS FOR:  LIQUID HOLDUP       FLOW REGIME
              HUGHMARK            BAKER

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.
```

LINE NODE	LENGTH FT	ELEV. FT	TEMP. DEG F	DELP-FRIC PSI	DELP-ELEV PSI	PRESS. PSIA	VELOC. FT/SEC	LIQUID HOLDUP	FLOW REGIME
0	0.0	0.0	166.1			1000.00	10.5	0.3092	DISP
1	5280.0	10.0	165.1	22.26	0.87	976.88	10.7	0.3026	DISP
2	10560.0	15.0	164.0	22.68	0.43	953.77	11.0	0.2960	DISP
3	15840.0	17.0	163.0	23.11	0.17	930.49	11.4	0.2896	DISP
4	21120.0	22.0	161.9	23.57	0.41	906.51	11.7	0.2830	DISP
5	26400.0	17.0	160.8	24.05	-0.41	882.87	12.1	0.2766	DISP
6	31680.0	27.0	159.5	24.54	0.80	857.53	12.5	0.2698	DISP
7	36960.0	30.0	158.3	25.09	0.23	832.20	12.9	0.2631	DISP
8	42240.0	34.0	157.0	25.67	0.31	806.22	13.4	0.2563	DISP
9	47520.0	36.0	155.6	26.28	0.15	779.79	13.9	0.2494	DISP
10	52800.0	28.0	154.2	26.93	-0.59	753.45	14.4	0.2426	DISP

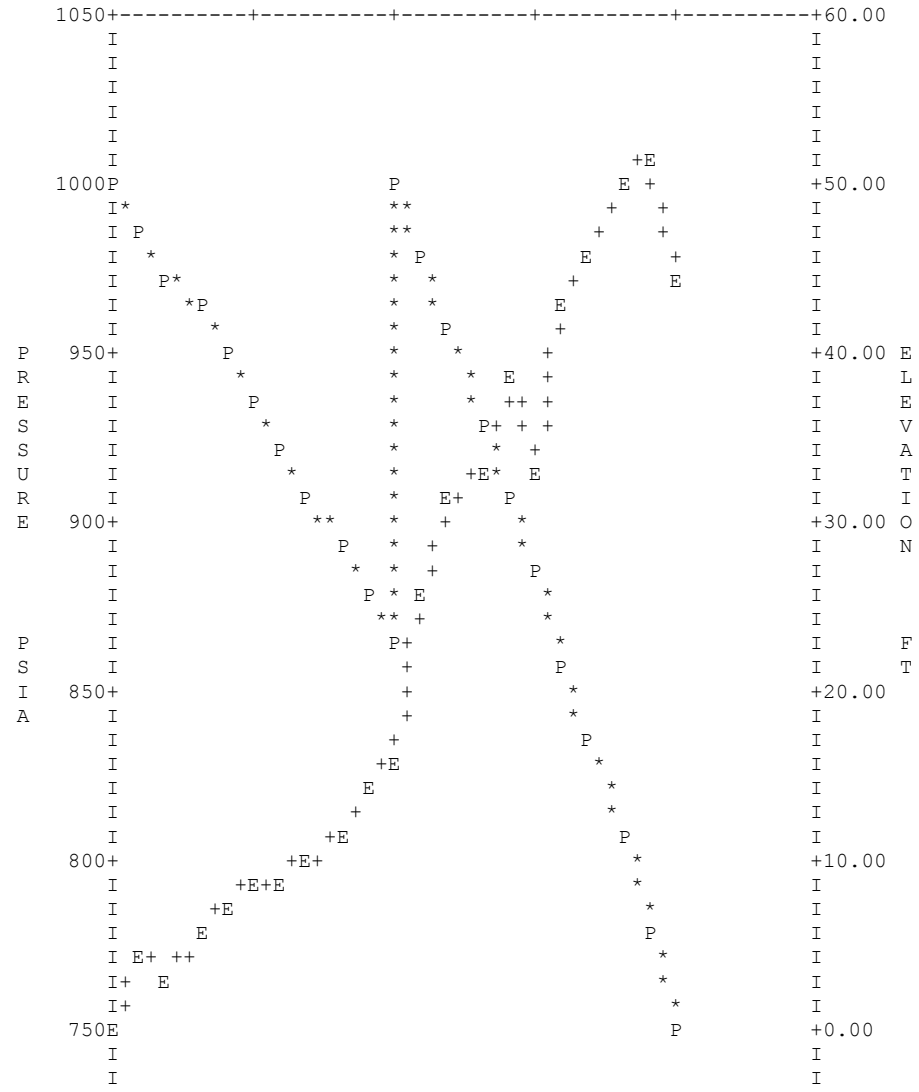
```
++NOTE: NO SLUG FLOW REGIME IS FOUND, SLUG ANALYSIS BYPASSED

      ESTIMATED SONIC VELOCITY IN GAS PHASE
        =   1022.0 FT/SEC   AT  INLET CONDITIONS.
        =   1030.2 FT/SEC   AT OUTLET CONDITIONS.
     *******  BAKER TWO-PHASE FLOW REGIME MAP FOR LINE    20 *******
       -1.0-0.5   0.0   0.5   1.0   1.5   2.0   2.5   3.0   3.5  4.0
       -1.0-0.5   0.0   0.5   1.0   1.5   2.0   2.5   3.0   3.5  4.0
    5.0+-----+----.+-----+-----+-----+-----+--.--+-----+-----+-----+5.0
       I..          ..                    .                      I
       I  ..          ...        DISPERSED  .                     I
       I    ..           .....              .                     I
       I      ..               .........XXX...                    I
       I        ..                          .                     I
L      I          ..                        .                     I
O      I    WAVE    ..          ANNULAR     . ..           BUBBLE  I
G   4.0+              ..                    .   .                  +4.0
1      I.............    ....              ..     ..               I
0      I             ...... .......    ...         .              I
       I                   ......  ....             .             I
       I                       ....            SLUG  .            I
B      I                           ...                .           I
Y      I                              ...              ..         I
```



BAKER FLOW REGIME COORDINATES

SEG NO.	X	Y	LOG10(X)	LOG10(Y)
1	67.214348	29006.177734	1.827462	4.462491
2	64.482086	29520.214844	1.809439	4.470119
3	61.824677	30055.173828	1.791162	4.477919
4	59.181705	30624.833984	1.772187	4.486074
5	56.664265	31206.027344	1.753309	4.494238
6	54.057697	31851.724609	1.732858	4.503133
7	51.542576	32522.601562	1.712166	4.512185
8	49.053478	33238.515625	1.690670	4.521642
9	46.609001	33998.683594	1.668470	4.531462
10	44.255524	34791.738281	1.645967	4.541476

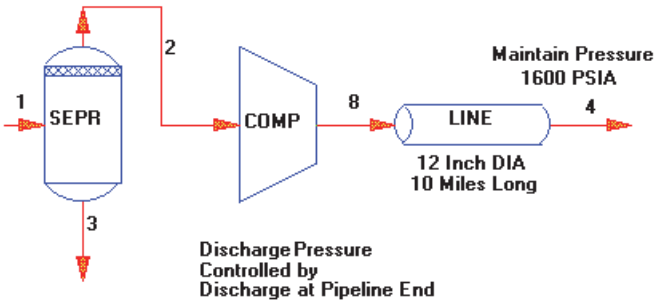
LINE PLOT 1 OF PRESSURE AND ELEVATION AGAINST LENGTH
FOR LINE(S) 10 20



```
        I                                                    I
        I                                                    I
        I                                                    I
        I                                                    I
    700+----------+----------+----------+----------+----------+-10.00
       0.00      5.00       10.00      15.00      20.00     25.00
                          LENGTH      MILES
```

Line2.psd - Back Pressure Calculation

An offshore transmission line has to meet an onshore pressure of 1600 PSIA. The coding below shows how this problem can be solved using a CONTroller module to back calculate the compression discharge pressure. Pressure drops due to friction and changes in elevation are calculated by the AGA method.



Pipeline with 1600 PSIA Discharge Pressure

```
FLAsh drum 1 = SEPR, -1,-2,-3,
COMPREssor 4 = COMP, 2,-8,
   ENTropy,PREssure OUT = 1700
LINE 7 = LINE, 8, 11,
   nominal DIAmeter (IN) = 12,
   DOWN,
   LENgth = 150,
   METhod for FRIction = AGA,
   DRAg factor = 0.96,
   METhod for ELEvation = AGA
LINE 8 = LINE, 11,12,
        DIAmeter (IN) = 12,
        LENgth (MILES) = 10*1,
        ELEvation (FT) = 3*10, 5,2,20,30,3*10,
        METhod for FRIction = AGA,
        DRAg factor = 0.96,
        METhod for ELEvation = AGA
CONTroller 10 = FIXP, 12,13,
        VARY PRE OUT (MIN=1500,MAX=2500) OF COMPRE 4
        UNTIL PREssure FROm STReam 12 = 1600 (TOL=25)
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,11,12,13,14,62
        IMMiscible = 62
        FLOw 1 = 5, 3, 5000, 3000, 1000, 700, 500, 233,
         339, 226, 427, 207, 266, 234, 100
        TP 1 = 120,1400
        SOAVEK, SOAVEH, STDD
END
```

Line3.psd - Diameter Case Studies

If you are studying alternative pipe sizes for a line with known flow and topography, you can calculate all of your cases in a single LINE module. In this case the downstream pressure out of the LINE module will be the outlet pressure from the last diameter calculated (20 inch).

14, 16, 18, 20 Inch
Line Diameter Study

```
LINE 2 = LINE,1,2,
        nominal DIAmeter (IN) = 14.0,16.0,18.0,20.0
        LENgth (MILES) = 10*1.0
        ELEvation (FT) = 5,2,3,4*1,3*2
GENeral
        COMponents = 2,3,4
        FLOw 1 = 500,1000,500
        TP 1 = 150,100
        TRANSport properties = NBS81
END
```

Line4.psd - Two Phase Line Calculations

An offshore transmission line is transporting 300 MMSCFD of gas through a 24 inch line over a distance of 35 miles. The example below uses a heat transfer coefficient and a temperature of the surrounding to determine the pressure drop and slugging characteristics of the line.

```
AB123.
*Two Phase Line Calculations

C- EQUIPMENT MODULES

LINE 1 = OFFSHORE, 1,-2,
  DIA (INCH) = 24
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 35. MILE
  LEN (MILE) = 25.,10.,
  ELE(FT) = 1.,5.,
  SEG LEN (MILE) = 3.5
  MET FRI = BEG
  TEM SUR(F) = 2*29.,
  U(BTU/HR/FT2/F) = 2*2.,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 46,49,2,3,4,5,6,7,8,10,11,12,13,
  PENK
  PENH
  STDD
  VIS = NBS81
  THE CON = NBS81
  BINPAR = PENG1
  MAX = 1
  CON TOL = 1.E-003
  FLO(FRA)1 = 3.E-003,1.1E-002,0.802,8.9E-002,6.E-002,7.E-003,1.7E-002,3.E-003,
  4.E-003,2.E-003,1.6E-003,3.E-004,1.E-004,
  TOT FLO(MMSCF/DAY)1 = 300.
  TP (F,PSIA) 1 = 50.,1350.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
END
```

Offshore Pipeline

The tabular results generated by the line module for the above input file is shown on the following page.

```
LIN         1 = OFFS
              RESULTS OF LINE     1      -      WET GAS FLOW
               INLET STREAM(S)    1
              OUTLET STREAM(S)    2
              LINE CHARACTERISTICS: HORIZONTAL
```

```
                                  HEAT TRANSFR
              ELEVATIONS ARE REPORTED RELATIVE TO INLET NODE.

VAPOR FLOWRATE SCF/HR       1.2256E+07  LIQUID FLOWRATE GAL/HR        6009.
PIPE ROUGHNESS    FT        1.5000E-04  PIPE DRAG FACTOR             1.000
INTERNAL DIAMETER INCHES   23.250       OUTSIDE DIAMETER INCHES     24.000
NOMINAL DIAMETER  INCHES   24.000       PIPE WALL THICKNESS CODE    STD

METHODS FOR:  FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
              BEGGS AND BRILL     BEGGS AND BRILL     PHASE DENSITY
METHODS FOR:  LIQUID HOLDUP       FLOW REGIME
              BEGGS AND BRILL     BEGGS AND BRILL

PRESSURE DROPS ARE REPORTED AS FROM THE PREVIOUS NODE TO THE CURRENT NODE.

LINE    LENGTH   ELEV.  TEMP. DELP-FRIC DELP-ELEV PRESS. VELOC. LIQUID  FLOW
NODE       FT      FT   DEG F   PSI       PSI     PSIA   FT/SEC HOLDUP REGIME

  0        0.0     0.0  50.0                     1350.00    8.3 0.1038  SEGR
  1    18480.0     0.1  44.3      8.41     0.01  1341.58    8.1 0.1262  SEGR
  2    36960.0     0.3  40.1      8.20     0.01  1333.37    8.0 0.1428  SEGR
  3    55440.0     0.4  37.0      8.06     0.01  1325.29    7.9 0.1552  SEGR
  4    73920.0     0.6  34.7      7.98     0.01  1317.31    7.9 0.1645  SEGR
  5    92400.0     0.7  33.0      7.93     0.01  1309.37    7.9 0.1714  SEGR
  6   110880.0     0.8  31.7      7.90     0.01  1301.45    7.9 0.1764  SEGR
  7   129360.0     1.0  30.8      7.90     0.01  1293.54    7.9 0.1801  SEGR
  8   132000.0     1.0  30.6      1.13     0.00  1292.41    7.9 0.1808  SEGR
  9   150480.0     2.8  29.9      7.93     0.13  1284.35    7.9 0.1833  SEGR
 10   168960.0     4.5  29.4      7.95     0.13  1276.27    8.0 0.1850  SEGR
 11   184800.0     6.0  29.1      6.84     0.11  1269.31    8.0 0.1857  SEGR
```

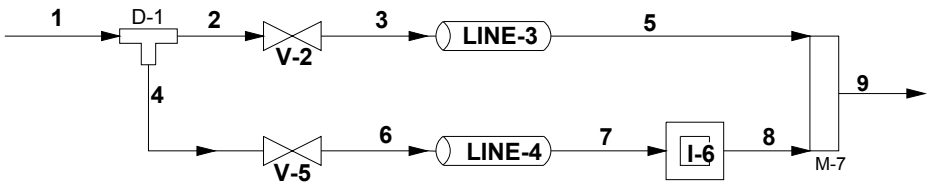
Line5.psd - Example Calculating Upstream Pressure

This flowsheet shows a line module calculating backwards by way of using a command for "calculating upstream pressure" (see Calculation Methodology below) when a target pressure out is needed for each line. After using this command one can set this calculated upstream pressure for the stream entering each of the line module. While doing this the divider flow fraction to each line can be varied to accommodate similar pressure fed to EACH line module. This is accomplished in the following way.

4. The first requires accessing CALculate UPStream PREssure command for a specified PRE OUT of each line module (see Inline Fortran and Keyword Input of each line module).
5. The valve pressure out is reset from the line module's calculated upstream pressure. Note that we have used user specified CALculation SEQuence command from the Specify...Keyword Input dialog.
6. The controller is set-up to vary the divider flow fraction until the stream pressures of #3 and #6 are close to each other. In this example the values were calculated as 365.48 and 365.28 psia.

Note that this example is specific for the line module specifications and feed pressure, temperature and flow specifications. If any change is done then you would have re-specify, change criteria in appropriate modules and re-run the case.

CALCULATE UPSTREAM PRESSURE (#3 & #6)



Calculation Methodology

The LINe module now allows a downstream pressure specification. When the down stream pressure is specified, the required upstream pressure is calculated. The calculated upstream pressure is available to other DESIGN II modules through Inline FORTRAN. For example, this feature may be used to specify an upstream compressor discharge pressure. The calculated upstream pressure is only reported when specified with Inline FORTRAN (see example below).

Major benefits of this improvement include direct calculation of downstream pressure. Previously, an iterative loop with a controller module was used to adjust the upstream pressure in order to achieve a specified downstream pressure. The calculated upstream pressure can be passed to an upstream module using Inline FORTRAN. You will need two additional commands in the GENeral section so that the pressure changing module and the Line module will be recalculated with the

new pressure value.

CALSEQ = n1, n2, ... or REC SEQ = n1, n2, ...

where n1, n2, ... are equipment module numbers, and MAX = 10

Example:

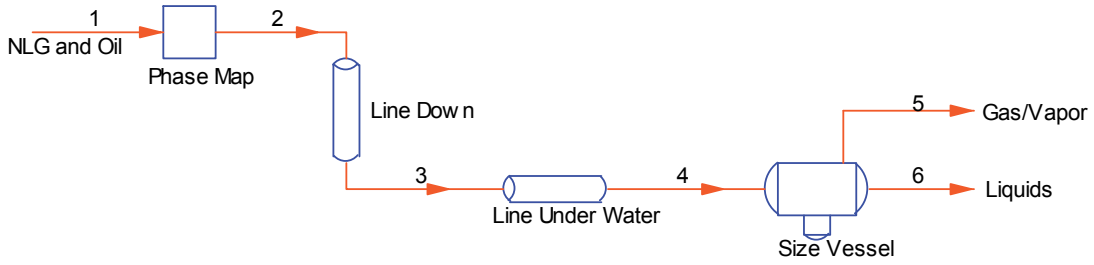
```
AB123.
*Calculate Upstream Pressure

C- EQUIPMENT MODULES

DIV 1 = D-1, 1,-4,-2,
  FLO(FRA) = 5.E-002

VAL 2 = V-2, 2,-3,
  PRE OUT(PSIA) = 10.

