CHAPTER VII

DYNAMIC SIMULATION

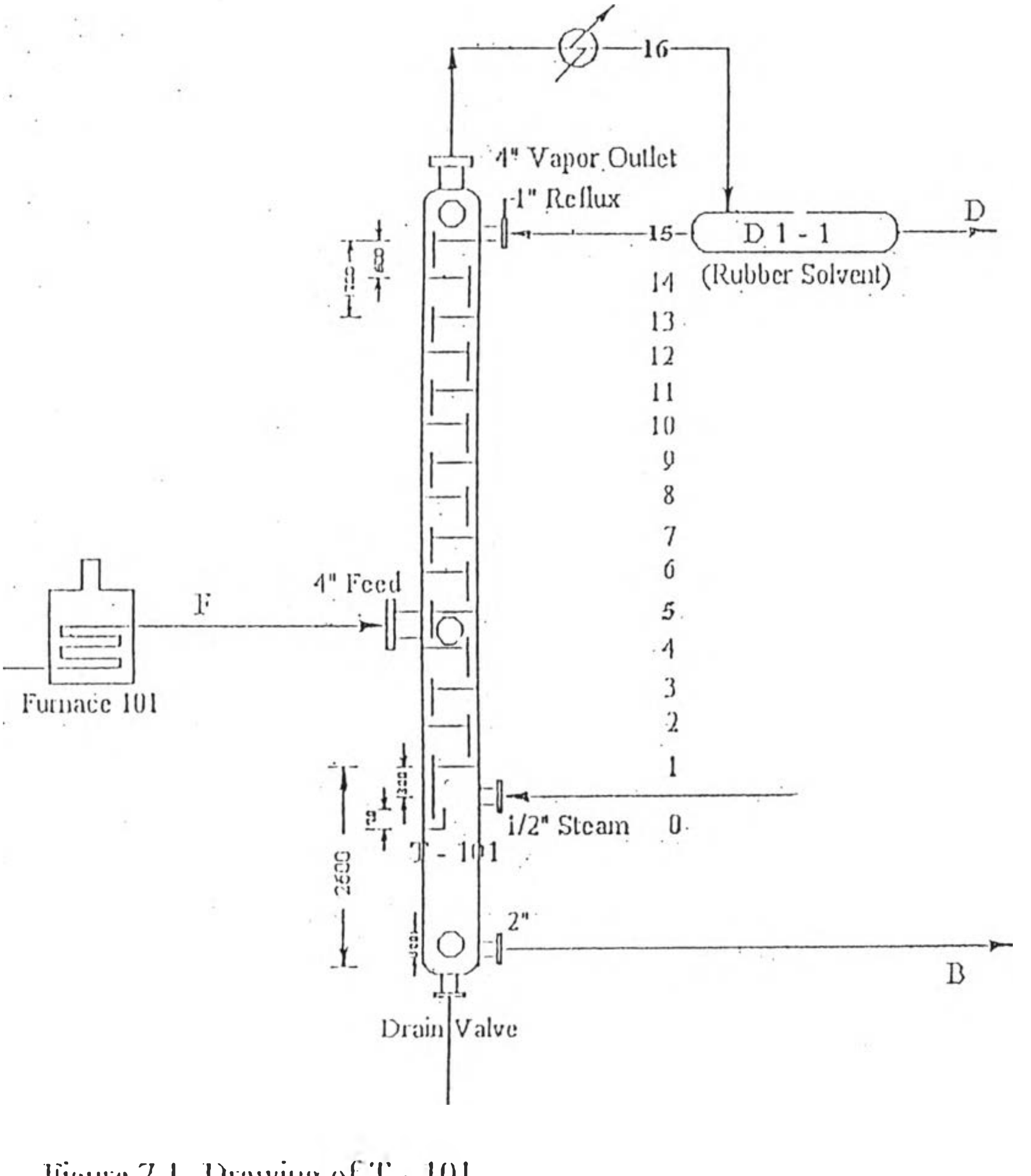

Figure 7.1 Drawing of T - 101

7.1 Changing of Feed Inlet Temperature

Feed inlet temperature 140 ^{O}C

Operating condition :			
	Pressure	14.7	psia
	Feed inlet temperature	140	^{O}C
	Product output temperature	89	^{O}C
	Bottom output temperature	128	^{O}C
	Feed flowrate	45	GPM
	Distillate (Rubber Solvent)	3.9	GPM
	Reflux feed rate	21.7	GPM

Table 7.1 Equilibrium Constant vs. Temperature

Tray No.	Temperature (^{O}C)	K_{C5}	K_{C6}	K_{C7}	K_{C8}
15	89.00	3.83	1.65	0.70	0.30
14	91.60	4.03	1.76	0.75	0.32
13	94.20	4.24	1.86	0.80	0.35
12	96.80	4.45	1.97	0.86	0.38
11	99.40	4.67	2.08	0.91	0.41
10	102.00	4.90	2.20	0.97	0.44
9	104.60	5.13	2.32	1.03	0.48
8	107.20	5.36	2.45	1.10	0.52
7	109.80	5.60	2.58	1.16	0.56
6	112.40	5.85	2.71	1.23	0.60
5	115.00	6.10	2.85	1.30	0.65
4	117.60	6.36	2.99	1.37	0.70
3	120.20	6.63	3.13	1.45	0.75
2	122.80	6.90	3.28	1.53	0.81
1	125.40	7.17	3.43	1.61	0.87
0	128.00	7.45	3.59	1.69	0.93

Table 7.2 Equilibrium Constant vs. Temperature (con't)

Tray No.	Temperature (^{O}C)	K_{C9}	K_{C10}	K_{C11}
15	89.00	0.118	0.045	0.0452
14	91.60	0.128	0.050	0.0503
13	94.20	0.140	0.056	0.0559
12	96.80	0.151	0.062	0.0620
11	99.40	0.164	0.069	0.0687
10	102.00	0.178	0.076	0.0760
9	104.60	0.192	0.084	0.0839
8	107.20	0.207	0.093	0.0926
7	109.80	0.222	0.102	0.1020
6	112.40	0.239	0.112	0.1122
5	115.00	0.257	0.123	0.1233
4	117.60	0.275	0.135	0.1353
3	120.20	0.295	0.148	0.1484
2	122.80	0.315	0.162	0.1624
1	125.40	0.336	0.178	0.1776
0	128.00	0.358	0.194	0.1940