LINE 3 = LINE-3, 3,-5,
  DIA (INCH) = 18
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 10. MILE
  LEN (MILE) = 10.,
  SEG LEN (MILE) = 1.
  MET FRI = BEG
  ADI
PRE OUT (PSIA) = 360
F-      POST
F-      UPRE =GET(LINE(3),CAL UPS PRE)
F-      WRITE(2,95)
F-  95  FORMAT (' ')
F-      WRITE(6,100)UPRE
F-      WRITE(1,100)UPRE
F- 100  FORMAT(2X,'UPSTREAM PRESSURE LINE 3 (PSIA)=',G12.5)
F-      SET(VALVE(2),PRE OUT)=UPRE

LINE 4 = LINE-4, 6,-7,
  DIA (INCH) = 18
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 14. MILE
  LEN (MILE) = 14.,
  SEG LEN (MILE) = 1.4
  MET FRI = BEG
  ADI
PRE OUT (PSIA) = 320
F-      POST
F-      UPRE =GET(LINE(4),CAL UPS PRE)
F-      WRITE(2,95)
F-  95  FORMAT (' ')
F-      WRITE(6,100)UPRE
F-      WRITE(1,100)UPRE
F- 100  FORMAT(2X,'UPSTREAM PRESSURE LINE 4 (PSIA)=',G12.5)
F-      SET(VALVE(5),PRE OUT)=UPRE

VAL 5 = V-5, 4,-6,
  PRE OUT(PSIA) = 10.

CONT 6 = I-6, 7,-8,
VARY FLO (MIN=0.05, MAX=1.0, STEP=0.05) OF DIV 1
UNTIL PRE FROM STR 3/PRE FROM STR 6=1.0 (TOL=0.005,LOO=100)

MIX 7 = M-7, 5,8,-9,

C- GENERAL SECTION
GENERAL
  COMPONENTS = 49,2,3,4,5,6,7,8,10,11,
  PENK
  PENH
  STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
  THE CON VAP = NBS81
  THE CON LIQ = NBS81
  BINPAR = PENG1
  MAX = 5
```

```
  CON TOL = 1.E-003
  FLO(FRA)1 = 1.1E-002,0.802,8.9E-002,6.E-002,7.E-003,1.7E-002,3.E-003,4.E-003,
  2.E-003,1.6E-003,
  TOT FLO(MMSCF/DAY)1 = 100.
  TP (F,PSIA) 1 = 50.,900.
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
  NAM STR 8 = Strm 8
  NAM STR 9 = Strm 9
CAL SEQ=1,2,3,2,5,4,5,6,7,
END
```

For Information on retrieving and printing calculated values such as upstream pressure, see Inline Fortran: Table 2. Variables Available to GET and SET.

Line6.psd - Line Surroundings with Heat Transfer

The example is to illustrate the heat transfer calculations equipped in the line module. Line down is exposed to Air at 70 F and 4 Ft/sec and the line under water is exposed to water flowing at .05 m/min. The external resistance due to heat transfer (conduction & convection) is included for the line modules.

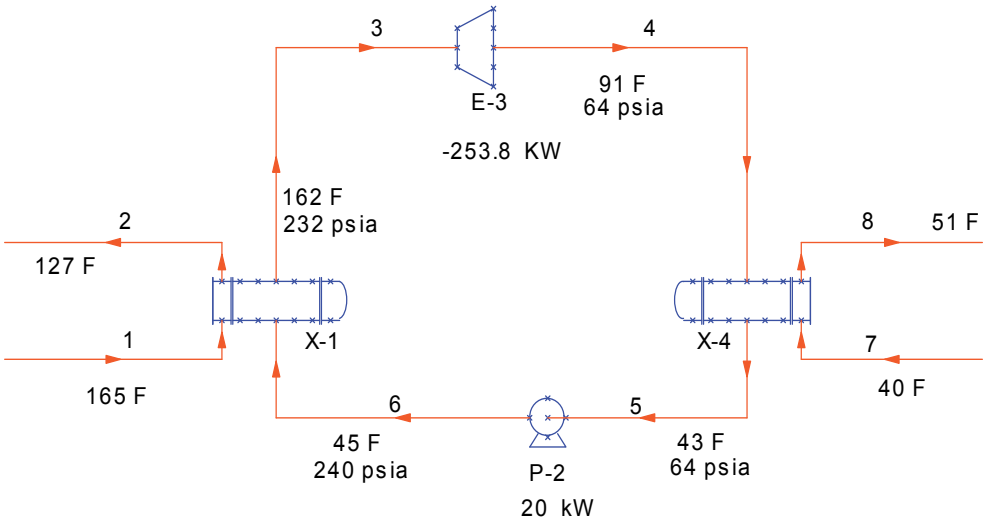



```
LINE 2 = Line Down, 2,-3,
  DIA (INCH) = 16
  PIP WAL = STD
  DOWN
C- ENTRIES IN LEN COMMAND TOTAL TO 45.72 M
  LEN (M) = 45.72,
  SEG LEN (M) = 4.572
  MET HOL = DUN
  TEM SUR(F) = 70,
  PIP MAT = STA
  SUR AIR
  VEL SUR(FT/SEC) = 4,

LINE 3 = Line Under Water, 3,-4,
  DIA (INCH) = 16
  PIP WAL = STD
  HORI
C- ENTRIES IN LEN COMMAND TOTAL TO 16000 M
  LEN (M) = 10*1600,
C- NOTE: the absolute elevation specification was changed to elevation change
  ELE (M) = 3.048,3.048,3.048,1.524,0.609,6.097,9.144,3.049,3.047,3.047
  SEG LEN (M) = 1600
  MET HOL = MUK
  TEM SUR(F) = 10*60,
  PIP MAT = CAR
  SUR WAT CON(KCAL/HR/M/C) = 10*0.1709,
  VEL SUR(M/MIN) = 10*0.051,
```

Chapter 67: Power - c:\designii\samples\power

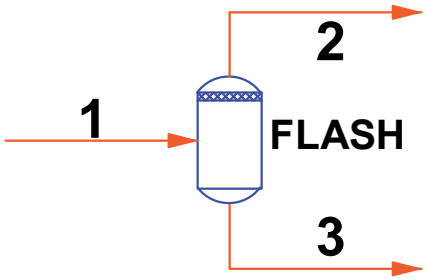
geothermal_power_system.psd – Geothermal Power in Alaska

Geothermal Power in Alaska Holds Hidden Model for Clean Energy (Popular Mechanics, February 2008) -
At Chena Hot Springs Resort, a visionary owner and an ingenious engineer tap into one of the world's most overlooked energy resources—not fossil fuels—to produce electricity, heat buildings and soon, they hope, generate hydrogen.



Chapter 68: Process Line Sizing Samples - c:\designii\samples\sizeline

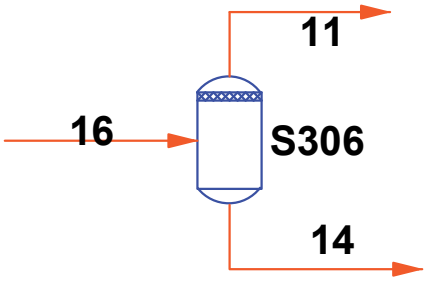
Size1.psd - Liquid DEA Solution Line Sized With Default Values



Sizing of a DEA solution stream with default values (of pressure drop per 100 ft, velocity, sonic velocity fraction, or nominal diameter).

```
AB123.
*SIZE LINES - EXAMPLE 1
FLAsh 1 = FLASH,1,-2,-3
SIZe LINes
        USE stream = 3
GENeral
        COMponents = 4051,62
        MDEAK,MDEAH
        FLOw (LB/HR)1 = 18967.5, 56902.5
        TP 1 = 120, 25
END
```

Size2.psd - Flare Gas Header Sized Using Sonic Velocity Specification

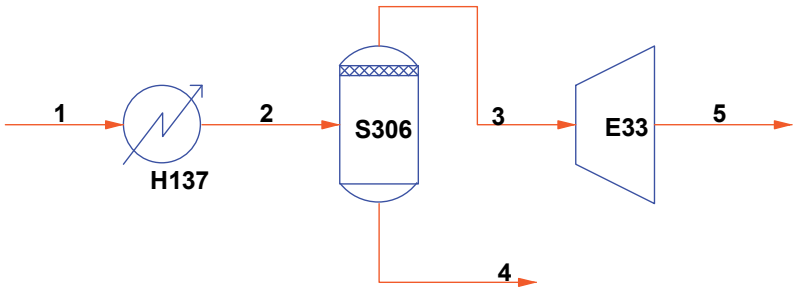


Sizing of a flare gas header stream using sonic velocity fraction.

```
AB123.
*SIZE LINES - EXAMPLE 2
FLAsh drum 19 = S306,16,-11,-44
GENeral
        COMponents = 2,3,4
        FLOw 16 = 10000,100,100
        TP 16 = 100,1200
        APISOAVEK
SIZe LINes
        USE stream = 11,
        DIRection of flow = UP,
        LENgth = 50,
        MAXimum SONic fraction = 0.6
END
```

Size3.psd - Two-Phase Vertical Reboiler Return Line Sized Using Pressure Drop Specification

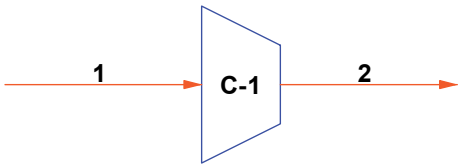
Flowsheet example of sizing reboiler return lines having two phase flow by coding the reboiler as a heat exchanger and sizing its outlet stream.




```
AB123.
*SIZE LINES EXAMPLE
HEAt EXChanger H7 = H137,21,-16,
        DELta pressure = 10,
        TEMperature OUT = -45
FLAsh drum 19 = S306,16,-11,-44
EXPander 37 = E33,11,-31,-33,
        PREssure OUT = 300
GENeral
        COMponents = 2,3,4
        FLOw 21 = 2300,400,600
        TP 21 = 100,800
        APISOAVEK,APISOAVEH
SIZe LINes
        USE streams = 2,3,5,
        DIRection of flow = HORIzontal
        MAXimum DELta P = 6.0, 1.0, 0.9
END
```

Size4.psd - Line with Two-Phase Upwards Flow using Specified Line Size and Roughness

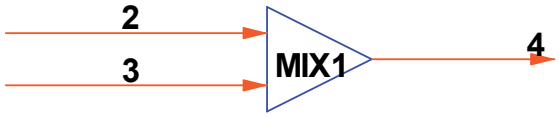
Example to size a two phase stream flowing upwards using a specified line and roughness factor.




```
AB123.
*SIZE LINES - EXAMPLE 5
COMPREssor 1 = PRES,1,-2, PREssure OUT = 1000
GENeral
        COMponents = 49,46,2,3,4,5,6,7,8,100
        AMW100 = 190, NAM100 = C6+,
        API(SPG)100 = 0.8, AMB100 = 490
        APISOAVEK,APISOAVEH,STDD
        PETroleum UNIts OUT
        FLOw 1 = 80,40,1500,440,190,56,60,37,26,310
        TP 1 = 100.0,800
SIZe LINes
        USE stream = 2
        NOMinal DIAmeter (IN) = 6,
        INSide DIAmeter (IN) = 6.065, DIRection of flow = UP,
        ROUghness (FT) = .0002
END
```

Size5.psd - Three-Phase Line With Inclined Vertical Upwards Flow

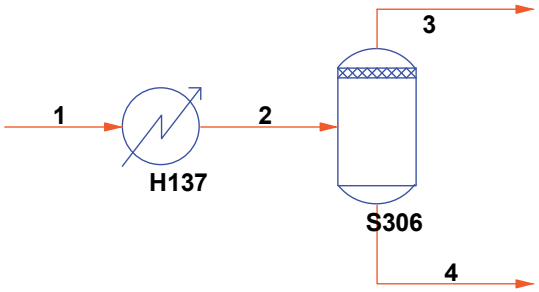
Example to size a three phase stream flowing inclined upwards using a specified pressure drop requirement.




```
AB123.
*SIZE LINES - EXAMPLE 6
MIX 1 = FLASH,2,3,-4
GENeral
        COMponents = 46,49,2,3,4,5,6,7,8,10,100,62
        PENK,PENH,PEND
        API100 = 38.5,AMW100 = 200,AMB100 = 500,NAMe100 = C7+
        FLOw (MMSCF/DAY) 2 = 10*200,10,170, TP 2 = 100,1000
        FLOw (BBL/DAY) 3 = 0,95000,9*0,170
        TP 3 = 100,1000
        IMMiscible  = 62
        TRANSport properties = NBS81
SIZe LINes
        USE stream = 4, DIRection of flow = UP,
        LENgth = 100, ELEvation = 90, PIPe WALl CODE = XS, MAXimum DELta P per hundred feet = 25,
        METhod for FRIction = ORKiszewski
END
```

Size6.psd - Different Sizing Criteria On Different Lines

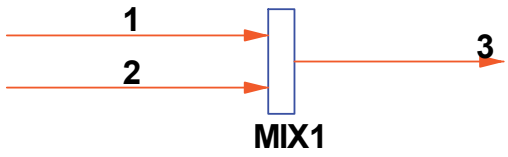
Example to size different lines with different sizing criteria.