Product composition (from GC) : D = 3.9 GPM

C_5	0.0175
C_6	0.3290
C_7	0.5977
C_8	0.0534
C_9	0
C_{10}	0
C_{11}	0
$>C_{11}$	0.0006
Polynaphthene	0.0001

Bottom composition (from GC) : B = 41.04 GPM

C_5	0.0008
C_6	0.0105
C_7	0.0623
C_8	0.2289
C_9	0.2807
C_{10}	0.1804
C_{11}	0.0620
$>C_{11}$	0.1648
Polynaphthene	0.0097

Feed composition (from GC) : F = 45 GPM

C_5	0.0161
C_6	0.3018
C_7	0.5521
C_8	0.0687
C_9	0.0244
C_{10}	0.0157
C_{11}	0.0054
$>C_{11}$	0.0149
Polynaphthene	0.0009

From experiment :

L =	21.7	GPM
Density of Rubber solvent =	0.7397	
Mass =	0.0608	kg/min.
MW. of Rubber solvent =	0.0954	kg/mol
L =	0.6368	mol/min.
D =	3.9	GPM
D =	0.1144	mol/min.
V =	L + D	
V =	0.7512	mol/min.

F =	45	GPM
Density of feed =	0.7531	
Mass =	0.1283	kg/min.
MW. of feed =	0.0963	kg/mol
F =	1.3313	mol/min.
B =	F - D	
B =	1.2169	mol/min.

Component A is a light non-key for this case is C_5 - C_{11}
Component B is the light key which choose component C_{11} as the reference
The operating equations for rectifying section are:

$$y_{i,j-1} = (L/V)X_{i,j} + (1 - L/V)X_{i,dist}$$

For the stripping section are :

$$y_{i,k-1} = (L'/V')X_{i,k} + (L'/V' - 1)X_{i,bot}$$

Constant relative volatility systems

$$\alpha_{AB} = K_A/K_B$$

For any component i on stage j

$$X_{ij} = y_{ij}/(\alpha_{iB,j} * K_{Bj})$$

$$K_{Bj} = \Sigma (y_{ij} / \alpha_{iB})$$

Table 7.3 Relative Volatility

Tray	α C_5C_{11}	α C_6C_{11}	α C_7C_{11}	α C_8C_{11}	α C_9C_{11}	α $C_{10}C_{11}$	α $C_{11}C_{11}$
15	84.81	36.60	15.56	6.55	2.61	1.00	1
14	80.21	34.91	14.96	6.40	2.55	1.00	1
13	75.89	33.30	14.38	6.26	2.50	1.00	1
12	71.84	31.78	13.83	6.12	2.44	1.00	1
11	68.03	30.34	13.30	5.98	2.39	1.00	1
10	64.45	28.97	12.79	5.85	2.34	1.00	1
9	61.08	27.66	12.30	5.73	2.28	1.00	1
8	57.92	26.43	11.83	5.61	2.23	1.00	1
7	54.94	25.25	11.39	5.49	2.18	1.00	1
6	52.14	24.14	10.96	5.38	2.13	1.00	1
5	49.50	23.08	10.54	5.27	2.08	1.00	1
4	47.01	22.07	10.15	5.17	2.03	1.00	1
3	44.67	21.11	9.77	5.06	1.99	1.00	1
2	42.46	20.20	9.40	4.97	1.94	1.00	1
1	40.37	19.34	9.06	4.87	1.89	1.00	1

Table 7.4 Multicomponent Distillation Calculation for Vapor Phase

Tray (j)	$y_{C5,j-1}$	$y_{C6,j-1}$	$y_{C7,j-1}$	$y_{C8,j-1}$	$y_{C9,j-1}$	$y_{C10,j-1}$	$y_{C11,j-1}$
15	0.0175	0.3290	0.5977	0.0534	0.0000	0.0000	0.0000
14	0.0059	0.1880	0.6756	0.1302	0.0000	0.0000	0.0000
13	0.0036	0.1152	0.6327	0.2481	0.0000	0.0000	0.0000
12	0.0031	0.0843	0.5220	0.3902	0.0000	0.0000	0.0000
11	0.0030	0.0721	0.4012	0.5233	0.0000	0.0000	0.0000
10	0.0030	0.0672	0.3067	0.6227	0.0000	0.0000	0.0000
9	0.0030	0.0653	0.2463	0.6851	0.0000	0.0000	0.0000
8	0.0030	0.0645	0.2123	0.7198	0.0000	0.0000	0.0000
7	0.0030	0.0643	0.1948	0.7375	0.0000	0.0000	0.0000
6	0.0030	0.0645	0.1867	0.7455	0.0000	0.0000	0.0000
5	0.0228	0.3757	0.5399	0.0482	0.0092	0.0032	0.0011
4	0.0049	0.1684	0.5311	0.1272	0.0881	0.0598	0.0205
3	0.0006	0.0344	0.2328	0.1389	0.2260	0.2734	0.0939
2	0.0001	0.0044	0.0498	0.0811	0.2325	0.4705	0.1616
1	0.0001	0.0019	0.0159	0.0545	0.1815	0.5553	0.1908

Table 7.5 Multicomponent Distillation Calculation for Liquid Phase

Tray (j)	$x_{C5,j}$	$x_{C6,j}$	$x_{C7,j}$	$x_{C8,j}$	$x_{C9,j}$	$x_{C10,j}$	$x_{C11,j}$
15	0.0175	0.3290	0.5977	0.0534	0.0000	0.0000	0.0000
14	0.0038	0.1627	0.6896	0.1440	0.0000	0.0000	0.0000
13	0.0011	0.0768	0.6390	0.2831	0.0000	0.0000	0.0000
12	0.0006	0.0403	0.5085	0.4507	0.0000	0.0000	0.0000
11	0.0004	0.0259	0.3659	0.6078	0.0000	0.0000	0.0000
10	0.0004	0.0202	0.2544	0.7250	0.0000	0.0000	0.0000
9	0.0004	0.0179	0.1832	0.7986	0.0000	0.0000	0.0000
8	0.0004	0.0170	0.1431	0.8396	0.0000	0.0000	0.0000
7	0.0004	0.0168	0.1224	0.8604	0.0000	0.0000	0.0000
6	0.0004	0.0169	0.1129	0.8699	0.0000	0.0000	0.0000
5	0.0054	0.1897	0.5967	0.1065	0.0513	0.0375	0.0129
4	0.0055	0.1915	0.5987	0.1050	0.0507	0.0362	0.0124
3	0.0005	0.0376	0.2562	0.1184	0.2090	0.2816	0.0967
2	0.0000	0.0032	0.0460	0.0520	0.2165	0.5079	0.1745
1	0.0000	0.0003	0.0071	0.0214	0.1580	0.6052	0.2079

Flash Distillation for Feed (Tray No. 5)

First Guess :

V/F = 1

(V/F) 0.079278

(df/d(V/F)) = 1.456692

next guess for V/F = 1.054424

In this case V/F = 1 gives, f (V/F) = close to zero

Therefore all feed fraction are evaporated.