```
AB123.
*SIZE LINES - EXAMPLE 7
HEAt EXChanger 7 = H137,21,-16, DELta pressure = 10,
        TEMperature OUT = -45
FLAsh drum 19 = S306,16,-11,-44
GENeral
        COMponents = 2,3,4
        FLOw  21 = 2300,400,600, TP 21 = 100,800
        APISOAVEK,APISOAVEH
        TRANSport properties = NBS81
SIZe LINes
        USE  streams = 21,16,
        MAXimum VELocity (FT/SEC) = 2*10
SIZe LINes
        USE stream = 11,
        DIRection of flow = UP, NOMinal DIAmeter (IN) = 4, PIPe WALl code = STD
SIZe LINes
        USE stream = 44,
        MAXimum DELta P per hundred feet = 0.5
END
```

Size7.psd - Natural Gas Line Sized Using Two Different Friction Factor Methods

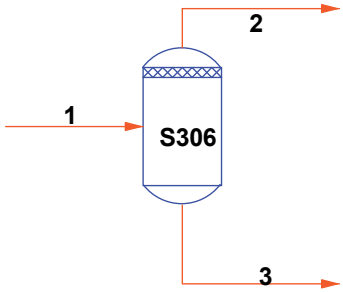
Sizing of a natural gas line with two different friction factor methods for comparison.




```
AB123.
*SIZE LINES - EXAMPLE 8
MIXer 10 = MIXR,1,2,-3
GENeral
        APISOAVEK,APISOAVEH,APISOAVED
        COMponents = 49,46,2,3,4,5,6,7,8,10,11,12,13,14,15
        TP (F,PSIA) 1 = 40,900
        TP (F,PSIA) 2 = 40,900
        FLOw (FRA) 1 = .00056, .00029, .10818, .07039, .10875, .09692, .1142, .10886, .07205, .13468,
        .0892, .06328, .02411, .00727,.00127
        TOTal FLOw (GAL/HR) 1 = 500
        FLOw (FRA) 2 = .001772, .0093702, .8492089, .0778564, .0366486, .0115022, .0080801, .0030361,
        .00127, .0009179, .000274,  .0000561, .0000064, .0000011, .0000001
        TOTal FLOw (SCF/HR) 2 = 750000
SIZe LINes
        USE streams = 3,3,
        MAXimum DELta P per hundred feet (PSI) = 0.35,0.35,
        METhod FRIction = WEYmouth,MODified Panhandle, EFFiciency = 0.96,0.96,
        LENgth (FT) = 10000,10000, ELEvation (FT) = 800,800
END
```

Size8.psd - CO2 Line With Bends Sized Using Line Of Specified Size and AGA Method Specification

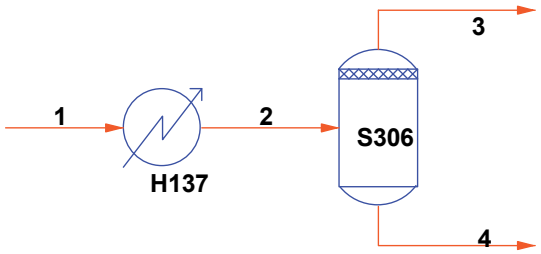
Flowsheet example to size a CO2 line with bend (equivalent length) with known line size and American Gas Association Method.




```
AB123.
*SIZE LINES - EXAMPLE 9
FLAsh 10 = FLA1,1,-2,-3
GENeral
        COMponents = 49
        FLOw (SCF/HR) 1 = 1000000
        TP1 = 150,1000
        BWRSD,BWRSH,BWRSK
SIZe LINes
        USE streams = 1,
        NOMinal DIAmeter (IN) = 12.0,
        LENgth (FT) = 10000.0,
        ELEvation (FT) = 10.0,
        EQUivalent length (FT) = 25.0,
        METhod of FRIction = AGA,
        DRAg = 0.97
END
```

Size9.psd - Two Phase Inclined Horizontal Line Sized With Maximum Velocity Specification

Example to size a two phase inclined horizontal line using a maximum allowable velocity specification.




```
AB123.
*SIZE LINES - EXAMPLE 10
C-  EQUIVALENT LENGTH IS USED FOR BENDS AND FITTINGS
HEAt EXChanger H7 = H137,21,-16,
        DELta pressure = 10,
TEMperature OUT = -45
FLAsh drum 19 = S306,16,-11,-44
GENeral
        COMponents = 2,3,4
        FLOw 21 = 2300,400,600
        TP 21 = 100,800
        APISOAVEK,APISOAVEH
SIZe LINes
        USE stream = 2,
        DIRection of flow = HORIzontal, LENgth (FT) = 1000,
        ELEvation (IN) = 18,
        EQUivalent length (FT) = 15,
        MAXimum VELocity (FT/SEC) = 10
END
```

Results of the SIZing LINes calculation for example 10 are shown on the following pages. The column labeled "TARGET" shows results for the minimum diameter which meets the maximum velocity specification. The results for the next larger and next smaller diameters are also shown. The second page shows the flow regime for each of the 3 sizes on a Baker chart.

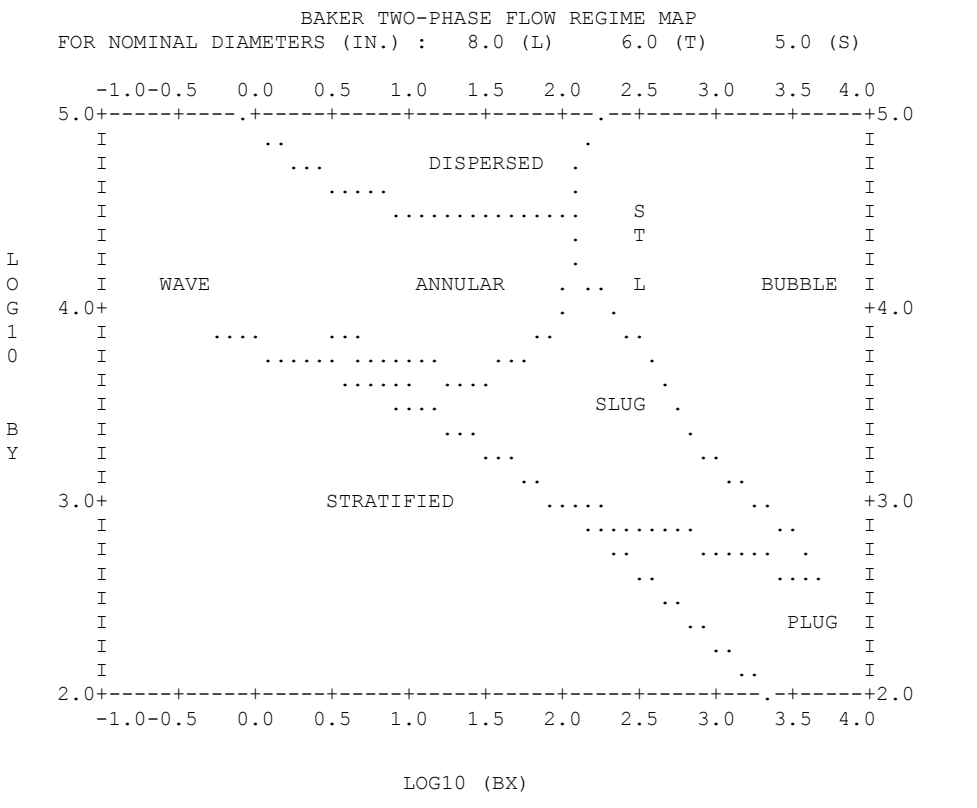
```
************** LINE SIZING RESULTS FOR 2 PHASE STREAM   16 **************
LINE IS ISOTHERMAL AND HORIZONTAL
METHODS USED : FRICTION DELTA P    FRICTION FACTOR     ELEVATION DELTA P
               LOCKHART-MARTINELLI MOODY               PHASE DENSITY

VAPOR  FLOW RATE           LB/HR      0.2294E+05
LIQUID FLOW RATE           LB/HR      0.5244E+05
LIQUID VOLUME FRACTION                0.3025
TEMPERATURE                DEG F      -45.00
INLET PRESSURE             PSIA        790.0
INTERNAL DIAMETER          IN.         6.065          DEFAULT
MAX VELOCITY               FT/SEC     10.000          SPECIFIED
LINE LENGTH                FT          1000.          SPECIFIED
EQUIVALENT LENGTH          FT          15.00          SPECIFIED
ELEVATION                  FT          1.500          SPECIFIED
PIPE WALL CODE                          STD           DEFAULT
PIPE ROUGHNESS             FT         0.1500E-03      DEFAULT
DENSITY          (VAPOR)   LB/FT3      4.707
                 (LIQUID)  LB/FT3      24.81
MOLECULAR WEIGHT (VAPOR)               17.71
                 (LIQUID)              26.16
VISCOSITY        (VAPOR)   CP         0.1132E-01
                 (LIQUID)  CP         0.6579E-01
Z-FACTOR(AT INLET PRESSURE)           0.6679
SURFACE TENSION                       1.855      DYNES/CM
SONIC VELOCITY IN GAS                 1032.      FT/SEC
*  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *  *
                                LARGER        TARGET         SMALLER
I.D. FOR BAKER MAP                L             T              S
NOMINAL DIAMETER    IN.         8.000         6.000          5.000
INTERNAL DIAMETER   IN.         7.981         6.065          5.047

DELTA P (FRICTION)  PSI         3.302         12.02          28.71
DELTA P (ELEVATION) PSI        0.1407        0.1383         0.1366
```

DELTA P (TOTAL)	PSI	3.443	12.16	28.85
PRESSURE DROP/100FT	PSI	0.3443	1.216	2.885
MEETS VELOCITY SPEC		YES	YES	NO
OUTLET PRESSURE	PSIA	786.6	777.8	761.2
LIQUID VELOCITY	FT/SEC	1.690	2.927	4.227
VAPOR VELOCITY	FT/SEC	3.897	6.748	9.744
2-PHASE VELOCITY	FT/SEC	4.464	7.864	11.49
NO SLIP VELOCITY	FT/SEC	5.587	9.675	13.97
REYNOLDS NUMBER	VAPOR	0.1603E+07	0.2109E+07	0.2535E+07
REYNOLDS NUMBER	LIQUID	0.6302E+06	0.8293E+06	0.9966E+06
FROUDE NUMBER		1.459	5.756	14.42
FLOW REGIME - BAKER		BUBBLE	BUBBLE	BUBBLE
FRICTION FACTOR	VAPOR	0.1465E-01	0.1527E-01	0.1577E-01
FRICTION FACTOR	LIQUID	0.1538E-01	0.1577E-01	0.1615E-01
2-PHASE DENSITY	LB/FT3	13.50	13.27	13.12
LIQUID HOLDUP		0.4376	0.4262	0.4184

THESE RESULTS ARE SUBJECT TO THE LIMITATIONS AND ACCURACY OF THE
METHODS USED. LITERATURE REFERENCES ARE GIVEN IN THE USER MANUAL.

THESE RESULTS ARE SUBJECT TO THE LIMITATIONS AND ACCURACY
OF THE METHODS USED. LITERATURE REFERENCES ARE GIVEN IN THE USER MANUAL.

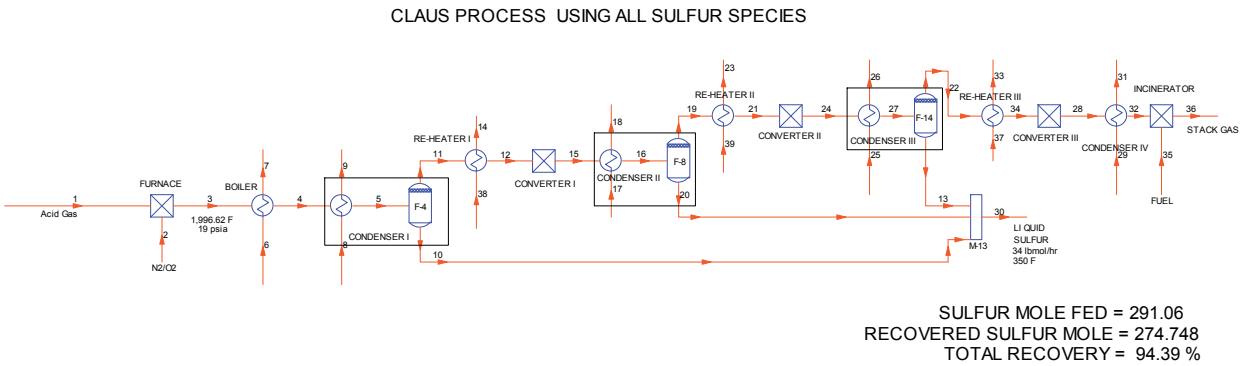
********** LINE SIZING RESULTS FOR TWO PHASE STEAM 16 **********



Chapter 69: Sulfur Samples-c:\designii\samples\sulfur

ClausAllSulfurspecies.psd – GPSA Data Book Claus Process

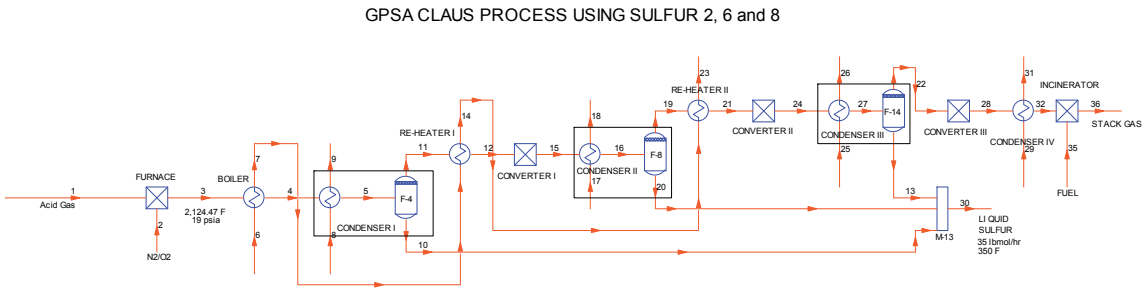
This sample flowsheet models the Claus Process in the GPSA data book using Sulfur 2, Sulfur 3, Sulfur 4, Sulfur 5, Sulfur 6, Sulfur 7, and Sulfur 8.



SULFUR MOLE FED = 291.06
RECOVERED SULFUR MOLE = 274.748
TOTAL RECOVERY = 94.39 %

ClausGPSA.psd – GPSA Data Book Claus Process

This sample flowsheet models the Claus Process in the GPSA data book using Sulfur 2, Sulfur 6, and Sulfur 8.



SULFUR MOLE FED = 291.06
RECOVERED SULFUR MOLE = 280.48
TOTAL RECOVERY = 96.36 %

Sulfurreactor1.psd – Converting Hydrogen Sulfide to Elemental Sulfur

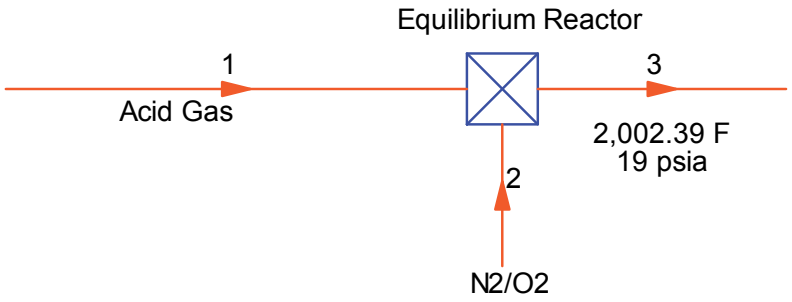
This flowsheet illustrates a sulfur process that utilizes the following chemical reactions to convert hydrogen sulfide to elemental sulfur (using Sulfur 2, Sulfur 3, Sulfur 4, Sulfur 5, Sulfur 6, Sulfur 7, and Sulfur 8) :

(1) $H_2S + {}^3/_2\, O_2 \rightleftharpoons SO_2 + H_2O$

(2) $2\, H_2S + SO_2 \rightleftharpoons {}^3/_n\, S_n + 2\, H_2O$

The overall reaction for the process is:

(3) $3\, H_2S + {}^3/_2\, O_2 \rightleftharpoons {}^3/_n\, S_n + 3\, H_2C$



SulfurreactorGPSA.psd – Converting Hydrogen Sulfide to Elemental Sulfur

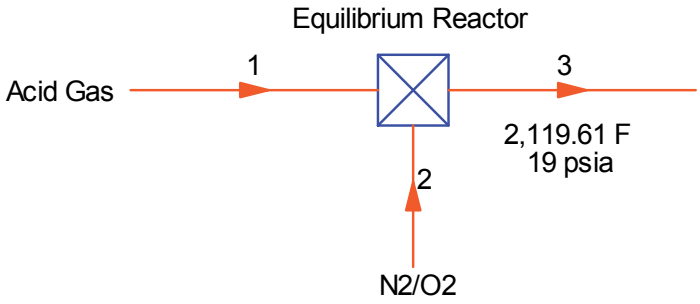
This flowsheet illustrates a sulfur process that utilizes the following chemical reactions to convert hydrogen sulfide to elemental sulfur (using Sulfur 2, Sulfur 6, and Sulfur 8) :

(1) $H_2S + {}^3/_2\, O_2 \rightleftharpoons SO_2 + H_2O$

(2) $2\, H_2S + SO_2 \rightleftharpoons {}^3/_n\, S_n + 2\, H_2O$

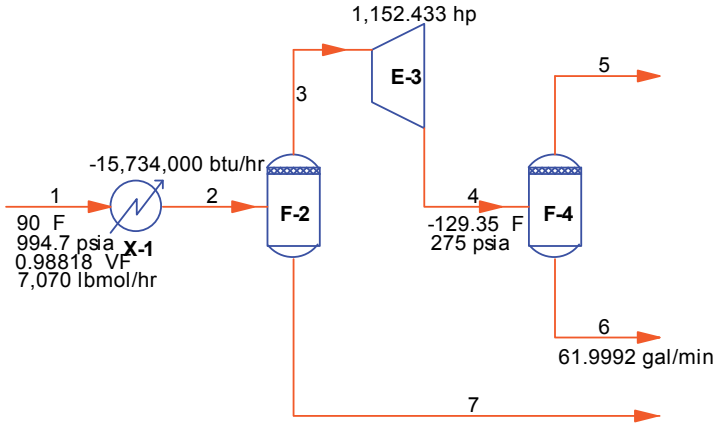
The overall reaction for the process is:

(3) $3\, H_2S + {}^3/_2\, O_2 \rightleftharpoons {}^3/_n\, S_n + 3\, H_2O$



Chapter 70: Thermodynamics Samples-c:\designii\samples\Thermodynamics

Expander_Apisoave.psd – Cooling the Feed Stream



The purpose of this simulation is to cool the feed stream (90 F at 980 psig) to -35 F, separate the resulting stream into two phases, and then expand the vapor phase to 275 psia. The feed contains 6100 lbmol/hr of methane, 500 lbmol/hr of ethane, 200 lbmol/hr of propane, 200 lbmol/hr of butane, 200 lbmol/hr of pentane and 70 lbmol/hr of hexane.

```
AB123.
*EXPANDER PLANT PROBLEM
C- EQUIPMENT MODULES

HEA EXC 1 = X-1, 1,-2,
  U(BTU/HR/FT2/F) = 50
  TEM OUT(F) = -35
  SHE PAS = 1
  TUB PAS = 1
  SHE PAR = 1
  SHE SER = 1
  TUB FEE = 1
  INL = OPP
  HOR
  DEL(PSI) = 10

FLA 2 = F-2, 2,-3,-7,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

EXP 3 = E-3, 3,-4,
  PRE OUT(PSIA) = 275
  EFF = 0.8

FLA 4 = F-4, 4,-5,-6,
  ADI, HEA(BTU/HR) = 0
  DEL(PSI) = 0

C- GENERAL SECTION
GENERAL
  COMPONENTS = 2,3,4,6,8,10,
  NAM 2 = 'METHANE'
  NAM 3 = 'ETHANE'
  NAM 4 = 'PROPANE'
  NAM 6 = 'N-BUTANE'
  NAM 8 = 'N-PENTANE'
  NAM 10 = 'N-HEXANE'
  APISOAVEK
  APISOAVEH
  STDD
  DEN LIQ = STDD
  VIS VAP = NBS81
  VIS LIQ = NBS81
```

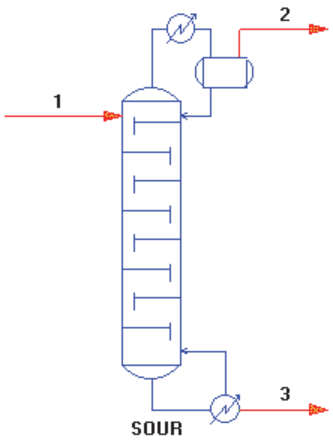
```
  THE CON VAP = NBS81
  THE CON LIQ = NBS81
  SUR TEN = STD
  AME UNI OUT
  MAX = 1
  CON TOL = 0.001
  BULK = 1,
  FLO(LBMOL/HR)1 = 6100,500,200,2*100,70,
  TP (F,PSIG) 1 = 90,980
  NAM STR 1 = Strm 1
  NAM STR 2 = Strm 2
  NAM STR 3 = Strm 3
  NAM STR 4 = Strm 4
  NAM STR 5 = Strm 5
  NAM STR 6 = Strm 6
  NAM STR 7 = Strm 7
END
```

Apisour.psd (Distl8.psd) – Sour Water Stripper with CO2

This example has to consider the solubilities and vapor phase interactions of H2S, NH3 and CO2 in water. The effect of CO2 on the pH of the system is also taken into account.

Preliminary steps:

5. Use COM CON = 30,3 with regular convergence technique or COM CON = 12,4 and DVS = 0.4 with SUPer to insure stability of K-values.
6. Use APISOUR K-values and YEN-Alexander enthalpies. The APISOUR technique also handles the solubilities of H2, N2, methane, ethane, propane, isobutane, and normal butane in water. K-values for heavier hydrocarbon are calculated using the Chao-Seader correlation.
7. If you do not enter a TEMperature PROfile, use the top tray temperature rather than the condenser temperature as an initial guess.
8. The APISOUR technique can also model use of caustics or organic acids to control pH. See Thermodynamics Section, for details.



Feed	Stream 1
Temperature, F	230
Pressure, PSIA	34.7
Flowrate, LB/HR	
NH3	155
H2S	234
CO2	40
H2O	49,426
CH4	800
C2H6	100

Guesses: Temperature of Top = 128 F
Temperature of Bottom = 239 F
Vapor Profile = 75, 825, 8 * 0, 708

Specifications: Vapor Product rate = 75.0 LBMOL/HR, Reflux = 10.0

NOTE: If you are entering LIQuid COMposition profiles for CO2 or NH3, you do not need COM CON = 12,4 or SUPer.

```
AB123.
* SOUR WATER COLUMN
C-
DIStillation 1 = SOUR, 1, -2, -3
        PARtial CONdenser
        TRAys = 9,
        LOCation of FEEd = 3,
        PREssure of TOP = 180,
        DELta pressure = 3
        TEMperature of TOP  = 128,
        TEMperature of BOTtom  = 239
C-
C- SPECIFICATIONS
        PROduct rates = 75, 0,
        REFlux = 10.0
C-
C- OPTIONAL INPUT
        PREssure of CONdenser = 26.7
```

```
C-
C- INITIAL GUESSES
        VAPor PROfile = 825, 7 * 0, 722
        REBoiler VAPor GUEss = 708
C-
C- CONVERGENCE CONTROL
        SUPer, COMpositional CONvergence = 12, 4, DVS = 0.4
GENeral
        COMponents = 63, 50, 49, 62, 2, 3
        FLOw rates (LB/HR) 1 = 155, 234, 40, 49426, 800, 100
        TP 1 = 230, 34.7
C-
C- THERMODYNAMICS OPTIONS
C-  K-VALUES     ENTHALPY
            APISOUR,       YEN
END
```

Beychok_Sour.psd (Distl6.psd) – Sour Water Stripper

This example illustrates the simulation of Sour water strippers using the Beychok method. NH3, H2S, and H2O interaction date is provided automatically. Because of the high degree of compositional dependence of the K-values for H2S, NH3, H2O systems the following steps are recommended.

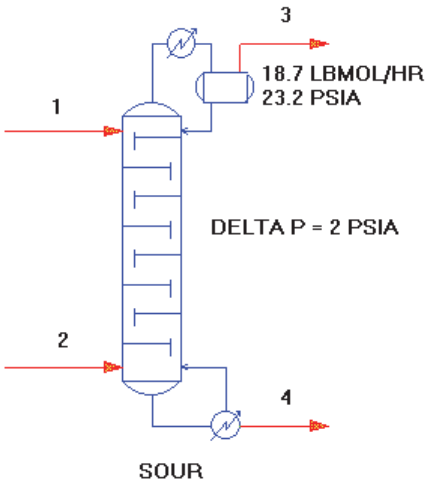
6. Use the Steam Tables* and an educated guess on NH3 and H2S effects to make an accurate guess for the TEM TOP and TEM BOT. If you do not enter a TEMperature PROfile, enter the value for the top tray as your guess for TEMperature of TOP rather than the product temperature. The column profile if fairly linear, except between trays 0 and 1. A change in the temperature guess of 5 F may be the difference between convergence and nonconvergence.
7. Use the Steam Tables* with the assumption that feeds are all water or steam, to estimate a VAPor PROfile.
8. Use COM CON = 10,2 (COM CON 12,4 and DVS = 0.4 with SUPER) to insure stability in K-values.
9. Tighter overall column balances can be obtained by changing the TOLerance to 1E-6 (Default is 1E-5) with negligible effect in the costs.
10. Use SOUR for K-values and YEN-Alexander heats for enthalpies. SOUR uses the Beychok technique for H2O, H2S and NH3 K-values. For other components Chao-Seader K-values are used unless overridden by entering tabular K-values via ChemTran.

* A preliminary run with VAPor PREssure K-values and LATent heat enthalpies can be used to generate this information.

Feeds	Stream 1	Stream 2
Temperature, F	180	309
Pressure, PSIA	23.2	24.7
Flowrate, LB/HR		
H2O	74500	11800
NH3	149.6	0
H2S	89.6	0

Guesses:

Reflux ratio = 32.7
Temp top product = 199 F
Temp bottoms product = 240 F
Vapor profile = 411, 695, 5*0, 725 Specification:
Top product rate = 18.7 LBMOL/HR

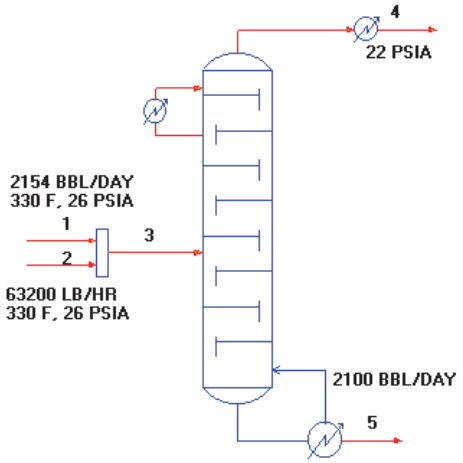



```
AB123.
* SOUR WATER STRIPPER
C-
DIStillation 1 = SOUR , 1, 2, -3, -4
        STRipper
        TRAys = 8,
        LOCation of FEEds = 1, 9
        PREssure of TOP = 23.2,
        DELta pressure = 2
        TEMperature of TOP  = 199,
        TEMperature of BOTtom  = 240
C-
C- SPECIFY TOP PRODUCT RATE
        PROduct rates = 18.7, 0
        REFlux GUEss = 32.7
        VAPor PROfile = 411, 695, 5 * 0, 725
        COMpositional CONvergence = 10, 2,
        TOLerance = 1E-6
        PRInt SWItch = 2
GENeral
        COMponents = 62, 63, 50
        TP 1 = 180, 23.2,
        TP 2 = 309, 24.7
```

```
        FLOwrates (LB/HR) 1 = 74500, 149.6, 89.6
        FLOwrates (LB/HR) 2 = 11800, 0, 0
C-
C- FORCE STREAM 2 TO BE ALL VAPOR
        VAPor = 2
C-
C- THERMODYNAMIC CHOICES
C-   K-VALUES    ENTHALPIES
        SOUR , YEN
C-
C- SET OUPUT UNITS OF MASS FLOWS TO POUNDS
        QUAntity UNIts OUT = LB
END
```

Braun K-10.psd (Refi4.psd - Pumparound Internal Reflux with Total Condenser (PUM TOT)

Example of refine with internally refluxed column having a top tray pumparound to generate reflux. There is no liquid returned from the external condenser to the main column. Water, if present, is decanted automatically from the external condenser.



Feed Stream 1 Component Flowrates :

	Water	0
	Toluene	63200 lb/hr

Temperature = 330 F Pressure = 26 psia

Feed Stream 2 Flowrate = 63200 LB/HR

Temperature = 330 F Pressure = 26 psia

Crude Characterization :

Volume percent	TBP(F)
0	400
5	475
10	609
20	785
30	885
40	956
50	1006
60	1044
70	1083
80	1122
90	1161
95	1181
100	1200

Bulk API Gravity = 8.1

Guesses: Liquid Distillate Flowrate = 691 lbmoles/hr
Bottom Product Flowrate = 69 lbmoles/hr
Temperature of Condenser = 259 F
Temperature Profile = 265,277,297,316,400,7*0,766 F
Vapor Profile = 689,850,824,787,87,120,202,240,4*0,140 lbmoles/hr

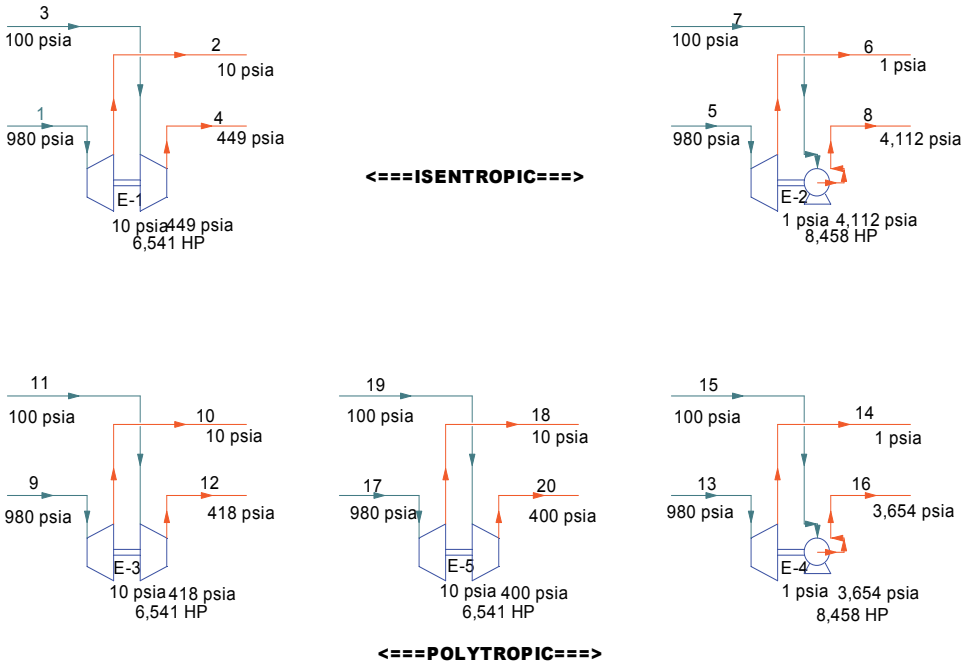
Specification: Top Pumparound Duty = 2.7 MMBtu/hr