So, V' = V = 0.75121

$L' = L + L_F$ 0.63677

F = 1.38797

Table 7.6 Enthalpy for nth Tray

Tray	ΔH_{C5}	ΔH_{C6}	ΔH_{C7}	ΔH_{C8}	ΔH_{C9}	ΔH_{C10}	ΔH_{C11}
15	-8E+09	-3E+12	-1.3E+13	-2E+11	0	0	0
14	-2E+09	-2E+12	-1.6E+13	-5E+11	0	0	0
13	-1E+09	-1E+12	-1.7E+13	-1E+12	0	0	0
12	-1E+09	-8E+11	-1.5E+13	-2E+12	0	0	0
11	-1E+09	-7E+11	-1.2E+13	-3E+12	0	0	0
10	-1E+09	-7E+11	-9.6E+12	-3E+12	0	0	0
9	-1E+09	-7E+11	-8E+12	-4E+12	0	0	0
8	-1E+09	-8E+11	-7.2E+12	-4E+12	0	0	0
7	-2E+09	-8E+11	-6.9E+12	-5E+12	0	0	0
6	-2E+09	-9E+11	-7.1E+12	-5E+12	0	0	0
5	-2E+10	-7E+12	-2.9E+13	-6E+11	-9.4E+09	-4E+09	-5E+08
4	-6E+09	-5E+12	-3.1E+13	-9E+11	-2.3E+10	-1E+10	-1E+09
3	-7E+08	-1E+12	-1.4E+13	-1E+12	-7.8E+10	-6E+10	-8E+09
2	-1E+08	-1E+11	-3E+12	-6E+11	-8.6E+10	-1E+11	-1E+10
1	-9E+07	-3E+10	-7.8E+11	-4E+11	-7E+10	-2E+11	-2E+10

The other operating conditions are shown in Appendix B

7.2 Multicomponent Nonideal Distillation Column

The assumptions are :

1. Liquid on the tray is perfectly mixed and incompressible.
2. Tray vapor holdups are negligible.
3. Dynamics of the condenser and the reboiler will be neglected.
4. Vapor and Liquid are in thermal equilibrium but not in phase equilibrium.

A general nth tray is sketched in Figure 7.2

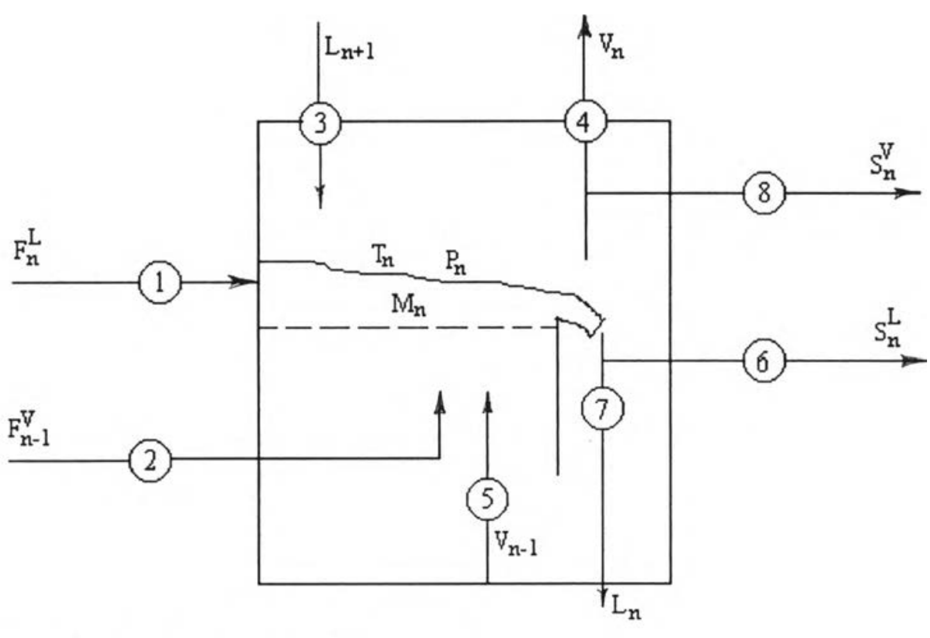

Figure 7.2 n^{th} tray of multicomponent column

The equation describing this tray is :

$$\frac{dM_n}{dt} = L_{n+1} + F_n^L + F_{n-1}^V + V_{n-1} - V_n - L_n - S_n^L - S_n^V \qquad (7.1)$$

$$\frac{dM_{15}}{dt} = L_{16} + F_{15}^{L} + F_{14}^{V} + V_{14} - V_{15} - L_{15} - S_{15}^{L} - S_{15}^{V}$$

Where: L_{16} = 0

L_{15} = Reflux flow rate = 21.7 GPM

F_{15}^{L} = No liquid inlet @ tray 15 = 0

F_{14}^{V} = No vapor inlet @ tray 14 = 0

V_{14} = $L_{14}/K_{\text{operating Line}}$ = $L_{14}/1.688$

V_{15} = $L_{15}/K_{\text{operating Line}}$ = 2.3099/1.688

= 1.3684

L_{14} = Liquid over weir = h_{ow}

= $0.48\ F_{w}^{*}\ (Q/l_{w})^{2/3}$

Fw = Weir constriction correction factor (From Chart)