```
AB123.
* BITUMEN RECOVERY ; REFINE PUM TOT EXAMPLE
MIXer 1 = MIXR, 1, 2, -3
REFI 2 = COAL, 3, -4, -5
        PUMped TOTal condenser
        STAges = 13
        PREssure of TOP = 25 ,
        DELta pressure = 1.5
        PREssure of CONdenser = 22
        LOCation of FEEd = 5
        PROduct rates GUEss = 691, 69
        TEMperature of CONdenser GUEss = 259
        LOCation of PUMparound = 2, 1
        PROduct of PUMparound = 400
        HEAt from PUMparounds = 2.7E6
        REBoiler (BTU/HR) = 9.4E6
        TEMperature PROfile =265,277,297,316,400,7*0,766
        VAPor PROfile =689,850,824,787,87,120,202,240,4*0,140
        REPort = 1, 1, 1, 0, 0, 1
        PROduct NAMes = SOLVENT, BITUMEN
```

```
        COLumn sizing OFF
C-
C- CRUDE FEED DESCRIPTION
C-
CRUde
        FEEd rate (BBL/DAY) 1 = 2154
        FEEd TBP 1 = 400, 475, 609, 785, 885, 956, 1006, 1044,
                1083, 1122, 1161, 1181, 1200
        FEEd GRAvity 1 = 9.1
        FEEd NAMe 1 = BITUMEN FEED
        FEEd NAMe 2 = TOLUENE FEED
        FEEd rate (LB/HR) 2 = 0, 63200
        TP 1 = 330 , 26 , TP 2 = 330, 26
C-
C- GENERAL SECTION
C-
GENeral
        COMponents = 62, 41 , STDH , BRAUN
END
```

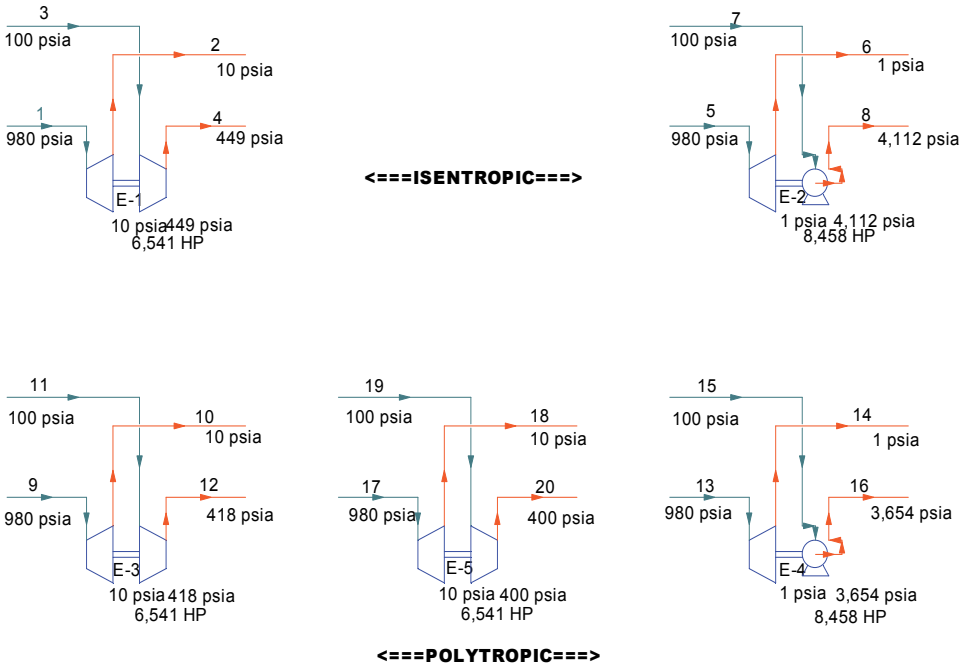
BWR.psd – Using BWR Thermodynamic Method

Sample illustrating the use of the Benedict-Webb-Rubin (BWR) equation of state. It is an eleven parameter equation developed for light hydrocarbon mixtures containing methane through pentane, nitrogen and carbon dioxide in the range of 26 F to 400 F.

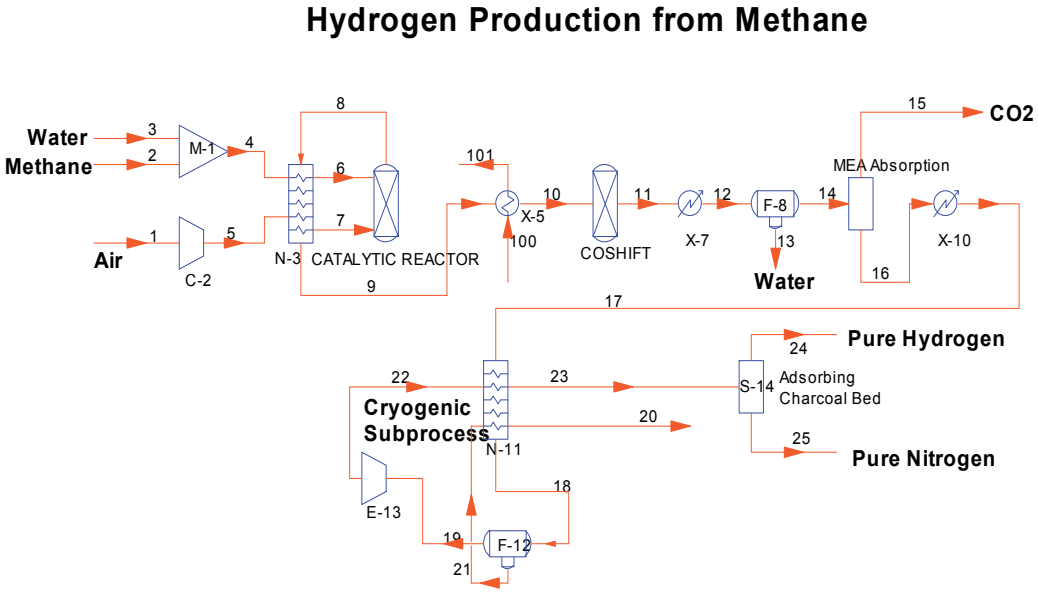


BWRS.psd - Using BWRS Thermodynamic Method

Sample illustrating the use of the 11 parameter Benedict-Webb-Rubin equation of state by Starling, et. al. It has an extended list of interaction coefficients and generalized correlation for parameters by Starling.



Chao_Seader.psd – Hydrogen Production from Methane

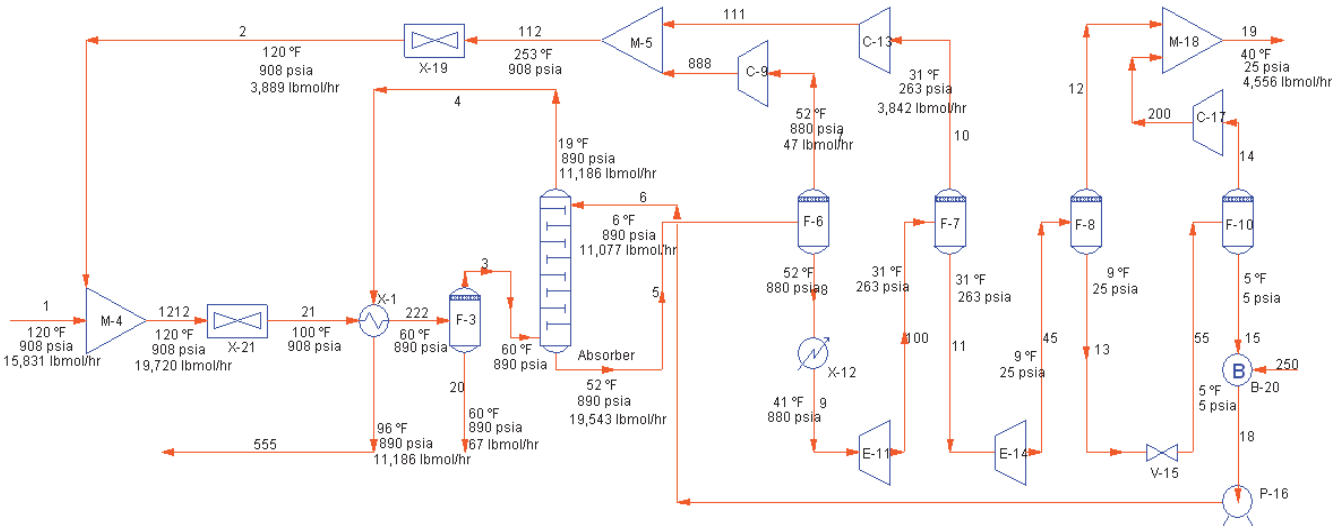


The above process is an illustration of hydrogen manufacture via a catalytic reactor. Methane, steam and air are each pre-heated to 600-700 C, and fed into a catalytic reactor containing a bed of refractory nickel catalyst (secondary reformer is used for simulation purposes). Initial combustion results in a temperature up to 1000 C. The product gas passes through heat exchange and a heat recovery boiler before entering the multi-stage CO shift converter. The gas then passes through a CO2 wash tower (can use MEA if needed in a separate flowsheet simulation...please see General: Gas Sweetening with MEA Example). Following CO2 absorption, the gas passes through a cryogenic separator where the hydrogen is separated from the CO, nitrogen, and methane.

The simulation described here produces 50 MMSCF/DAY of 99.93% pure hydrogen from methane and steam.

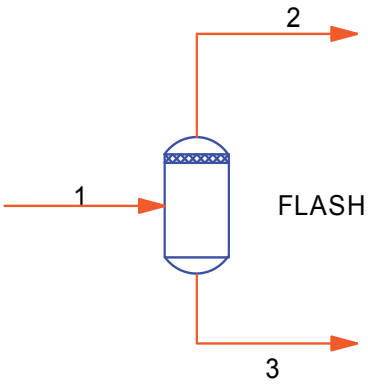
dmpeg1.psd - Using DMPEG Thermodynamic Method

This sample flowsheet models the removal of acid gases such as CO2, H2S, COS, and mercaptans from natural and synthesis gas systems with DMPEG, a physical solvent that is a mixture of polyglycols. DMPEG is also known by the trade names such as Coastal AGR, Selexol and Genosorb 1753.



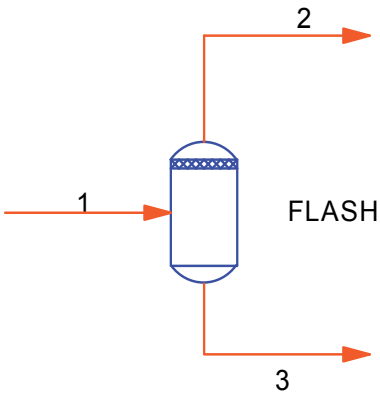
Edwards_K_with Chemtran.psd – Edwards K-Value Sample

This sample simulation uses K-values for weak, aqueous electrolytic solutions based on the Edwards et al., model. The solvent must be water. Total solute concentrations should be <= 10 molal (gram moles of solute per kilogram of solvent) and total ionic strength should be less than or equal to 6 molal.



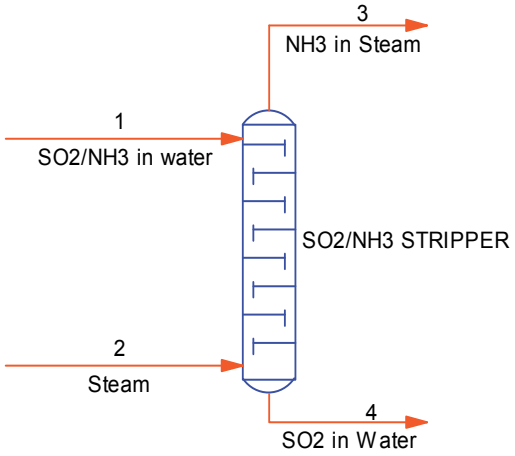
Edwards_K_without Chemtran.psd – Edwards K-Value Sample

This sample simulation uses K-values for weak, aqueous electrolytic solutions based on the Edwards et al., model. Henry's constants in water for the following example are built into the Pure Component Database and does not use ChemTran.



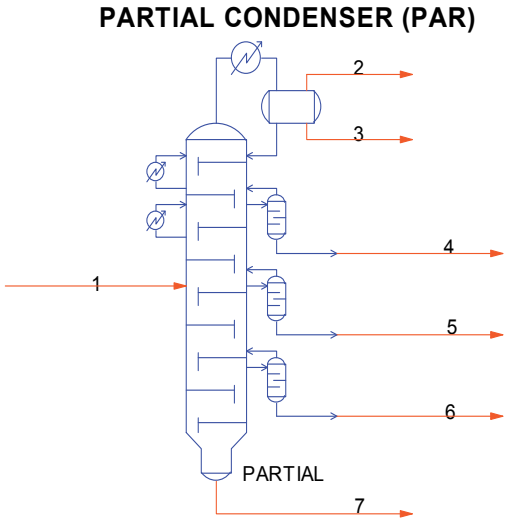
Edwards_Sour.psd – Aqueous Edwards Model

This sample simulation uses K-values for weak, aqueous electrolytic solutions based on the Edwards et al., model. This method is applicable for all ions, reactions, and components which are in the Design II ionic library. Henry's constants in water for the following example are built into the Pure Component Database and does not use ChemTran.



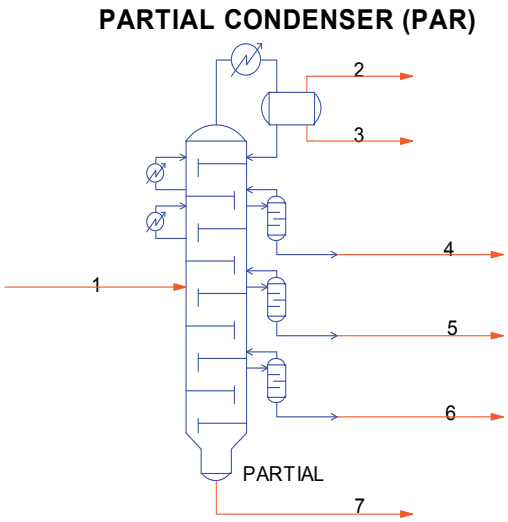
Esso_Maxwell_Bonnell.psd – Partial Condenser

This simulation uses Curve-fitted equations (developed at Exxon Research and Engineering Co.) from the Maxwell-Bonnell vapor pressure charts which are used for heavy hydrocarbon systems (crude oils).



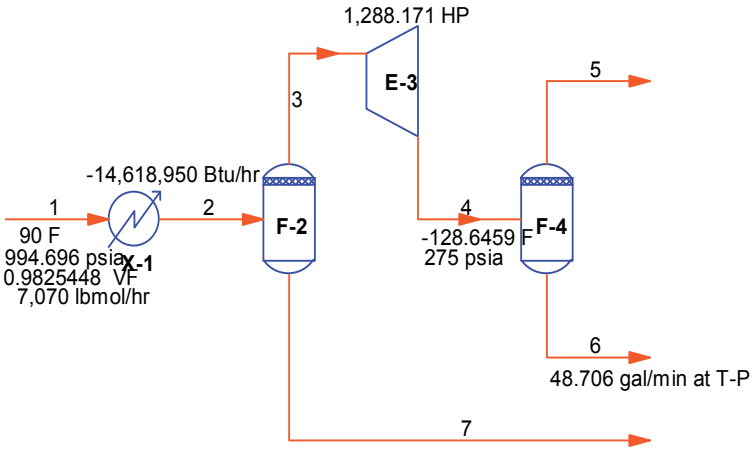
Esso_Tab.psd – Esso Tabular

This simulation uses a faster, tabular form of the ESSO correlation (vapor pressure charts from the API Technical Data Book). Primarily used for heavy hydrocarbon systems at pressures below 50 PSIA.



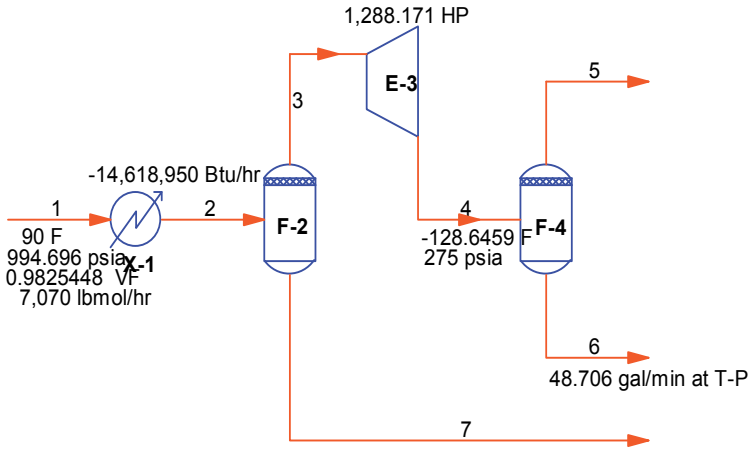
gerg2004.psd – Expander Plant

This sample uses GERG (Groupe Européen de Recherches Gazières) 2004 correlation which is a wide-range reference equation of state for natural gases and other mixtures (limited to 18 components) based upon a multi-fluid mixture model approximation explicit on the reduced Helmholtz free energy.



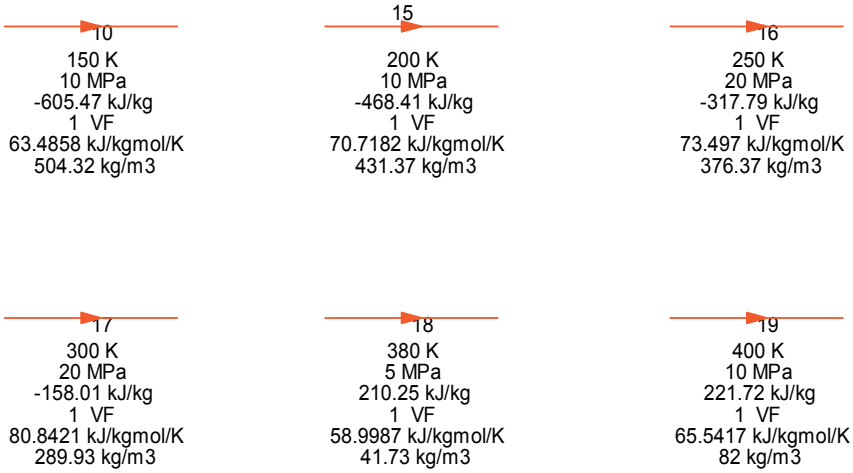
gerg2008.psd – Expander Plant

This sample uses GERG (Groupe Européen de Recherches Gazières) 2008 correlation which is a wide-range reference equation of state for natural gases and other mixtures (limited to 21 components) based upon a multi-fluid mixture model approximation explicit on the reduced Helmholtz free energy.



Gerg_iso_20765-1_samples.psd – GERG Thermodynamic Method

This sample uses GERG (Groupe Européen de Recherches Gazières) 2008 correlation which is a wide-range reference equation of state for natural gases and other mixtures (limited to 21 components) based upon a multi-fluid mixture model approximation explicit on the reduced Helmholtz free energy.



Gerg_iso_20765-2_samples.psd – GERG Thermodynamic Method

This sample uses GERG (Groupe Européen de Recherches Gazières) 2008 correlation which is a wide-range reference equation of state for natural gases and other mixtures (limited to 21 components) based upon a multi-fluid mixture model approximation explicit on the reduced Helmholtz free energy.

26
180 K
10 MPa
-56,419.93 kJ/kg
0.242 VF
746.71 kg/m3
139.08 kg/m3
162.7927 kJ/kgmol/K
160.5623 kJ/kgmol/K
744.55 kg/m3

32
250 K
10 MPa
-165.24 kJ/kg
0.95 VF
174.7513 kJ/kgmol/K
688.16 kg/m3
838.12 kg/m3

33
300 K
25 MPa
-42.15 kJ/kg
0.951 VF
187.599 kJ/kgmol/K
661.27 kg/m3
855.86 kg/m3

34
400 K
5 MPa
204.86 kJ/kg
0.999 VF
225.1252 kJ/kgmol/K
549.16 kg/m3
939.27 kg/m3

35
450 K
0.5 MPa
600.27 kJ/kg
1 VF
195.5182 kJ/kgmol/K
11.81 kg/m3

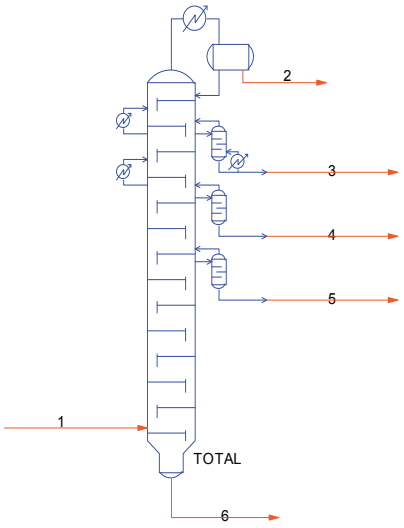
36
500 K
2 MPa
686.94 kJ/kg
1 VF
235.4879 kJ/kgmol/K
52.174 kg/m3

37
180 K
10 MPa
-12,266.21 kJ/kg
0 VF
1,164 kg/m3

Grayson_Streed.psd – Total Condenser

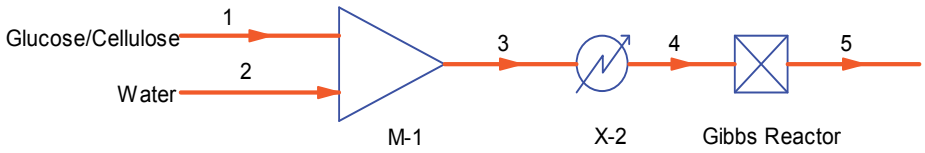
This sample uses a combination of regular solution theory for activity coefficients and empirical functions by Grayson-Streed to describe the liquid phase with the Redlich-Kwong equation for the vapor phase. The correlation was developed for hydrocarbons, hydrogen and gas oils and should give reasonable values for temperatures from 60 F to 800 F (or below critical temperature of mixture) for pressures up to 3000 PSIA.

TOTAL CONDENSER (TOT)



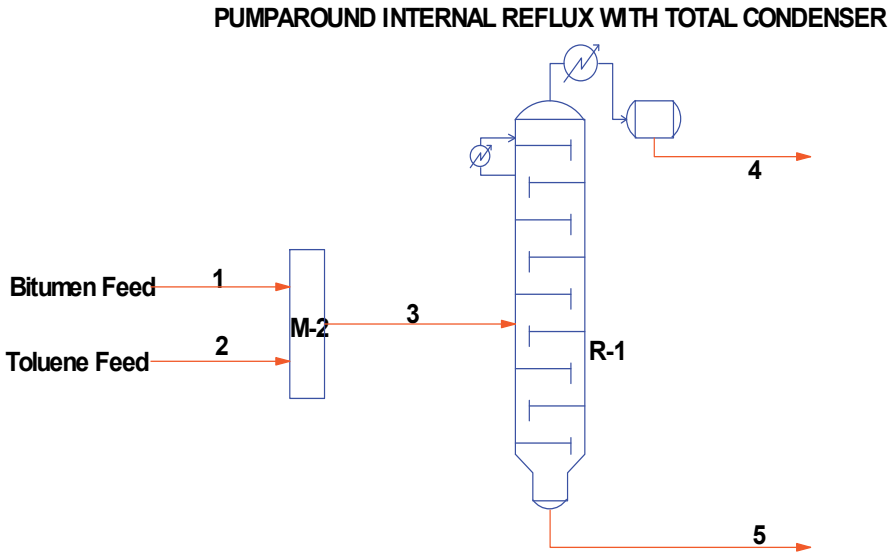
Chemtran_Glucose_Ideal.psd – Ideal Method

This is an illustration of ideal method using ChemTran defined Glucose and Cellulose species for their conversion to Ethanol via Gibbs Free energy minimization reactor.



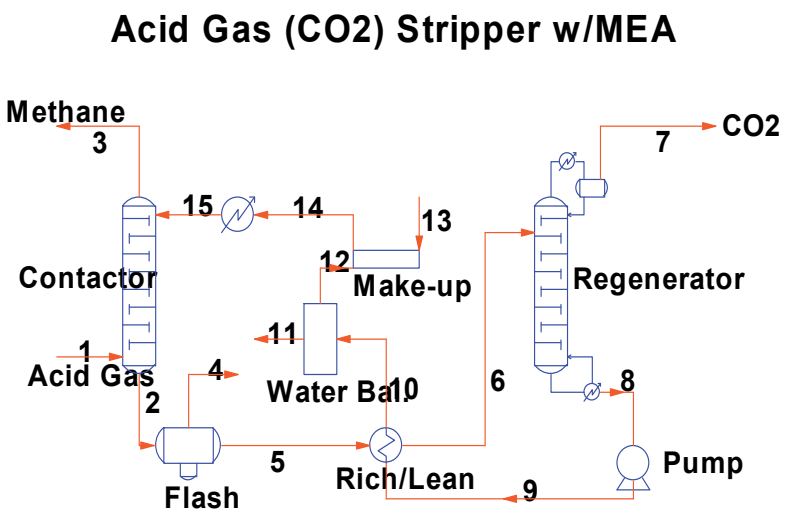
LKPK.psd – Pumparound using Lee-Kessler-Ploecker

This is an application of the BWR equation by Lee-Kesler-Ploecker. They use the BWR equation to represent a simple fluid (based on argon, krypton, and methane) and a reference fluid (octane). Mixture calculations use mixture rules, reduced temperature, reduced pressure, and acentric factors to combine simple and reference fluids together.



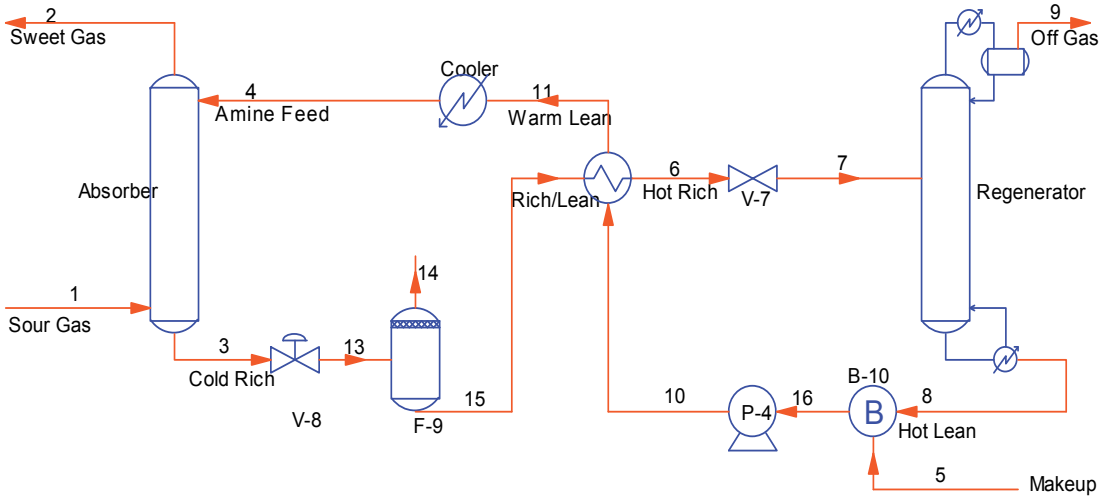
MEA_DEA_Ideal.psd – Gas Sweetening with MEA

This sample uses rigorous model for removal of CO2 and/or H2S from natural gas streams using either monoethanolamine (component ID no. 4155) or diethanolamine (component ID no. 4051) solutions.



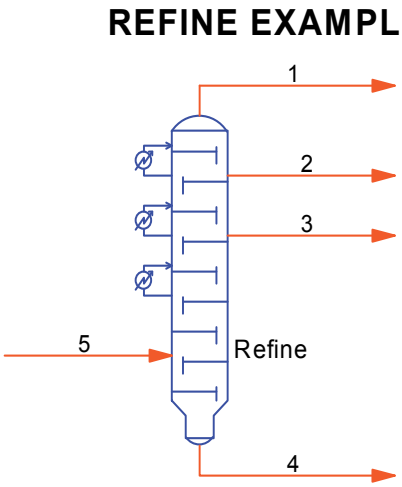
Mixed Amine.psd – Mixed Amine Plant for CO2/H2S Removal

This sample uses rigorous model for the removal of CO2 and/or H2S from natural gas streams using MEA, DEA, MDEA, DGA, and DIPA amine solutions. Please see Thermodynamics Table 6. Mixed Amine Modeling for other components for which this technique is applicable. K-values for all components are calculated using Kent-Eisenberg and Deshmukh-Mather models.