$Q/(l_{w})^{2.5}$ = Liquid load/Weir length = 21.7 GPM/(2.133 ft)$^{2.5}$

= 3.2657

Ratio weir length/Tower diameter = 650/850 = 0.7647

Therefore: F_{w} = 1.025

L_{15} = 2.3099

S_{15}^{L} = No side-draw of tray 15 = 0

S_{15}^{V} = Flow of product = 12.857 GPM

Table 7.7 Streams on n^{th} tray

Tray (n)	L_n	F_n^L	F_n^V	V_n	S_n^L	S_n^V
16	0	0	0	0	0	0
15	0.0006	0	0	0.0013	0	0.0007
14	0.0006	0	0	0.0013	0	0
13	0.0006	0	0	0.0013	0	0
12	0.0006	0	0	0.0013	0	0
11	0.0006	0	0	0.0013	0	0
10	0.0006	0	0	0.0013	0	0
9	0.0006	0	0	0.0013	0	0
8	0.0006	0	0	0.0013	0	0
7	0.0006	0	0	0.0013	0	0
6	0.0006	0	0	0.0013	0	0
5	0.0018	0.0012	0	0.0013	0	0
4	0.0018	0	0	0.0013	0	0
3	0.0018	0	0	0.0013	0	0
2	0.0018	0	0	0.0013	0	0
1	0.0018	0	0	0.0013	0.0005	0
0	0.0018	0	0	0.0013	0	0

7.3 Mathematical Models of Multicomponent Column

Tray 15 @ C_5 composition

Total continuity :

$$\frac{dM_n}{dt} = L_{n+1} + F_n^{L} + F_{n-1}^{V} + V_{n-1} - V_n - L_n - S_n^{L} - S_n^{V} \quad (7.1)$$

$$\frac{dM_{15}}{dt} = L_{16} + F_{15}^{L} + F_{14}^{V} + V_{14} - V_{15} - L_{15} - S_{15}^{L} - S_{15}^{V} \quad (7.2)$$

Component continuity equations :

$$\frac{d(M_nX_{nj})}{dt} = L_{n+1}X_{n+1,j} + F_n^{L}X_{n,j}^{F} + F_{n-1}^{V}y_{n-1,j}^{F} + V_{n-1}Y_{n-1,j} - V_nY_{n,j} - L_nX_{n,j} - S_n^{L}X_{n,j} - S_n^{V}Y_{n,j} \quad (7.3)$$

$$\frac{d(M_{15}X_{15,C5})}{dt} = L_{16}X_{16,C5} + F_{15}^{L}X_{15,C5}^{F} + F_{14}^{V}y_{14,C5}^{F} + V_{14}Y_{14,C5} - V_{15}Y_{15,C5} - L_{15}X_{15,C5} - S_{15}^{L}X_{15,C5} - S_{15}^{V}Y_{15,C5} \quad (7.4)$$

$$\frac{d(M_{15}X_{15,C6})}{dt} = L_{16}X_{16,C6} + F_{15}^{L}X_{15,C6}^{F} + F_{14}^{V}y_{14,C6}^{F} + V_{14}Y_{14,C6} - V_{15}Y_{15,C6} - L_{15}X_{15,C6} - S_{15}^{L}X_{15,C6} - S_{15}^{V}Y_{15,C6} \quad (7.5)$$

$$\frac{d(M_{15}X_{15,C7})}{dt} = L_{16}X_{16,C7} + F_{15}^{L}X_{15,C7}^{F} + F_{14}^{V}y_{14,C7}^{F} + V_{14}Y_{14,C7} - V_{15}Y_{15,C7} - L_{15}X_{15,C7} - S_{15}^{L}X_{15,C7} - S_{15}^{V}Y_{15,C7} \quad (7.6)$$

$$\frac{d(M_{15}X_{15,C8})}{dt} = L_{16}X_{16,C8} + F_{15}^{L}X_{15,C8}^{F} + F_{14}^{V}y_{14,C8}^{F} + V_{14}Y_{14,C8} - V_{15}Y_{15,C8} - L_{15}X_{15,C8} - S_{15}^{L}X_{15,C8} - S_{15}^{V}Y_{15,C8} \quad (7.7)$$

$$\frac{d(M_{15}X_{15,C9})}{dt} = L_{16}X_{16,C9} + F_{15}{}^{L}X_{15,C9}{}^{F} + F_{14}{}^{V}y_{14,C9}{}^{F} + V_{14}Y_{14,C9}$$

$$- V_{15}Y_{15,C9} - L_{15}X_{15,C9} - S_{15}{}^{L}X_{15,C9} - S_{15}{}^{V}Y_{15,C9} \quad (7.8)$$

$$\frac{d(M_{15}X_{15,C10})}{dt} = L_{16}X_{16,C10} + F_{15}{}^{L}X_{15,C10}{}^{F} + F_{14}{}^{V}y_{14,C10}{}^{F} + V_{14}Y_{14,C10}$$

$$- V_{15}Y_{15,C10} - L_{15}X_{15,C10} - S_{15}{}^{L}X_{15,C10} - S_{15}{}^{V}Y_{15,C10} \quad (7.9)$$

$$\frac{d(M_{15}X_{15,C11})}{dt} = L_{16}X_{16,C11} + F_{15}{}^{L}X_{15,C11}{}^{F} + F_{14}{}^{V}y_{14,C11}{}^{F} + V_{14}Y_{14,C11}$$

$$- V_{15}Y_{15,C11} - L_{15}X_{15,C11} - S_{15}{}^{L}X_{15,C11} - S_{15}{}^{V}Y_{15,C11} \quad (7.10)$$

Energy equation :

$$\frac{d(M_n h_n)}{dt} = L_{n+1}h_{n+1} + F_n{}^{L}h_n{}^{F} + F_{n-1}{}^{V}H_{n-1}{}^{F} + V_{n-1}H_{n-1} - V_nH_n$$

$$- L_nh_n - S_n{}^{L}h_n - S_n{}^{V}H_n \quad (7.11)$$

$$\frac{d(M_{15}h_{15})}{dt} = L_{16}h_{16} + F_{15}{}^{L}h_{15}{}^{F} + F_{14}{}^{V}H_{14}{}^{F} + V_{14}H_{14} - V_{15}H_{15}$$

$$- L_{15}h_{15} - S_{15}{}^{L}h_{15} - S_{15}{}^{V}H_{15} \quad (7.12)$$

There are 15 equations to solve for the multicomponent nonideal distillation column per tray. The following are the results from solving equations using Mathcad program.

7.4 Solving Dynamic Simulation by Mathcad Program

The total continuity equations are written in terms of moles per unit time. The multi-component nonideal distillation column differential equations were solved by Runge-Kutta technique in the Mathcad program as shown in appendix A. The computational procedures are as follows.