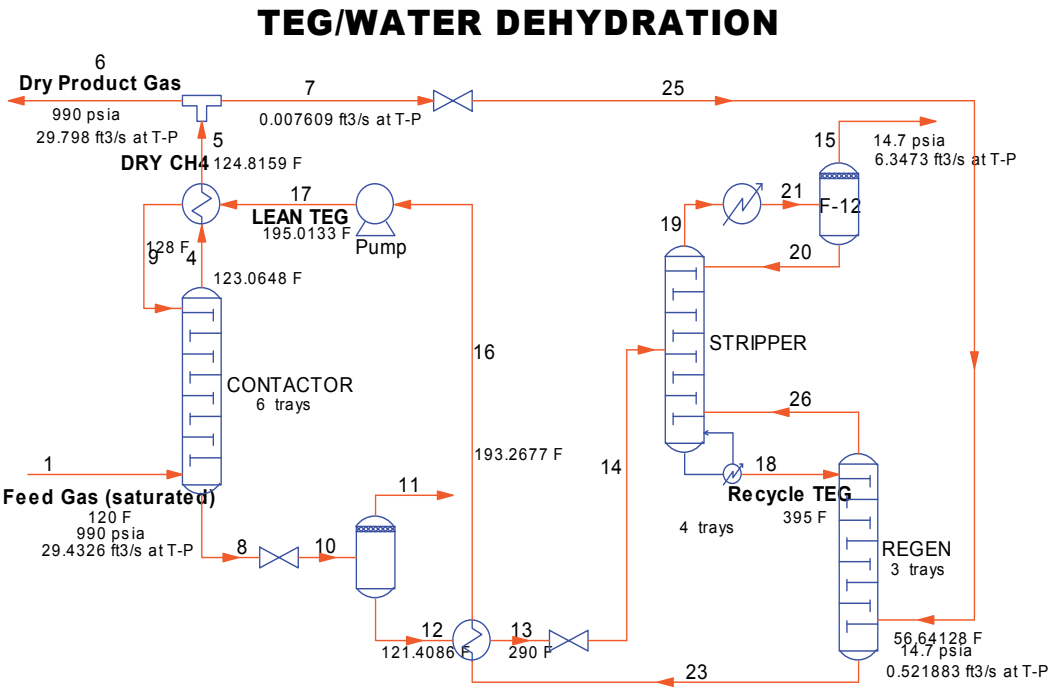
Modified Esso.psd – Pumparound Internal Reflux without Condenser

This is a simulation using the Lee-Kesler modification to the ESSO vapor pressure correlation. This option should improve prediction of K-values for heavy and aromatic hydrocarbons (Watson characterization factor which is not equal to 12).



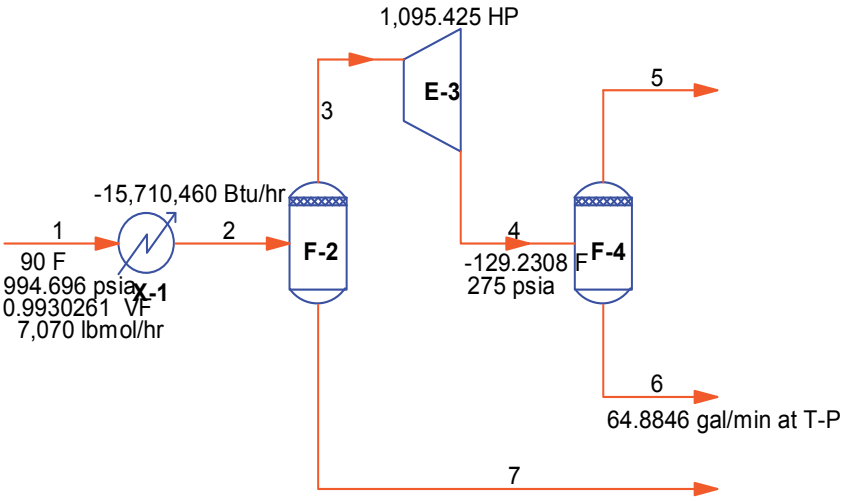
Teg1_MODPENK.psd – Modified Peng Robinson

This sample uses Design II modifications of Peng-Robinson equation of state based on work by Stryjek et al. This equation is applicable for non-ideal chemical systems at high pressure. This method predicts the liquid phase with the accuracy of activity coefficient methods with the in-built modified Peng Robinson parameter library.



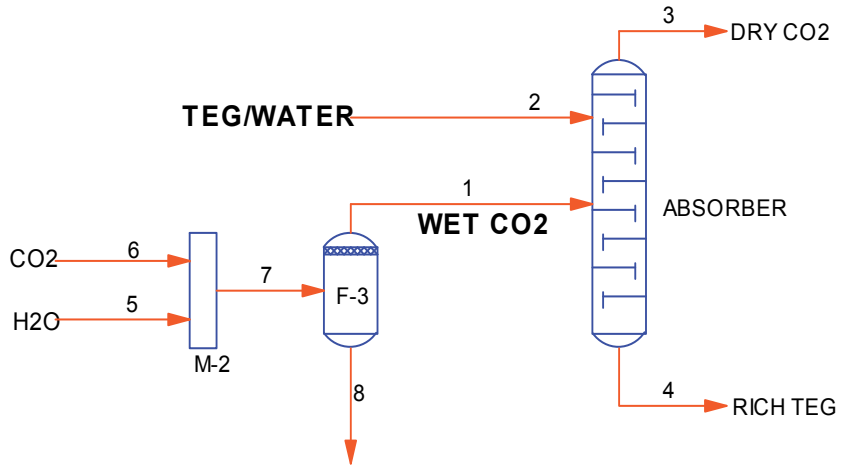
EXPANDER_Peng.psd – Expander Plant using Peng Robinson

This sample uses Peng-Robinson equation (with interaction parameters) that improves the applicability of the cubic equation of state for vapor-liquid equilibrium calculations in the vicinity of the critical region. With interaction parameters, this option is suitable for simulating cryogenic ethane and propane recovery, nitrogen rejection and systems where H2S and CO2 predominate.



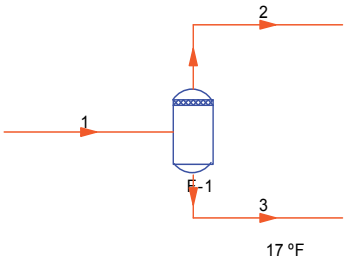
TEGCO2_Peng.psd – Water Dehydration using Peng Robinson

This sample uses Peng-Robinson equation (with interaction parameters) that improves the applicability of the cubic equation of state for vapor-liquid equilibrium calculations in the vicinity of the critical region. With interaction parameters, this option is suitable for simulating cryogenic ethane and propane recovery, nitrogen rejection and systems where H2S and CO2 predominate.



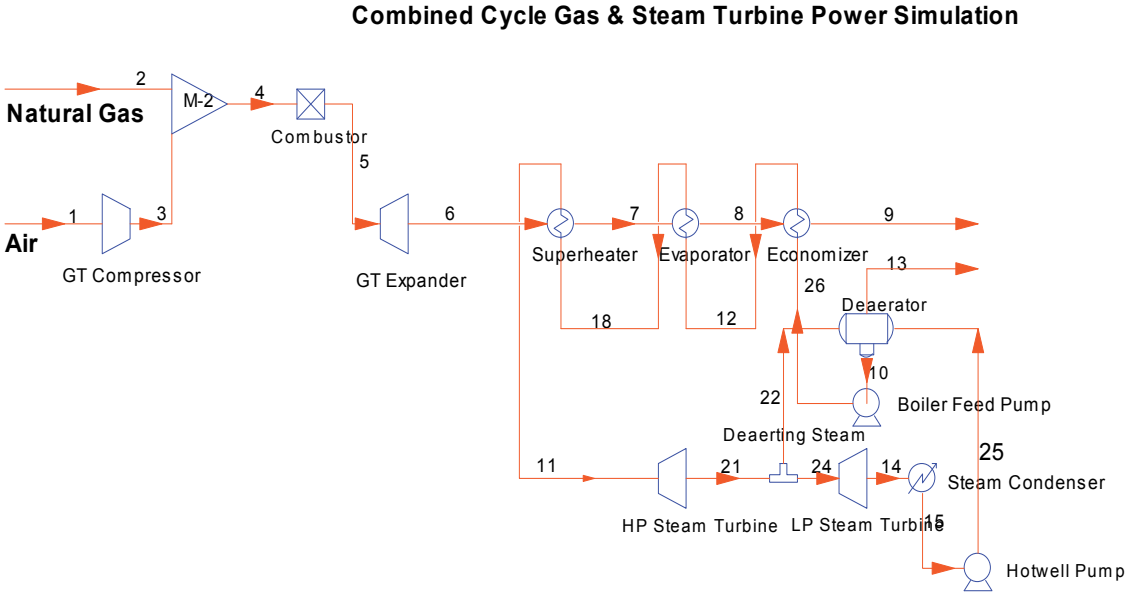
CO2flash_PPRK.psd – Predictive Peng Robinson

This sample uses Predictive Peng-Robinson correlation enhances the characterization of highly non-ideal systems over a wide range of temperature and pressure. It is generalized as a linear function of acentric factor used for the prediction of liquid densities for heavy hydrocarbons and polar components.



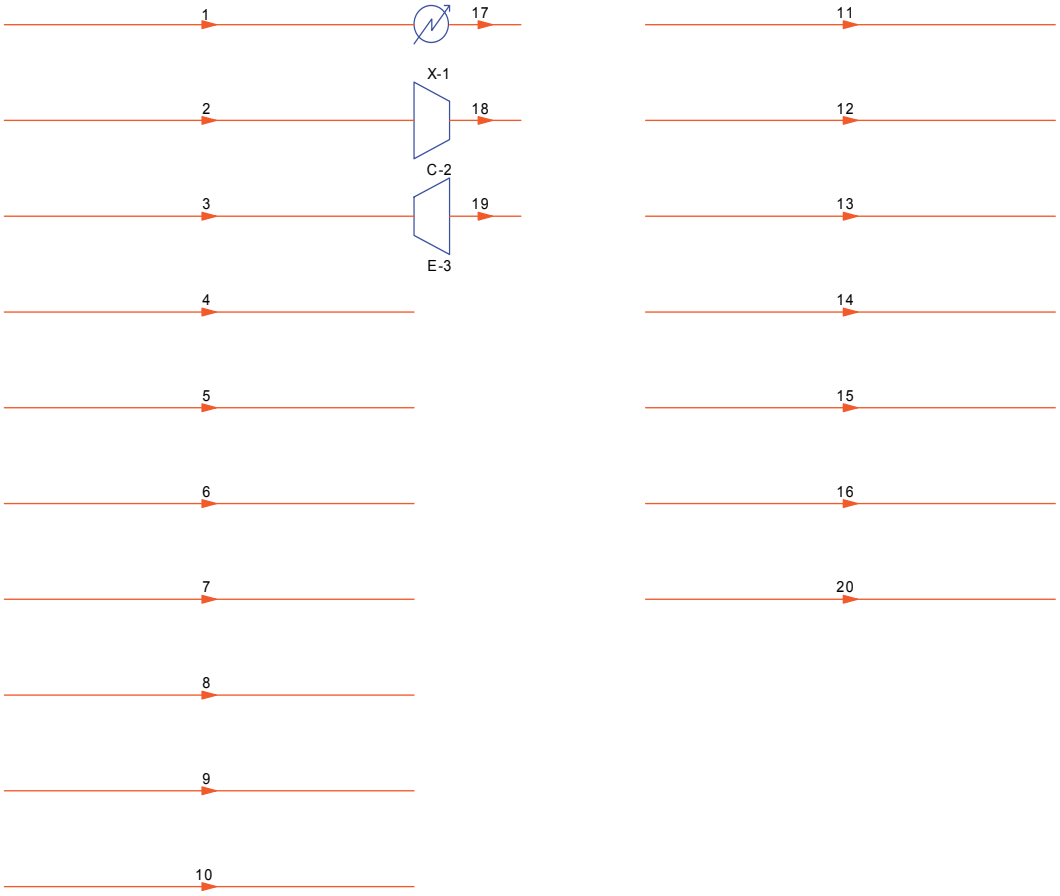
Redlich_Kwong.psd – Gas & Steam Turbine Power Simulation

This sample uses Redlich-Kwong equation of state with interaction coefficients for CO2 and H2S developed by G. W. Wilson is a cubic equation of state which can be used to model light hydrocarbon mixtures such as cryogenic ethane and propane recovery.



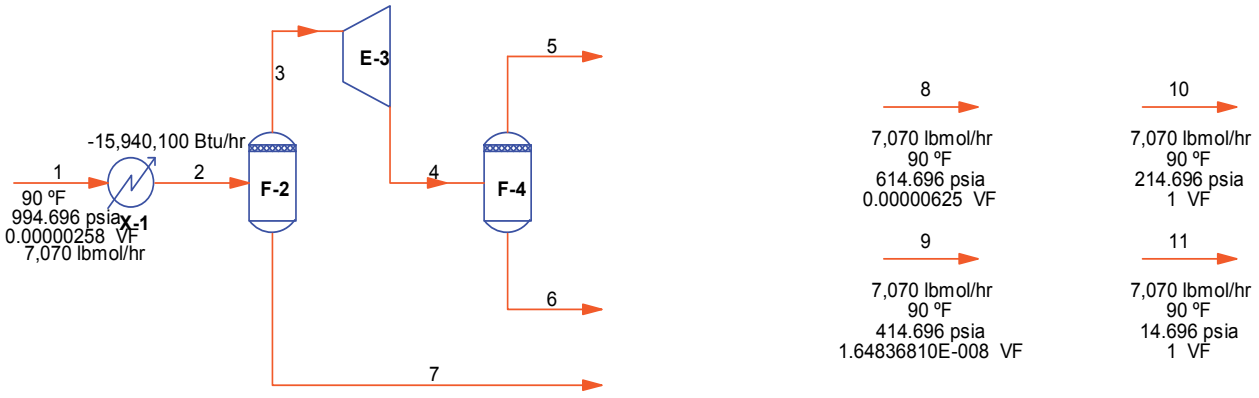
refprop.psd – REFerence fluid PROPerties Method

This sample uses REFPROP which is an acronym for REFerence fluid PROPerties. This program, developed by the National Institute of Standards and Technology (NIST), provides tables and plots of the thermodynamic and transport properties of industrially important fluids and their mixtures with an emphasis on refrigerants and hydrocarbons, especially natural gas systems.



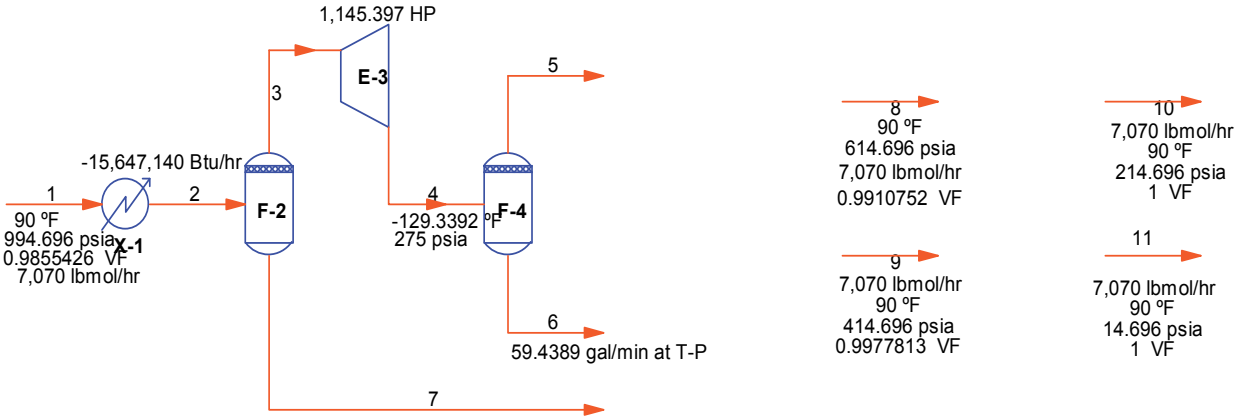
refprop-aga.psd – REFerence fluid PROPerties with AGA Method

This sample uses REFPROP which is an acronym for REFerence fluid PROPerties along with AGA method. This program, developed by the National Institute of Standards and Technology (NIST), provides tables and plots of the thermodynamic and transport properties of industrially important fluids and their mixtures with an emphasis on refrigerants and hydrocarbons, especially natural gas systems.



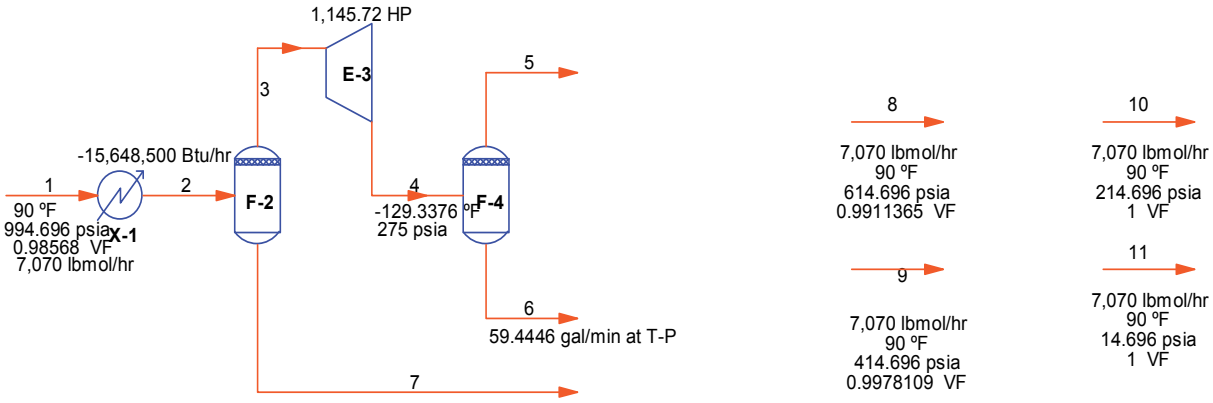
refprop-default.psd – REFerence fluid PROPerties Method

This sample uses REFPROP which is an acronym for REFerence fluid PROPerties. This program, developed by the National Institute of Standards and Technology (NIST), provides tables and plots of the thermodynamic and transport properties of industrially important fluids and their mixtures with an emphasis on refrigerants and hydrocarbons, especially natural gas systems.



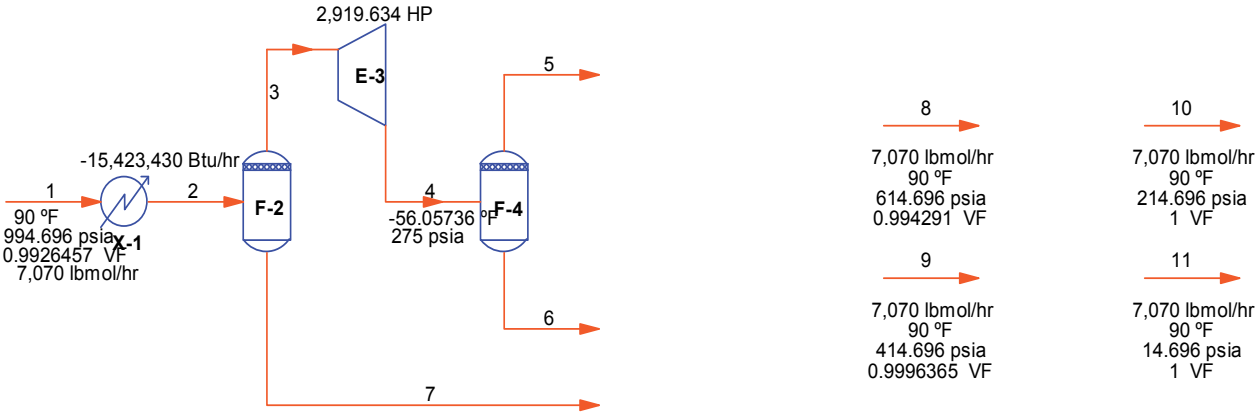
refprop-gerg.psd – REFerence fluid PROPerties with GERG Method

This sample uses REFPROP which is an acronym for REFerence fluid PROPerties along with its own GERG method. This program, developed by the National Institute of Standards and Technology (NIST), provides tables and plots of the thermodynamic and transport properties of industrially important fluids and their mixtures with an emphasis on refrigerants and hydrocarbons, especially natural gas systems.



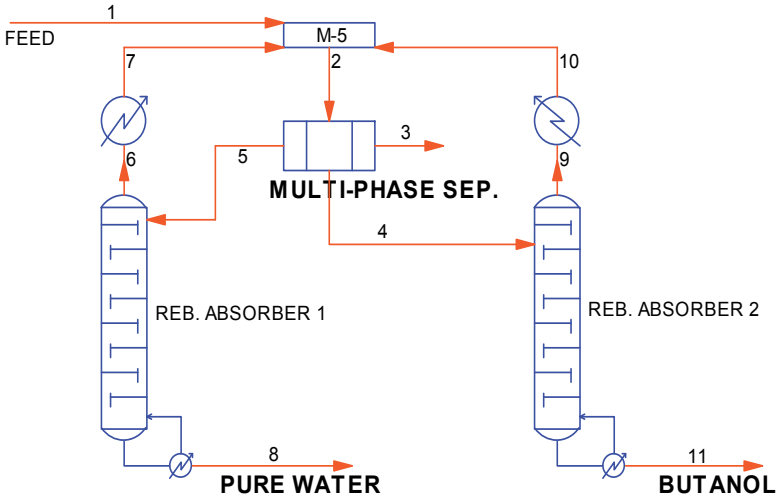
refprop-pengrob.psd – REFerence fluid PROPerties with PENK Method

This sample uses REFPROP which is an acronym for REFerence fluid PROPerties along with its own Peng Robinson method. This program, developed by the National Institute of Standards and Technology (NIST), provides tables and plots of the thermodynamic and transport properties of industrially important fluids and their mixtures with an emphasis on refrigerants and hydrocarbons, especially natural gas systems.



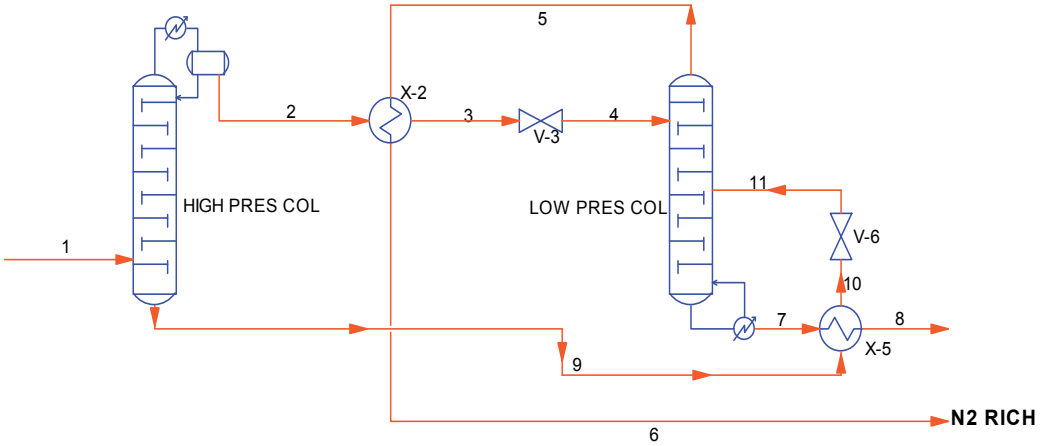
Renon (NRTL).psd – Azeotropic Distillation of Butanol/Water

This sample uses Renon thermodynamic method which is also known as the Non-Random Two-Liquid equation (NRTL). This is an activity coefficient correlation for modeling non-ideal systems which can be partly or totally miscible.



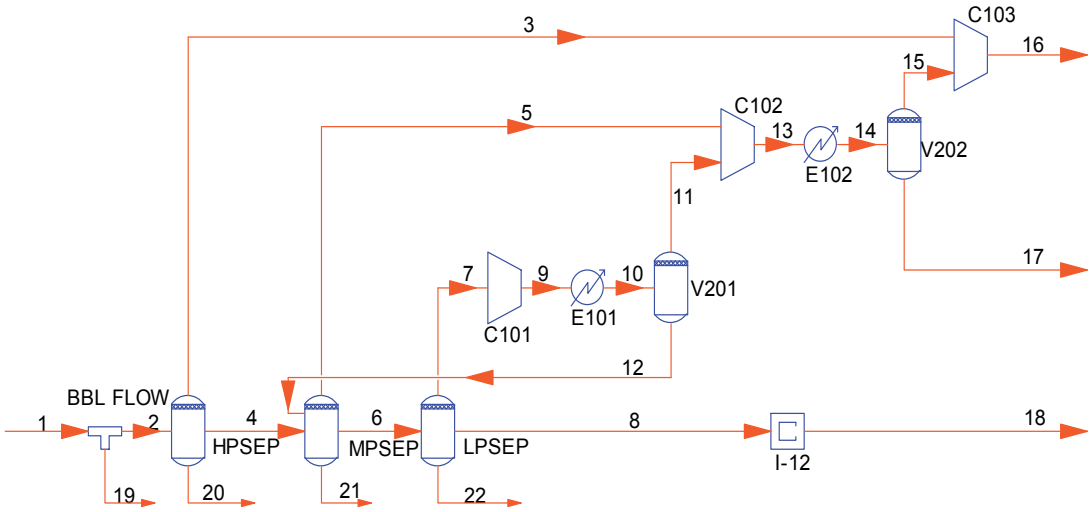
KVAL.psd – Nitrogen Rejection Unit

This sample uses the correlation that is based on data for methane systems and can be used for natural gas processes where methane is the predominant component for temperatures between 100F and -300 F (e.g. nitrogen rejection, cryogenic expander plants).



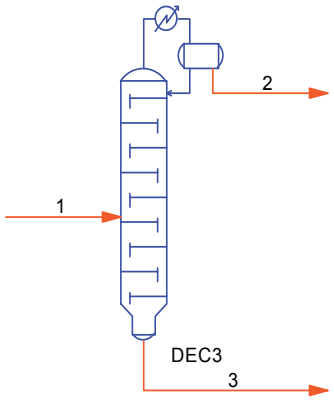
Gosp_SKD.psd – Gas Oil Separation Plant

This sample uses the Soave-Kabadi-Danner method which is based upon a modified Soave-Redlich-Kwong equation of state. This method applies only to hydrocarbon-water systems. If water is not present in the system, then the SKDK method defaults to APISOAVEK.



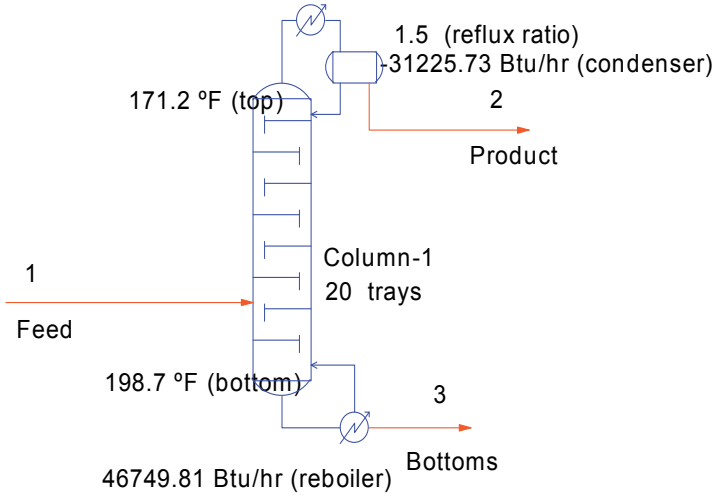
Soave (SRK).psd – Free Water Check in Refine Column Simulation

This sample uses Soave modification of the Redlich-Kwong cubic equation of state is most typically used for light hydrocarbon systems from cryogenic temperatures to 900F (or below the critical temperature of the mixture) for pressures up to 1000 PSIA.



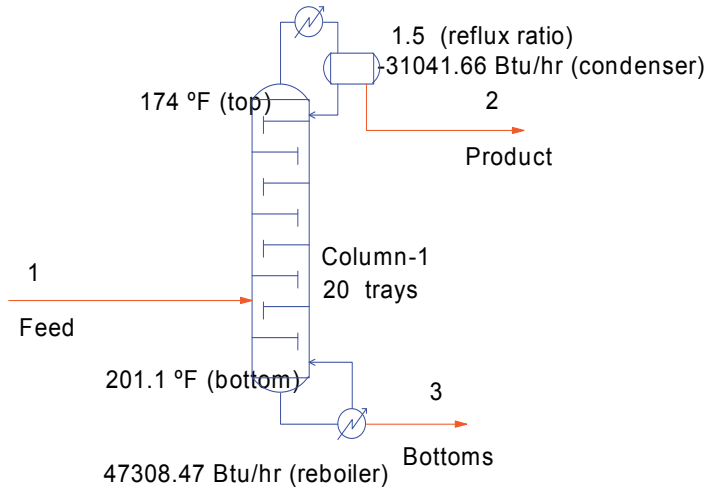
Ethanol_h2o_unifac.psd – Ethanol-Water Distillation

This is an ethanol water simulation using group contribution theory developed by J.M. Prausnitz, et.al. and extended by A. Fredenslund, et.al. It generates estimates of the non-ideal interactions between binary pairs of components based on the "groups" contained in each component.



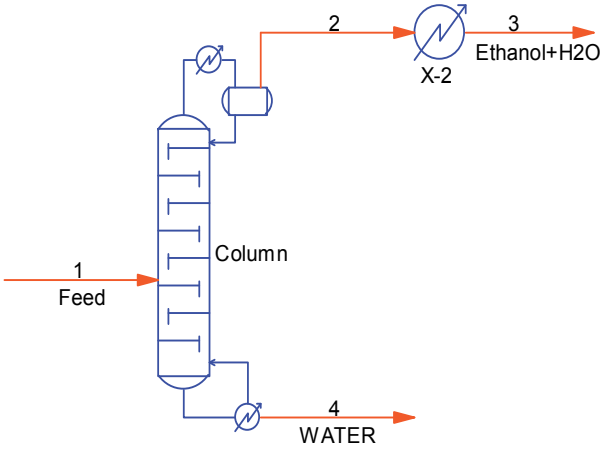
Ethanol_h2o_unifac lle.psd – Ethanol-Water Distillation

This is an ethanol water simulation using group contribution theory developed by J.M. Prausnitz, et.al. and extended by A. Fredenslund, et.al. It generates estimates of the non-ideal interactions between binary pairs of components based on the "groups" contained in each component. The UNIFAC model used to predict liquid-liquid phase behavior is the same as that used to predict vapor-liquid equilibrium (UNIFACK). However, the interaction parameters are different. UNIFAC LLE group interactions are available for 32 different groups representing hydrocarbons, water, alcohols, organic acids, and other compounds.



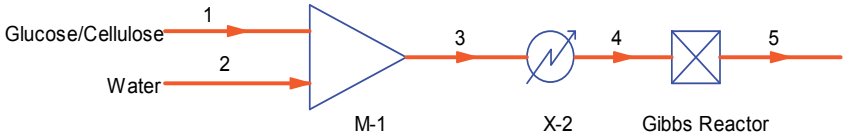
Ethanol_Water_Uniquac.psd – Ethanol-Water Distillation

This is an ethanol water simulation using the UNIversal QUAsi-Chemical equation can correlate vapor-liquid equilibrium data for partially and totally miscible systems. The UNIQUAC equation extends the Guggenheim theory of quasi-chemical representation for liquid mixtures by introducing Q (molecular volume) and R (molecular area) parameters and utilizes the Wilson concept of local composition. The main advantage of the UNIQUAC equation is its ability to represent both VLE and LLE with only two adjustable parameters per binary.



ChemTran_Glucose_Vapor_Pressure.psd – Vapor Pressure Method

This is an illustration of vapor pressure method using ChemTran defined Glucose and Cellulose species for their conversion to Ethanol via Gibbs Free energy minimization reactor.



HCL_Water_Wilson.psd – Using Wilson Thermodynamic Method

This is a sample using Wilson correlation which is an activity coefficient equation used for modeling non-ideal, miscible mixtures (CHEmtran FILe recommended). Vapor phase fugacities are calculated using Redlich-Kwong equation of state.