1. Set the mass balance value of reflux, multiple feeds of both liquid and vapor, and sidestream drawoffs.
2. Set the equations for calculating bubble point temperature of each tray to find the exceptional Raoult's law K-values which give $\Sigma y_i = 1$ ($y_i = K_i Z_i$) and use these K-values to determine the vapor component on every tray.
3. Set the equations for dew point temperature of each tray to find the exceptional Raoult's law K-values which give $\Sigma x_i = 1$ ($x_i = Z_i/K_i$) and use these K-values to determine the enthalpy of vapor on every tray.
4. Compute the hydrocarbon mole fractions for every tray by set the equations in Mathcad program which consists of 3 main equations ; component continuity equations, energy equations, and total continuity equations.
5. Calculate mole fractions and enthalpies by looping in Mathcad respect to time for make the data more accuracy.
6. Repeat step 2 and 3 with a corrector step for the same time increment. In this case every time step (5 min.) found that there is no disturbance to a process of dew point and bubble point temperatures. Therefore, in this case assume the every step change there is a constant temperature on each tray.

The results of dynamic distillation are shown below.

Table 7.8 Results for multicomponent nonideal dynamic-distillation

Time	Mole fraction of component on tray 15						
(min.)	C_5	C_6	C_7	C_8	C_9	C_{10}	C_{11}
0	0.0175	0.3290	0.5977	0.0534	0.0000	0.0000	0.0000
10	0.0192	0.3076	0.5960	0.0655	0.0039	0.0039	0.0039
20	0.0203	0.2900	0.5950	0.0767	0.0060	0.0060	0.0060
30	0.0212	0.2741	0.5915	0.0882	0.0084	0.0083	0.0084
40	0.0151	0.2622	0.5902	0.1004	0.0107	0.0107	0.0107
60	0.0180	0.2384	0.5749	0.1216	0.0157	0.0157	0.0157
90	0.0233	0.2103	0.5483	0.1510	0.0226	0.0219	0.0226
120	0.0283	0.1881	0.5212	0.1782	0.0283	0.0276	0.0283

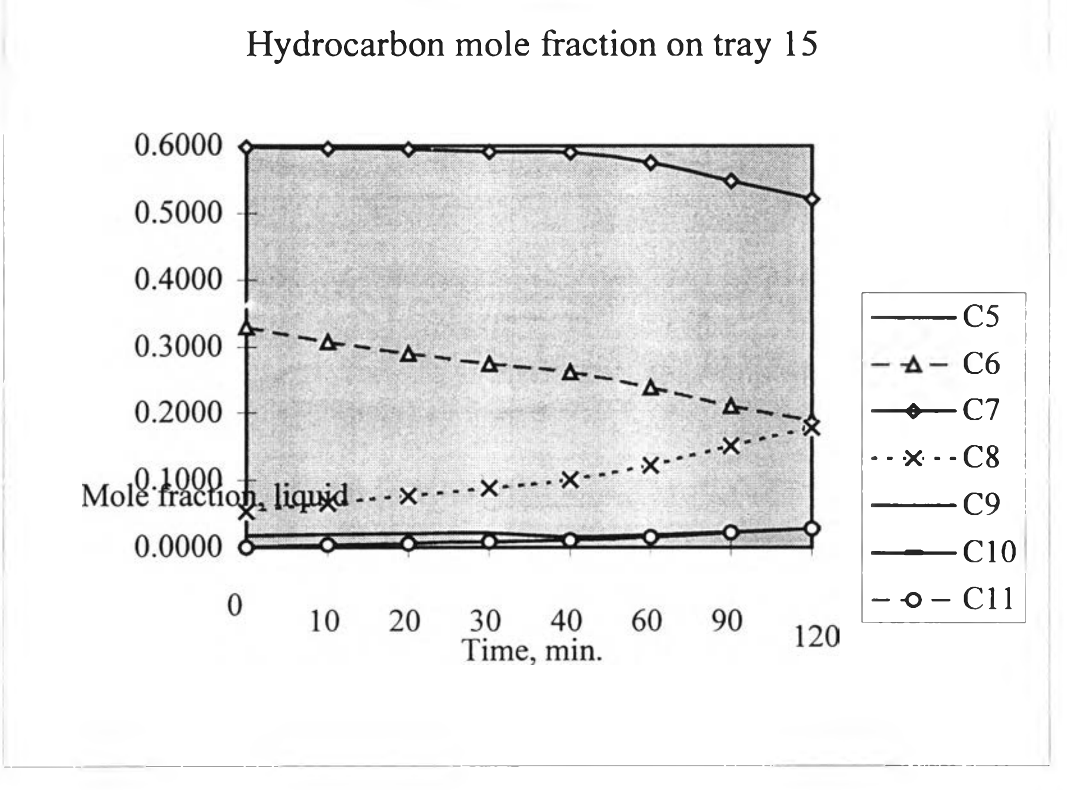

Figure 7.3 Plot of hydrocarbon mole fraction on tray 15.

Table 7.9 Results for multicomponent nonideal dynamic-distillation

Time (min.)	Mole fraction of component on tray 10						
	C_5	C_6	C_7	C_8	C_9	C_{10}	C_{11}
0	0.0004	0.0202	0.2544	0.7250	0.0000	0.0000	0.0000
10	0.0004	0.0205	0.2603	0.7186	0.0000	0.0000	0.0000
20	0.0004	0.0205	0.2662	0.7127	0.0001	0.0001	0.0001
30	0.0004	0.0205	0.2720	0.7068	0.0001	0.0001	0.0001
40	0.0004	0.0205	0.2778	0.7009	0.0001	0.0001	0.0001
60	0.0006	0.0199	0.2901	0.6874	0.0006	0.0006	0.0006
90	0.0035	0.0192	0.3077	0.6603	0.0033	0.0024	0.0036
120	0.0109	0.0218	0.3211	0.6204	0.0086	0.0073	0.0100

Table 7.10 Results for multicomponent nonideal dynamic-distillation

Time (min.)	Mole fraction of component on tray 5						
	C_5	C_6	C_7	C_8	C_9	C_{10}	C_{11}
0	0.0054	0.1897	0.5967	0.1065	0.0513	0.0375	0.0129
10	0.0065	0.0227	0.7116	0.1261	0.0648	0.0507	0.0177
20	0.0064	0.0448	0.6845	0.1214	0.0654	0.0570	0.0205
30	0.0064	0.0587	0.6627	0.1174	0.0652	0.0652	0.0242
40	0.0056	0.0712	0.6429	0.1128	0.0647	0.0749	0.0277
60	0.0055	0.0959	0.6033	0.1018	0.0623	0.0959	0.0353
90	0.0061	0.1332	0.5513	0.0850	0.0574	0.1217	0.0452
120	0.0067	0.1606	0.5184	0.0711	0.0535	0.1384	0.0512

Table 7.11 Results for multicomponent nonideal dynamic-distillation

Time (min.)	Mole fraction of component on bottom						
	C_5	C_6	C_7	C_8	C_9	C_{10}	C_{11}
0	0.00000	0.00030	0.00710	0.02140	0.15800	0.60520	0.20790
10	0.00000	0.00030	0.00690	0.02140	0.15840	0.60520	0.20800
20	0.00000	0.00030	0.00690	0.02142	0.15821	0.60521	0.20796
30	0.00000	0.00300	0.00690	0.02140	0.15780	0.60360	0.20740
40	0.00000	0.00300	0.00690	0.02140	0.15750	0.60390	0.20730
60	0.00000	0.00300	0.00670	0.02150	0.15760	0.60380	0.20730
90	0.00000	0.00280	0.06260	0.02030	0.14860	0.57010	0.19560
120	0.00000	0.00280	0.06270	0.02020	0.14850	0.57020	0.19560

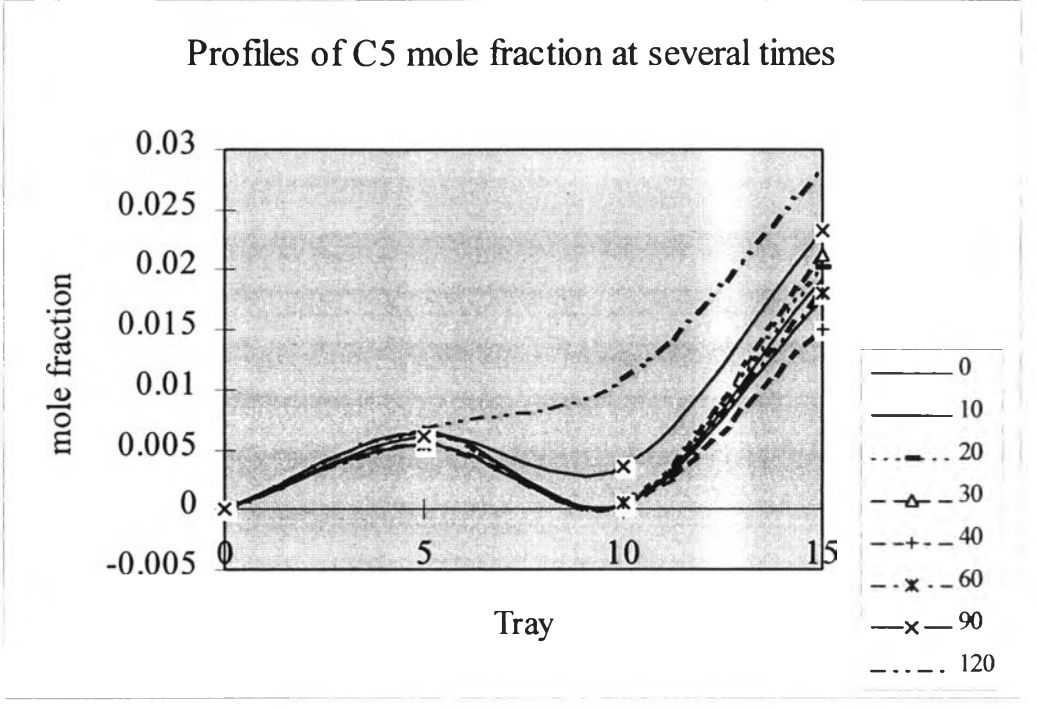

Figure 7.4 Plot of C_5 mole fraction at several times

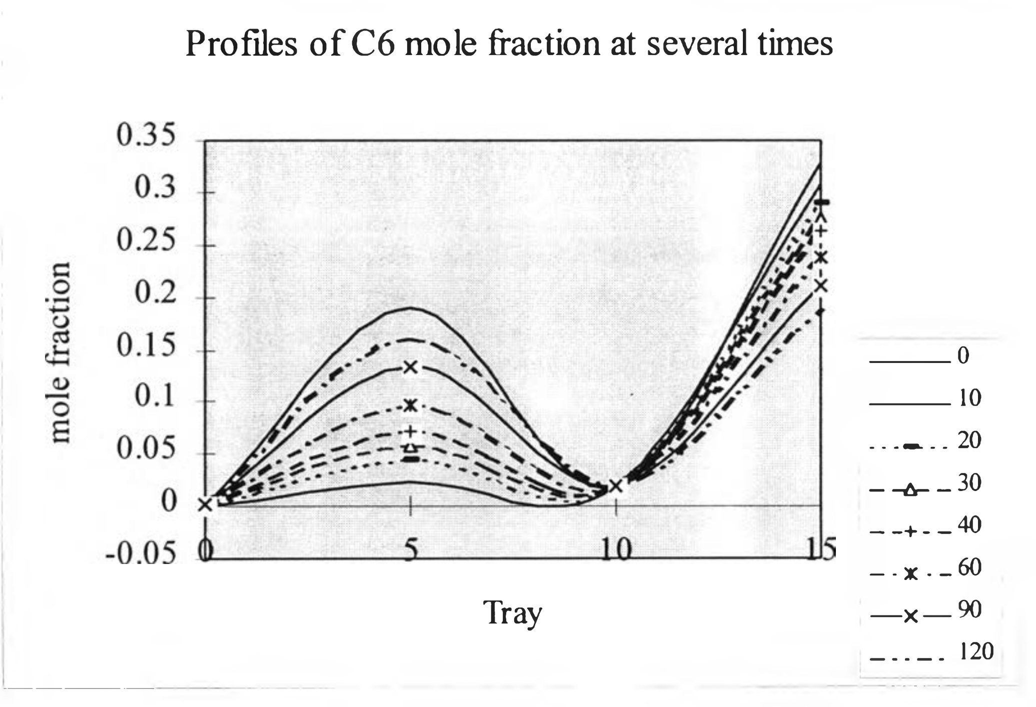

Figure 7.5 Plot of C_6 mole fraction at several times

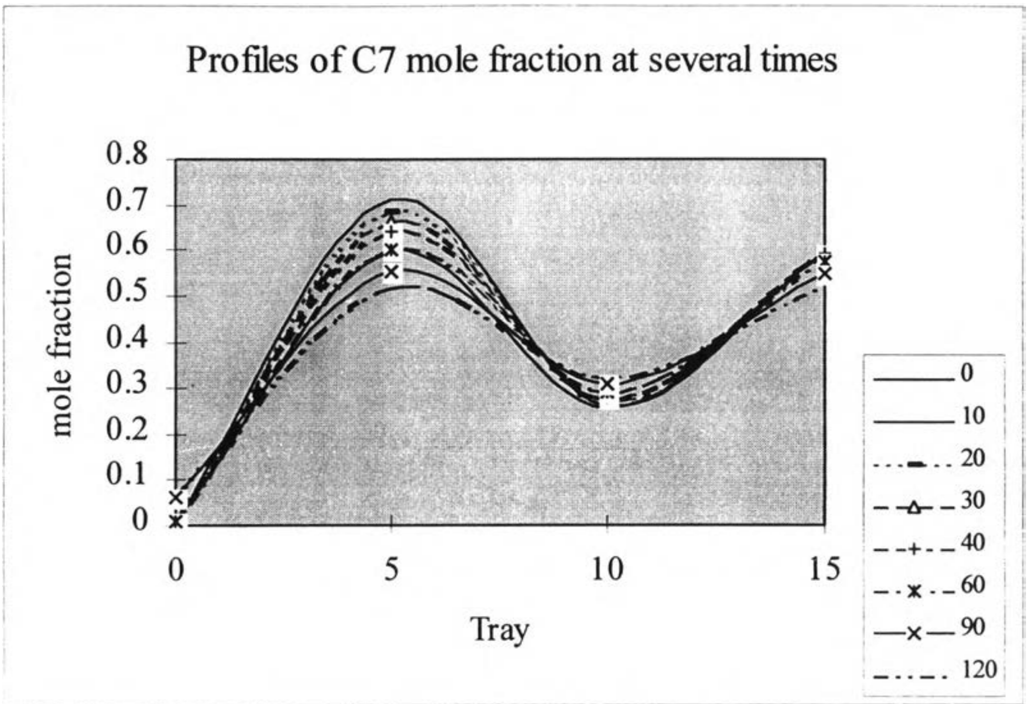

Figure 7.6 Plot of C_7 mole fraction at several times

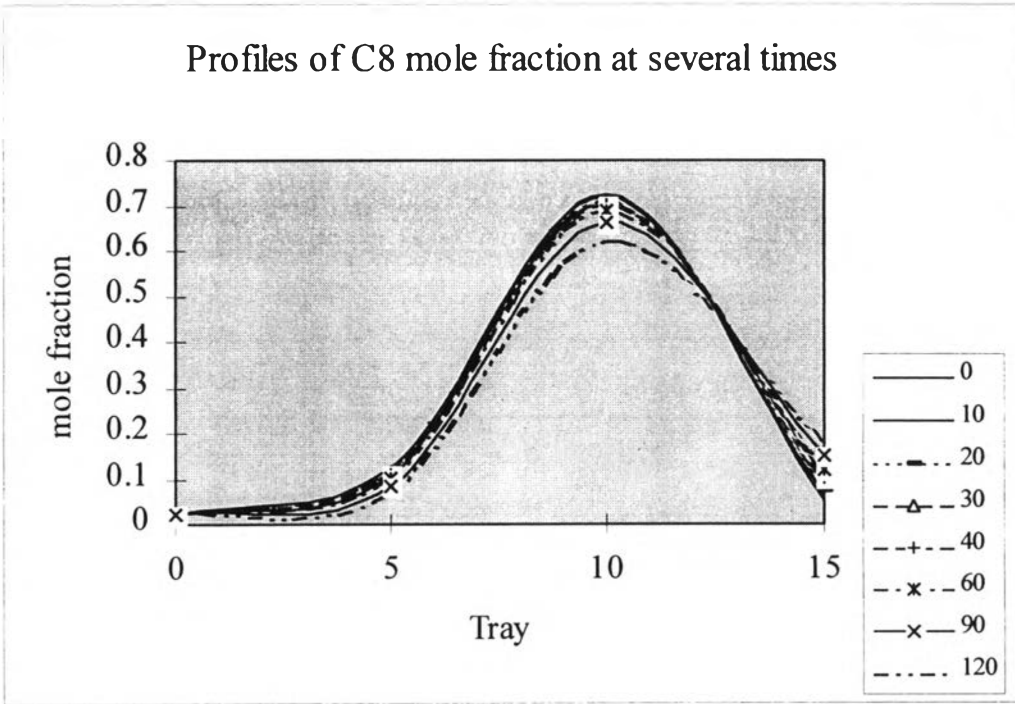

Figure 7.7 Plot of C_8 mole fraction at several times

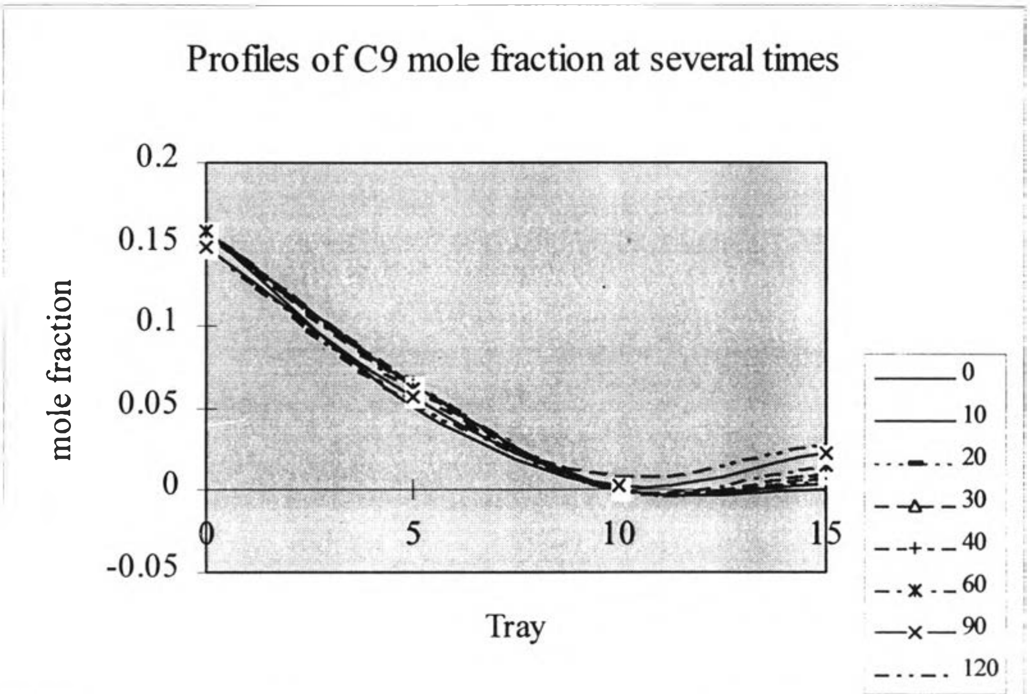

Figure 7.8 Plot of C_9 mole fraction at several times

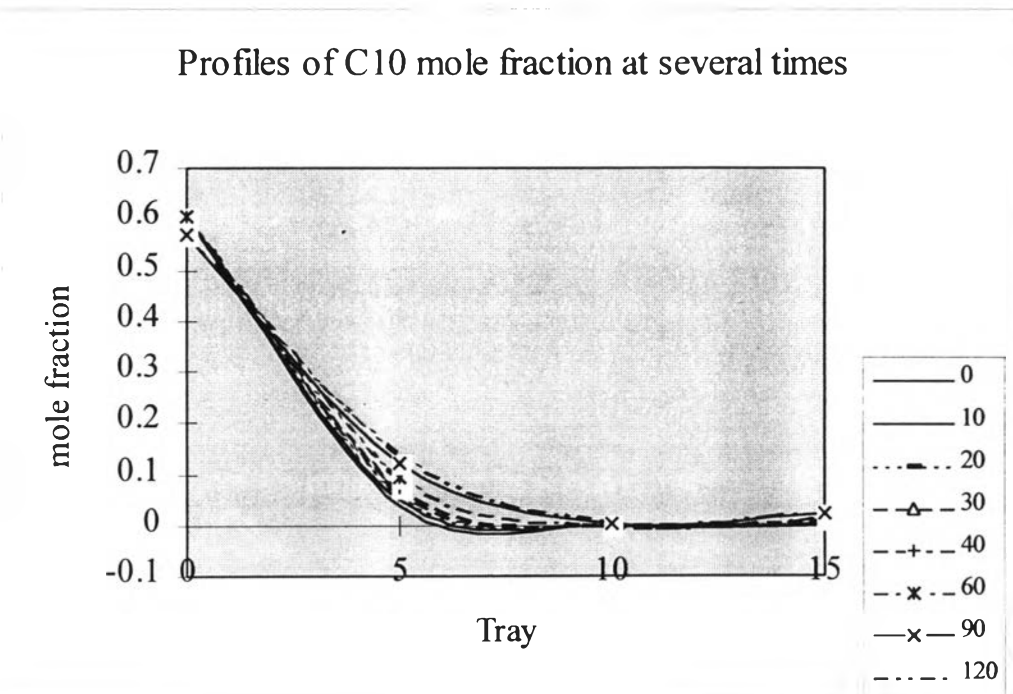

Figure 7.9 Plot of C_{10} mole fraction at several times

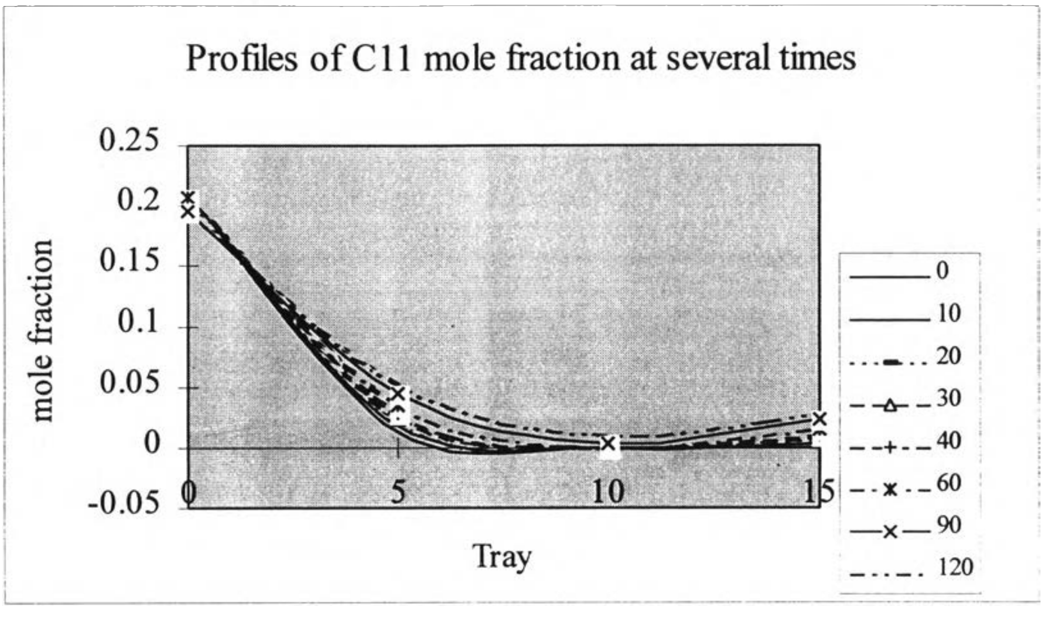

Figure 7.10 Plot of C_{11} mole fraction at several times

Table 7.12 In comparison of distillate (Rubber Solvent) @ 120 min.

Fraction	C_5	C_6	C_7	C_8	C_9	C_{10}	C_{11}
Model	0.0283	0.1881	0.5212	0.1782	0.0283	0.0276	0.0283
GC	0.0181	0.3290	0.5977	0.0534	0.0000	0.0000	0.0000

Table 7.13 Enthalpy values from dynamic simulation

Tray	Time (min.)				
	0	10	30	60	120
0	0	$1.10*10^{15}$	$4.98*10^{15}$	$2.01*10^{16}$	$2.05*10^{17}$
5	$5.40*10^{10}$	$7.72*10^{11}$	$6.76*10^{13}$	$3.68*10^{15}$	$9.46*10^{17}$
10	$3.18*10^{9}$	$2.38*10^{13}$	$8.95*10^{13}$	$2.36*10^{14}$	$2.09*10^{15}$
15	$1.39*10^{11}$	$1.72*10^{14}$	$5.04*10^{14}$	$1.01*10^{15}$	$2.37*10^{15}$